Optimization of Desalination Process

by

Lee Fuhan

Dissertation submitted in partial fulfillment of
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(Chemical Engineering)

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Certification of Approval

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Approved by,

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UNIVERSITI TEKNOLOGI PETRONAS
TRONOH, PERAK
January 2010

Certification of Originality

This is to certify that I am responsible for the work submitted in this project, that the original work is my own except as specified in the references and acknowledgements, and that the original work contained herein have not been undertaken or done by unspecified sources or persons.

LEE FUHAN

Abstract

Desalination is the process to separate the salts and minerals from seawater to produce fresh water. This research project presents an optimization study of a single stage flash chamber based on the operating and design criteria of multistage flash (MSF) desalination process. The optimization study is essential to develop an efficient MSF process especially in the energy consumption of the process. The optimization problem is to optimize the total annualized cost of the single stage flash chamber, which includes the operating and capital costs while meeting the constraints based on the mass and energy balances, requirements and design equations. A non-linear programming (NLP) optimization model has been developed by using GAMS to solve the optimization problem by implementing the objective function and constraints. The optimal operating parameters, capital cost factors, operating and capital costs, as well as the total annualized cost (TAC) obtained from the optimization model is analyzed. Lastly, it has been found that the major contribution to the TAC is the energy cost. Thus, the future optimization study on MSF process should focus on the optimization of energy consumption.

Acknowledgement

First and foremost, I would like to express my sincere gratitude to my supervisor, Dr. Shuhaimi Mahadzir for his valuable guidance throughout this project. Without his advices and supports, the accomplishment of this project would be impossible. Besides that, I would also with to thank the internal examiner, Dr. Yusmiza Yusoff for his helpful and constructive feedbacks and suggestions given during the Final Year Project I presentation. Lastly, I would also like to thank the Final Year Project Committee for their great work in coordinating the final year project.

Table of Content

List of Figures

List of Tables

Chapter 1 : Introduction

1.1 Background

Approximately 70% of the earth's surface area is covered in water, which is equivalent to about 1.4×10^9 m^3 of water. However, 97.5% of this large amount is salt water, or better known as seawater. The remaining 2.5% is the fresh water with 80% of this amount found in frozen icecaps or combined as soil moisture and the fresh water resources are unevenly distributed across the globe. The table below shows the distribution of water sources on Earth.

Table 1.1: Distribution of water sources

Location	**Amount (10^6 km^3)**	**Percentage of World Water**
Ocean	1338.0	96.5
Glaciers and permanent snow	24.1	1.74
Groundwater (brackish or saline)	12.9	0.94
Groundwater (fresh)	10.5	0.76
Ground ice/permafrost	0.30	0.022
Freshwater lakes	0.091	0.007
Freshwater stream channels	0.002	0.0002

(Source: Committee on Advancing Desalination Technology, 2008)

Besides that, with the combination effects of the continuous rapid increase in the world population, changes in life-style, weather and the limited natural resources of fresh water, many parts in the world are facing water shortage problems. Karagiannis I.C. and Soldatos P.G. (2007) claim that 25% of the world population does not have access to satisfactory quality and/or quantity of freshwater and more than 80 countries face severe water problems. By the year 2025, this percentage is expected to increase to more than 60% (El-Dessouky H.T. and Ettouney H.M., 2002, p.5). Ultimately, the abundant seawater becomes one of the best alternate water sources.

Desalination is the process to separate the salts and minerals from seawater to produce fresh water. The desalination processes can be achieved through thermal or membrane separation. The thermal separation processes include the multistage flash desalination (MSF), multiple effect evaporation (MEE), single effect vapor compression (SEE), humidification-dehumidification (HDH) and solar stills. On the other hand, the membrane separation processes consist of reverse osmosis (RO) and electrodialysis (ED). Based on the committee on Advancing Desalination Technology (2008), membrane-based desalination processes appear to have the largest percentage of the total capacity, with 56% in the United States of America and 96% worldwide as shown in the pie charts below.

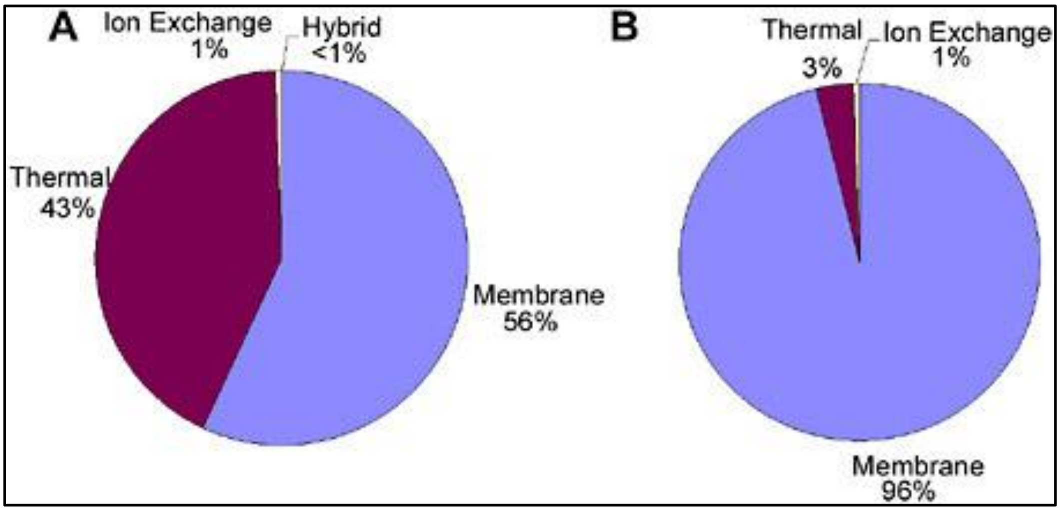

Figure 1.1: Percentage of currently operating desalination plants by technology at (A) United States of America and (B) Worldwide

(Source: Committee on Advancing Desalination Technology, 2008)

Besides that, the costs for different desalination methods are different as well. Global Water Intelligence (2006a) claims that the capital costs of seawater desalination by MED and MSF to be 1.5 to 2.0 times the capital costs of R.O desalination systems, respectively. Thermal desalination systems also consume more energy than RO systems. The table below shows the breakdown of desalination costs for different methods for 100,000m^3 desalination plants:

Table 1.2: Comparative total cost data for desalination processes

	RO	**MSF**	**MED**
Annualized capital costs	0.15	0.29	0.22
Parts/maintenance	0.03	0.01	0.01
Chemicals	0.07	0.05	0.08
Labor	0.10	0.08	0.08
Membrane (life not specified)	0.03	0.00	0.00
Thermal energy	0.00	0.27	0.27
Electrical energy ($0.05/kWh)	0.23	0.19	0.06
Total ($/$m^3$)	0.61	0.89	0.72

(Source: Committee on Advancing Desalination Technology, 2008)

The committee on Advancing Desalination Technology (2008) says that the global desalination water production capacity has been increasing exponentially since 1960 to its current value of 42 million m^3/day, as shown in Figure 1.2:

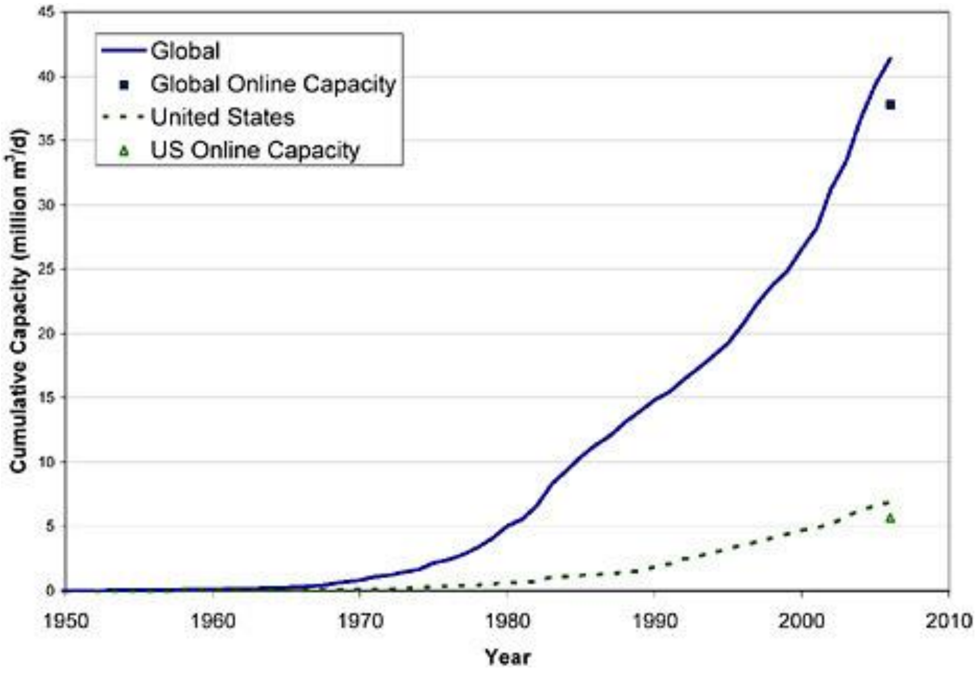

Figure 1.2: Cumulative capacity of installed desalination plants

(Source: Committee on Advancing Desalination Technology, 2008)

1.2 Multistage Flash (MSF) Desalination Process

MSF desalination process produces distilled water from the condensation of the flashed brine vapor through a series of flashing chambers. The seawater enters the preheater/condenser tubes at the last stage of the heat rejection stage and flows through the flashing chambers until the first stage at the heat recovery stage before entering the brine heater. The heat rejection stage is introduced in the MSF process to control the temperature of the brine by removing excess heat added to the system in the brine heater.

The drive of this flashing process is the low pressure (LP) steam with a temperature range of 97-117^{o}C. After the seawater flows through from the last stage to the first stage, the saturating steam heats up the brine at the brine heater. As the steam at the shell side condenses, the brine inside the tubes gains the latent heat of condensation, thus, heating up the brine to the desired top temperature.

The hot brine then enters the first stage of the flashing chambers, where vapor is formed through the flashing of the hot brine. Due to the flashing process, the temperature of the remaining brine solution drops. The temperature reduction across the flashing stages leads to the pressure drop across the stages as well. In other words the highest stage pressure is at the first stage, while the lowest stage pressure is at the last stage. This allows the brine to flow without the need of any pumping unit. In each flashing stage, the flashed vapor from the brine flows through the demister, where any entrained droplets of brine is removed to avoid contamination of the distillate product. As the flashed vapor is at a higher temperature than the seawater inside the preheater/condenser tubes, heat transfer occurs across the tubes. The flashed vapor condenses and forms distillate which is collected at the distillate trays across the flashing stages as the final distillate product. The latent heat of condensation which is released during the condensation of flashed vapor, heats up the seawater stream inside the tubes. The same process takes place in all the flashing stages in both heat recovery and heat rejection stages. Figure 1.3 shows a simplified flow diagram of MSF process.

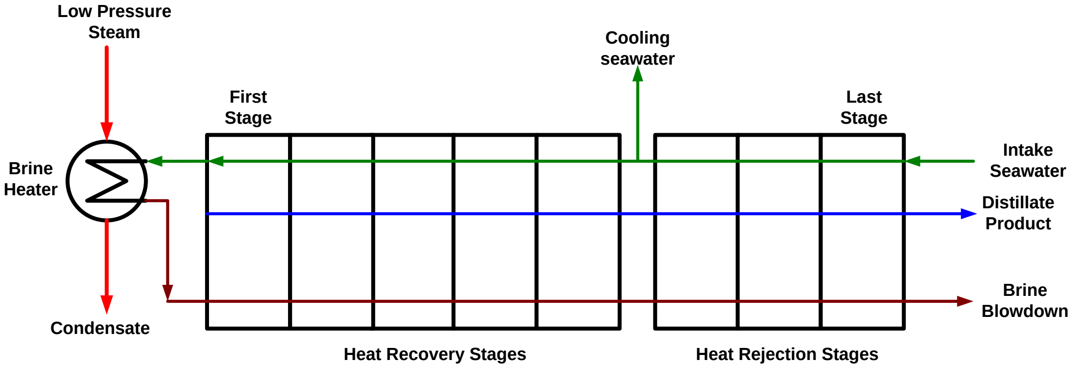

Figure 1.3: Simplified flow diagram of MSF process

1.3 Problem Statement

Optimization of the MSF process is essential to develop a more efficient desalination process, especially in terms of energy consumption as the intensive energy consumption of MSF process. In this research project, the optimization is focused on the operating and capital costs of a single stage of the MSF process. As it involves many variables, it is hoped that by the development of the mathematical programming with GAMS, the optimum synthesis of MSF process can be solved simultaneously.

1.4 Objectives

The main objective of this research project is to develop an automatic synthesis for optimization of the total annualized cost of a single stage of MSF desalination process by applying mathematical programming.

The automatic synthesis optimization model is formulated based on these constraints:

- Amount of distillate and brine
- Concentration of water in brine
- Energy required for the single stage flash chamber
- Dimensions of the single stage flash chamber
- Wall thickness of the single stage flash chamber

1.5 Scope of Study

In order to ensure the feasibility of this research project within the given time frame, the boundaries of the project work is narrowed down.

There are numerous types of desalination process available nowadays. In this research project, the selected desalination process is the multistage flash (MSF) desalination. However, the optimization model proposed in this research project only focus on a single stage of flash chamber. The minimum production capacity of this research project is set at 30000 m^3/day and the MSF desalination process is assumed to be in operation for 360 days annually.

The energy required in this optimization model is measured in kWh, and the cost of power is taken from Malaysia's largest electricity utility company, Tenaga Nasional Berhad (TNB), which is 0.0080606 USD/kWh.

Among the operating parameters which are involved in this optimization model is the amount of energy required, amount of distillate, and amount of brine and concentration of water in brine. The inlet pressure of the MSF desalination process is assumed to be 2 bar and outlet pressure at 1 bar. The amount of feed seawater is 50000 m^3/day, and the typical composition of seawater with salinity of 36000 ppm

The optimization of the capital cost in this research project ventures into the design of the single stage flash chamber. As the capital cost is related to the amount of 308 stainless steel required for the fabrication of the flash chamber, the design cost factors which are involved are the dimensions, i.e., length and diameter of the flash chamber, and the wall thickness of the flash chamber. Based on Treybal (1981), for economical reasons, the ratio of length over diameter (L/D) should be in the range of 3-5. As for this research project, the value for L/D selected is 3. On the other hand, the minimum wall thickness is also calculated. The cost for 308 stainless steel based on MEPS (International) Ltd. is 3565 USD/ton.

Chapter 2 : Literature Reviews

2.1 Operating and Capital Cost Factors

Similar to any other chemical processes, operating variables and cost factors play vital roles in determining the performance of certain processes. It is necessary to review and understand all the operating variables and cost factors and also their effects in MSF process as they are considered in the optimization of MSF process as well.

In the article by Rosso, Beltramini, Mazzotti and Morbidelli (1996), the effect of operating variables on the performance of MSF plant is analyzed. The operating variables studied are number of flashing stages, steam temperature and seawater temperature.

2.1.1 Number of Flashing Stages

The increasing number of stages yields an improvement in process performance. However, it is also mentioned that the improvement is due to the simultaneous increase of the product distillate flow rate and decrease of the steam flow rate. According to Rosso, Beltramini, Mazzotti and Morbidelli (1996), in designing a MSF plant, the number of stages selected is a compromise between the fixed costs and variable costs, where the fixed costs increase while the variable costs decrease with increasing number of stages. The results of their work are shown in Table 2.1:

Table 2.1: Effect of number of stages on the MSF plant performance

Stage no.	Number of stages in heat recovery section	Number of stages in heat rejection section	GOR	Distillate (kg/h)	Steam (kg/h)
11	9	2	5.1	8.5×10^5	1.75×10^5
16	13	3	6.9	9.3×10^5	1.3×10^5
22	18	4	9.0	9.9×10^5	1.1×10^5
27	22	5	10.6	10.2×10^5	0.96×10^5

In the research of Tanvir and Mujtaba (2008), it can also be observed that as the number of flashing stages increases, the energy required from steam decreases, thus leads to less amount of steam required and ultimately the decrease in TAC, as shown in Table 2.2 below:

Table 2.2: Summary of optimization results for different fixed water demand

T_{sea}, °C	*NR*	*R*, kg/h	C_w, kg/h	*TBT*, °C	W_R, kg/h	Q_{steam}	T_{steam}, °C	W_{steam}, kg/h	*TAC*, $/y
Case 1: Water demand, D_{end} = 700,000 kg/h									
20 (Winter)	15	2.40×10^4	4.54×10^6	90	6.78×10^6	3.99×10^7	93.12	62765	3.27×10^7
25	16	2.40×10^4	4.05×10^6	90	7.27×10^6	3.85×10^7	93.09	60498	3.34×10^7
30	18	2.40×10^4	5.47×10^6	90	7.85×10^6	3.64×10^7	93.00	57186	3.43×10^7
35	19	2.40×10^4	2.82×10^6	90	8.56×10^6	3.56×10^7	93.02	55874	3.55×10^7
40 (Summer)	21	2.40×10^4	1.90×10^6	90	9.42×10^6	3.44×10^7	93.01	54065	3.77×10^7
Case 2: Water demand, D_{end} = 800,000 kg/h									
20 (Winter)	21	4.70×10^5	4.01×10^6	90	7.76×10^6	3.65×10^7	93.00	57415	3.52×10^7
25	22	5.18×10^4	3.03×10^6	90	8.33×10^6	3.58×10^7	93.01	56261	3.61×10^7
30	24	2.19×10^6	4.49×10^6	90	9.00×10^6	3.48×10^7	93.00	54698	3.75×10^7
35	26	1.94×10^6	3.44×10^6	90	9.80×10^6	3.39×10^7	93.00	53350	3.91×10^7
40 (Summer)	28	5.62×10^4	5.38×10^6	90	1.08×10^7	3.35×10^7	93.04	52630	4.12×10^7
Case 3: Water demand, D_{end} = 900,000 kg/h									
20 (Winter)	25	3.42×10^4	2.57×10^6	90	8.76×10^6	3.73×10^7	93.19	58619	3.82×10^7
25	28	2.40×10^4	1.93×10^6	90	9.39×10^6	3.49×10^7	93.04	54799	3.93×10^7
30	28	2.40×10^4	1.14×10^6	90	1.02×10^7	3.60×10^7	93.22	56602	4.10×10^7
35	28	2.40×10^4	1.68×10^5	90	1.12×10^7	3.77×10^7	93.44	59163	4.30×10^7
40 (Summer)	28	4.41×10^4	3.38×10^6	90	1.23×10^7	4.08×10^7	93.82	64086	4.59×10^7

2.1.2 Steam Temperature

The steam temperature in a MSF plant is one of the parameters which could affect the performance of the plant. Increasing steam temperature will increase the performance; however, this also implies the requirement for a higher pressure steam at a higher cost. Rosso, Beltramini, Mazzotti and Morbidelli (1996), measures the effect of steam temperature on the top brine temperature (TBT) and bottom brine temperature (BBT) as TBT and BBT directly affects the production of the MSF plant. Higher TBT and BBT will leads to greater increase in the distillate product flow rate.

In other words, the steam temperature directly affects the energy cost, which is the operating cost factor considered in this research project. The energy cost greatly affects the operating cost of a MSF process. As mentioned by Mesa, Gomez and Azpitarte (1996), energy costs invariably represent 50 to 75% of the real operating costs, regardless of the technology used and the design of the seawater desalinator. Therefore, in order to optimize the MSF process, it is inevitably to include the energy cost. Figure 2.1, 2.2 and 2.3 below are from the research by Rosso, Beltramini, Mazzotti and Morbidelli (1996):

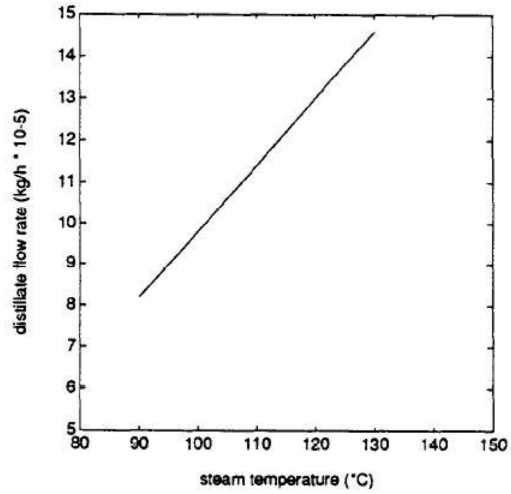

Figure 2.1: Effect of steam temperature on the distillate flow rate

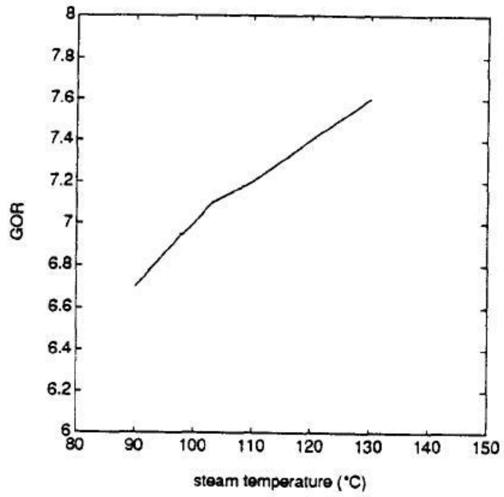

Figure 2.2: Effect of steam temperature on the performance parameter

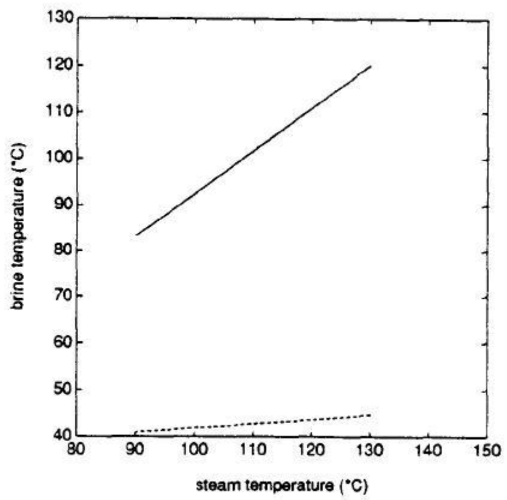

Figure 2.3: Effect of steam temperature on TBT (solid line) and BBT (dotted line)

2.1.3 Seawater Temperature

In the research work by Rosso, Beltramini, Mazzotti and Morbidelli (1996), the seawater temperature's effect on the MSF plant is discussed. Even the seawater temperature is a variable affected by external disturbances and subjected to seasonal and daily variations, it does has a significant impact on the MSF plant as the BBT is very sensitive to the seawater temperature variations. In other words, as the seawater temperature increases, the distillate product flow rate decreases. Figure 2.4, 2.5 and 2.6 below are from their work:

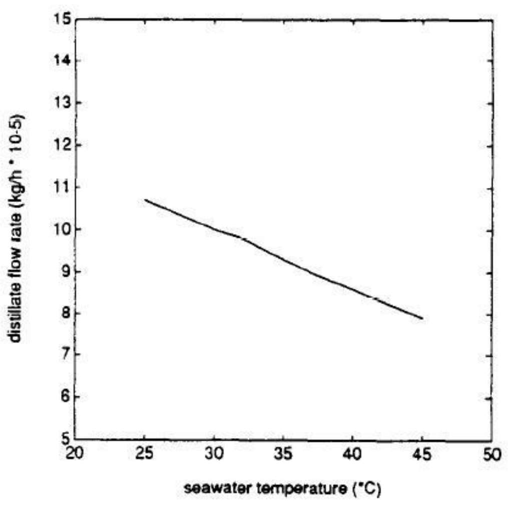

Figure 2.4: Effect of seawater temperature on distillate product flow

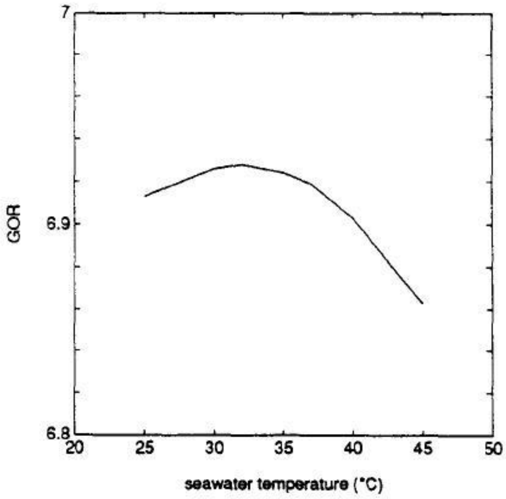

Figure 2.5: Effect of seawater temperature on process performance

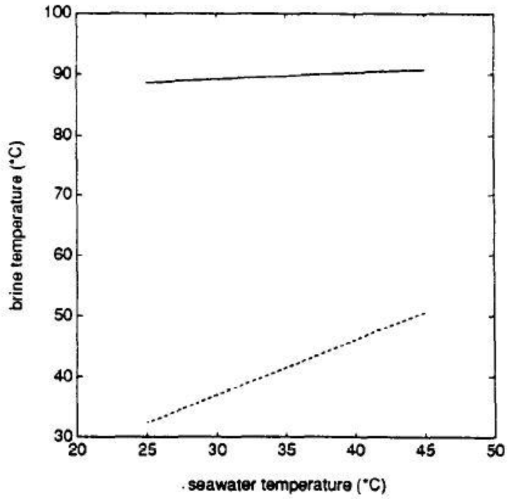

Figure 2.6: Effect of seawater temperature on TBT (solid line) and BBT (dashed line)

2.2 Desalination Cost

The objective of this research project is to optimize the total annualized cost (TAC) of a MSF process by optimizing the operating cost and capital cost. In the study of economics of thermal and membrane processes by Andrianne and Alardin (2002), in the field of optimization in desalination industry, the capital expenses (CAPEX) key parameters to be considered as suggested are site selection, desalination process equipment, electrical network, civil works, water intake and outfall, electromechanical equipment, fuel supply equipment, water distribution network, transportation, erection and commissioning, engineering and supervision, and financial charges. On the other hand, the operating expenses (OPEX) key parameters which to be considered are fuel consumption, electricity consumption, electricity export, chemicals, personnel costs, maintenance and overhaul. The capital and operating cost factors which are considered in this research project will be further reviewed.

The water desalination cost is also studied from the literatures as it measures the objective of this research project in dollars and cents. Karagiannis and Soldatos (2008) claims that the cost for MSF process varies between \$0.52/m^3 and \$1.75/m^3 and refers to systems with daily production from 23,000m^3 to 528,000m^3. On the other hand, another article by Wade (2001) states that based on the fuel cost of \$1.5 per gigajoule and for a MSF plant with the capacity of 31,822m^3/day, the estimated water cost for MSF process is \$1.04/m^3.

However, as the objective function in this research project is focused on the optimization of the total annualized cost (TAC), the best reference for the TAC is the research work done by Tanvir and Mujtaba (2008). In their research, various MSF units with different fixed water demand (700000 kg/h, 800000 kg/h and 900000 kg/h) are optimized at different seawater temperatures (20, 25, 30, 35, and 40) and number of recovery stages (15, 16, 18, 19 and 21).

For this research project, it is necessary to compare the results of this proposed optimization model with the values from literatures to determine the feasibility of the model. Thus, the optimization results from Tanvir and Mujtaba (2008) are being analyzed. The average values of TAC for each fixed water demand is taken and plotted against the production capacity, which is the fixed water demand.

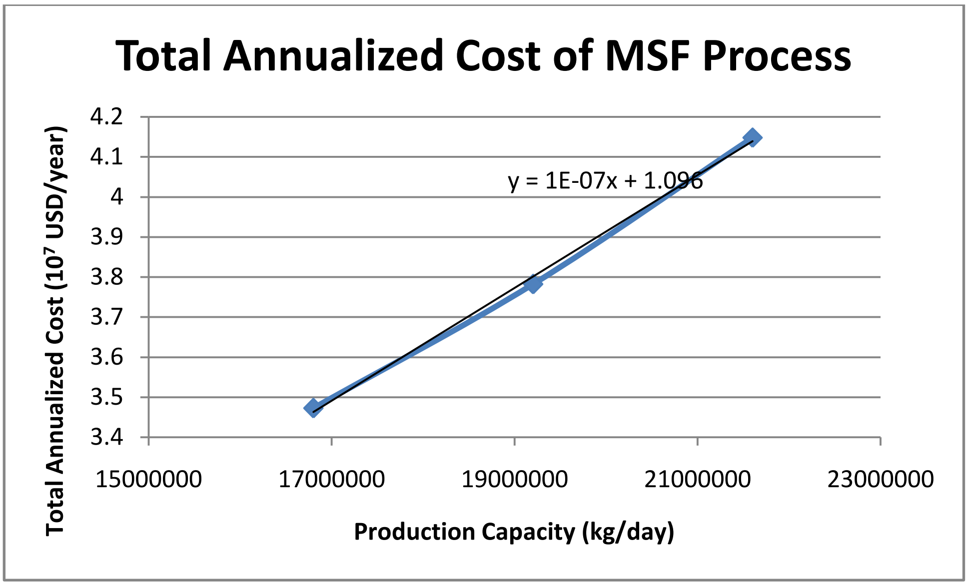

Figure 2.7: TAC of MSF process based on research by Tanvir and Mujtaba (2008)

By using the linear equation obtained from the graph above, the TAC for the production capacity of this research project can be estimated with this equation:

$$TAC = 1 \times 10^{-7}(Production\ capacity) + 1.0967$$

2.3 Optimization of Desalination Process

As one of the objectives of literature reviews, the previous similar researches conducted related to the optimization of desalination process are also reviewed.

Mussati S., Aguirre P. and Scenna N.J. (2001) presented a rigorous model for MSF system in a non-linear programming (NLP) model by using GAMS. In this model, only the costs of the heat transfer area and energy consumption are considered. Furthermore, a few assumptions made in their study are:

- A mean value of heat capacity coefficients is used
- The effects of brine concentration, temperature and pressure are neglected
- The effects of chamber geometry, temperature, pressure and fluid parameters are also neglected

\The objective function implemented in this model is to:

$$Minimize\ CRF\ C_A \left(\sum_{1}^{NS} A^{j}_{Recov} \right) + C_{Q^{Des}}\, Q^{Des}$$

where,

CRF	Capital recovering factor
C_A	Area transfer unit cost ($\$/m^2$)
NS	Number of stage
A^j_{Recov}	Heat recovery transfer area (m^2)
C_{QDes}	Heat consumption unit cost, \$/kcal
Q^{Des}	Heat consumption, kcal/h

Tanvir and Mujtaba (2008) have also conducted a research on optimization of design and operation of MSF desalination process using MINLP technique in gPROMS. In their work, the total annualized cost of the desalination (investment and operation cost) required are minimized for three different fixed water demand and for changing seawater temperature. Among the design parameter considered in this research is number of stages, and operating parameters such as top brine temperature (TBT), steam temperature (reflects utility/energy cost), recycled brine flowrate and rejected seawater flowrate (reflects pumping cost). In other words, the total annualized cost (TAC) in their research consists of annualized capital cost; annualized steam cost and annualized pumping cost. Seawater temperature is a very important parameters in Tanvir and Mujtaba (2008)'s research. However, in this research project, seawater temperature is not taken in account as one of the operating variables to be optimized as seawater temperature is affected by external factors.

Abduljawad and Ezzeghni (2010)'s research on optimization of Tajoura MSF desalination plant was conducted to maximize the gained output ratio (GOR) at different plant capacities by varying the top brine temperature. Besides that, the feasibility of increasing the plant capacity from the current capacity of 1200 to 1300 m3/day without the mechanical design alterations is conducted as well. In their work, based on the design data of the MSF plant, the optimal operating conditions are determined to maximize the GOR. From their research, the plant productivity can be augmented by 10% at 28°C and 14°C.

Chapter 3 : Methodology

3.1 Calculation of Enthalpies and Concentration of Water in Distillate

The enthalpies of feed seawater, distillate and brine streams are the function of pressure. With the inlet pressure of 2 bar and outlet pressure of 1bar, the enthalpies for each stream are taken from the saturated steam table, and the values are shown in Table 3.1 below.

Table 3.1: Enthalpies for each stream

Stream	Pressure	Enthalpy
Feed seawater, h_f	2 bar	504.7 kJ/kg
Distillate (vapor), h_d	1 bar	2675.4 kJ/kg
Brine, h_b	1 bar	417.5 kJ/kg

The concentration of water in distillate is also calculated based on the operating pressure in the flash chamber. By using the Antoine Equation,

$$log_{10}p^* = A - \frac{B}{T + C}$$

where,

p^* = Vapor pressure in mmHg

A, B and C = Antoine equation constants

T = Temperature in °C

In this research project, the Antoine equation constants A, B and C for water are 7.96681, 1668.210 and 228 respectively while the temperature is the saturation temperature of steam at 1 bar.

The concentration of water in distillate which is in vapor form, can then be calculated by using Raoult's Law,

$$p^* = y_i P$$

where,

y_i	=	Concentration of component i
P	=	Total pressure in mmHg

3.2 Operating Parameters

The overall mass balance for the system which includes the feed seawater stream (f), distillate stream (d) and brine stream (b) is as below:

$$f = d + b$$

The water mass balance is another form of mass balance, but specifically on the concentration of water in the feed seawater (x_f), distillate (x_d) and brine (x_b) in mass fraction:

$$\left(x_f \times f\right) = (x_d \times d) + (x_b \times b)$$

One of the constraints is the requirement for a minimum production capacity (d_{min}) of the MSF desalination process is 30000 m^3/day, which is 3 x 10^7 kg/day,

$$d \geq d_{min}$$

The second constraint in this optimization model specifies the amount of water which must be successfully flashed into the distillate vapor stream. In this constraint, it is assumed that the maximum of 80% of the water in the feed seawater entering the system, will ended up as the distillate product. The constraint is shown below:

$$(x_d \times d) \leq 0.8 \times (x_f \times f)$$

The energy balance of the system is also one of the constraints in this optimization model as the energy required (q) by the system is one of the variables. The energy balance for the system is as follow:

$$\left(h_f \times f\right) + q = (h_d \times d) + (h_b \times b)$$

Based on the energy required obtained from the energy balance, the cost of energy, i.e., operating cost can be calculated, with the assumption that the MSF plant is in operation for 360 days or 8640 hours annually.

$$Operating\ cost = q_r \times C_p$$

where,

C_p = Cost of power

= 0.0080606 USD/kWh

q_r = Energy required, kWh

= $q \times r_n$

r_n = Operating hours per year, h

3.3 Design of Single Stage Flash Chamber

There are two parts involved in the design of the single stage flash chamber. The first part is the calculation to obtain the dimensions, i.e., diameter and length of the flash chamber, the design equations are proposed by Ludwig (1997).

First, the maximum allowable vapor velocity is calculated by using the equation below:

$$V = k\sqrt{\frac{(dL - dV}{dV}}$$

where,

V = Maximum allowable vapor velocity, m/s

dL = Liquid density, kg/m^3

dV = Vapor density, kg/m^3

k = 0.107 m/s (when the drum includes a de-entraining mesh pad)

The values for dL, dV and k are as shown in Table 3.2 below:

Table 3.2: Values for maximum allowable vapor velocity calculation

Constant	Value
dL	958.204 kg/m^3
dV	0.5875 kg/m^3
k	0.107 m/s

The diameter of the flash chamber can then be calculated by using the equation below,

$$D = \left[\frac{4}{\pi}(CSA)\right]^{\frac{1}{2}}$$

where,

D = Diameter, m

CSA = Cross sectional area, m^2

= $\frac{W}{V}$

W = Vapor flow rate, m^3/s (amount of distillate vapor)

V = Vapor velocity, m/s

Then, the length of the flash chamber can be calculated by using the ratio of length over diameter (L/D), which in this research project, L/D is 3,

$$\frac{L}{D} = 3$$

The second part of the design of the flash chamber is the calculation of the minimum wall thickness of the flash chamber. The design equation by Peters, Timmerhaus and West (2003) is used in this section,

$$t = \frac{Pr_i}{SE_j - 0.6P} + C_c$$

where,

t = minimum wall thickness, m

P = maximum allowable internal pressure, kPa

r_i = inside radius of shell, before corrosion allowance is added, m

S = maximum allowable working stress, kPa

E_j = efficiency of joints expressed as a fraction

C_c = allowance for corrosion, m

The values for the constants are shown in Table 3.3 below:

Table 3.3: Values for minimum wall thickness calculation

Constant	Value
P	1000 kPa
S	72400 kPa
E_j	1
C_c	0.003 m

Referring to by Peters, Timmerhaus and West (2003), the value for maximum allowable working stress, S is chosen based on the material of the flash chamber, where in this research project, is 304 stainless steel. While the efficiency as joints expressed as a fraction, E_j is chosen by assuming the flash chamber is using double-welded butt joints which is fully radiographed. The allowance for corrosion is 0.003 m as it is assumed to be 0.003 m or 3 mm for 10 years life.

By taking the diameter and length obtained, the total surface area of the flash chamber wall can be calculated with the assumption that the flash chamber is in cylindrical shape,

$$SA = 2\pi(\frac{D}{2})^2 + 2\pi\left(\frac{D}{2}\right)L$$

where,

SA = Total surface area with top and bottom, m^2

By using the total surface area and wall thickness, the volume of the flash chamber wall can be calculated,

$$V = t \times SA$$

where,

V = Volume of flash chamber wall, m^3

Eventually, the weight of the stainless steel required can be calculated,

$$W = \frac{V}{\rho_{ss}}$$

where,

W = Weight of stainless steel, ton

ρ_{ss} = Density of stainless steel

= 7.83 ton/m^3

Based on the weight of the stainless steel required, the cost of the stainless steel, i.e., capital cost can be calculated,

$$Capital\ cost = W \times C_{ss}$$

where,

C_{ss} = Cost of stainless steel

= 3565 USD/ton

3.4 Total Annualized Cost

In this research project, the objective function in this optimization model is to minimize the total annualized cost which includes the operating and capital costs:

$$TAC = operating\ cost + capital\ cost$$

The objective function stated above is subjected to related constraints and design equations of the single stage flash chamber discussed above.

3.5 Development of Optimization Model

The NLP optimization model of single stage flash chamber is developed by using GAMS programming language.

First of all, the basis data required for the constraints and equations of the optimization model such as the enthalpy of each stream and concentration of water in the distillate stream are calculated. The basis data and values for constants in the design equations are being declared as scalar in the optimization model. Then, the objective function for the optimization model which is the total annualized cost (TAC) is formulated.

The first section of equations in the optimization model consists of constraints derived from the overall mass balance, water mass balance, requirements of the MSF desalination process and energy balance. With this section of equations, the equation for calculating the optimized operating cost is formulated.

The second section of equations consists of design equations to calculate the dimensions, i.e., length and diameter of flash chamber and wall thickness of the flash chamber. From the dimensions and wall thickness obtained, the equations to calculate the amount of stainless steel required, as well as the capital cost are formulated.

Lastly, the optimization model is executed to obtain the optimal value for TAC. The optimization model is considered to be completed when an optimal value for TAC can be obtained from the execution. In the other hand, if no optimal value is obtained due to infeasibilities and errors in the optimization model, the flow will go back and repeat from the formulation of objective function for modification.

The flow of development of the optimization model is shown in Figure 3.1 at next page:

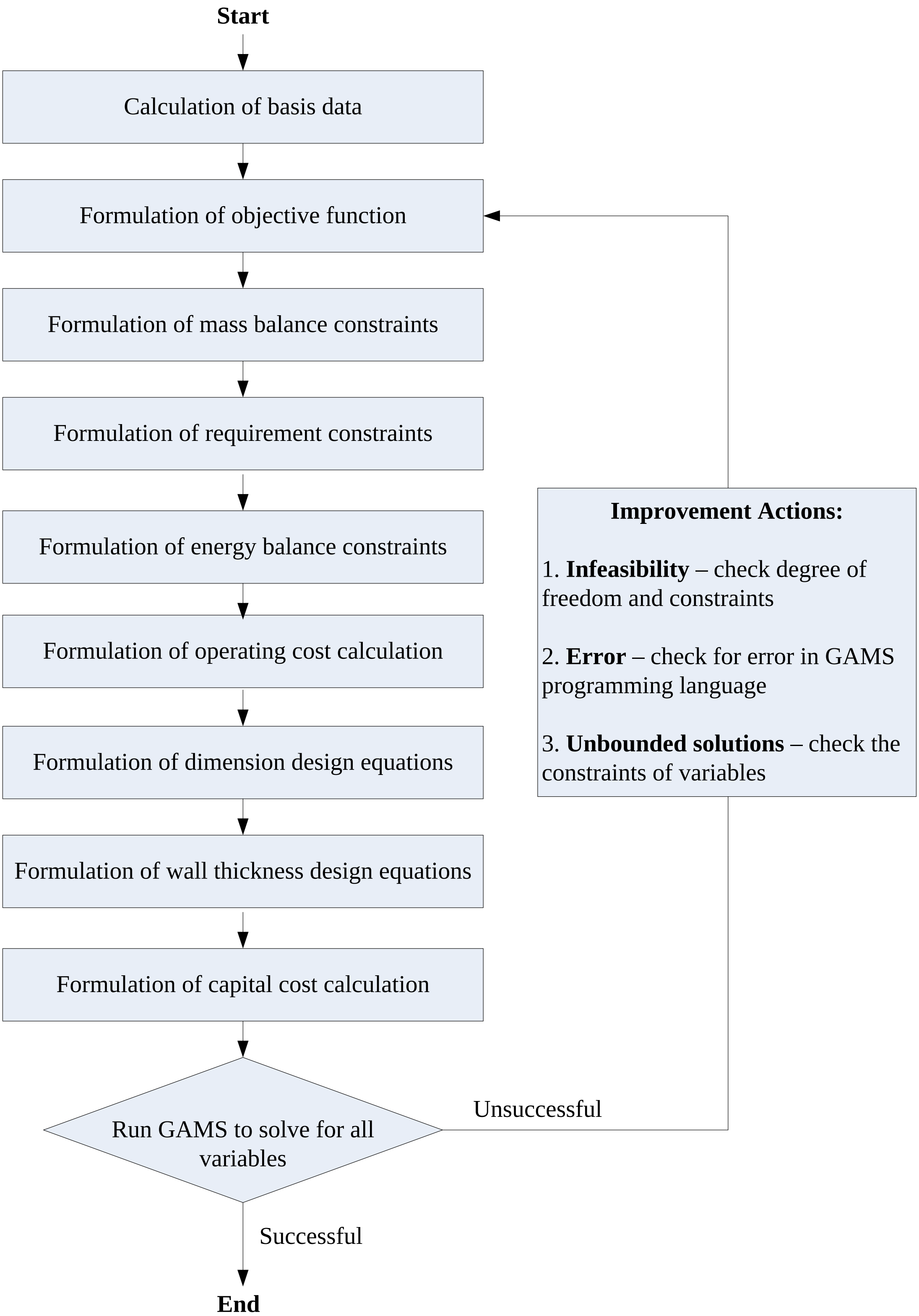

Figure 3.1: Flow of development of optimization model

The formulation of the NLP optimization model for this research project can be shown below:

Minimize $$TAC = \left(q_r \times C_p\right) + (W \times C_{ss})$$

s.t $$f - d - b = 0$$

$$\left(x_f \times f\right) - (x_d \times d) - (x_b \times b) = 0$$

$$\left(h_f \times f\right) + q - (h_d \times d) - (h_b \times b) = 0$$

$$q_r = q \times r_n$$

$$V - k\sqrt{\frac{(dL - dV)}{dV}} = 0$$

$$D - \left[\frac{4}{\pi} \times \left(\frac{W}{V}\right)\right]^{\frac{1}{2}} = 0$$

$$L - (3 \times D) = 0$$

$$SA - 2\pi\left(\frac{D}{2}\right)^2 - 2\pi\left(\frac{D}{2}\right)L = 0$$

$$t - \frac{Pr_i}{SE_j - 0.6P} - C_c = 0$$

$$V - (t \times SA) = 0$$

$$W - \left(\frac{V}{\rho_{ss}}\right) = 0$$

$$d - d_{min} \geq 0$$

$$(x_d \times d) - 0.8(x_f \times f) \leq 0$$

Chapter 4 : Results and Discussion

The results obtained from the optimization model will be discussed in three sections this chapter.

4.1 Operating Cost

The first section discusses about the results which are related to the operating parameters and operating cost. The results obtained are shown in Table 4.1 below:

Table 4.1: Results for operating parameters and cost

Variable	Value
Amount of distillate	3.00×10^{7} kg/day
Amount of brine	2.00×10^{7} kg/day
Concentration of water in brine	0.912
Amount of energy required	6.3377×10^{10} kJ/day
Operating cost	5.108566×10^{7} USD/year

The operating parameters and cost are bounded and affected directly by the mass and energy balance, as well as the requirements of the MSF desalination process.

The amount of distillate obtained from the optimization model is optimized and feasible as it met the requirement constraints, which the distillate product must be greater or equal to the minimum production capacity of 30000 m^3/day or 3.00×10^7 kg/day and maximum of 80% of the water from the feed seawater must ended up as the distillate product.

The amount of brine, which is 2.00×10^7 kg/day or 20000 m^3/day generated by the optimization model satisfies the overall mass balance of the system. With 50000 m^3/day of feed seawater and 30000 m^3/day of distillate, the remaining seawater leaves the system as the brine stream. Besides that, the concentration of water in brine of 0.912 in mass fraction obtained from the optimization model satisfies the water mass balance. This value ensures that the brine stream is still in liquid form while leaving the system and as seawater with a slightly higher concentration of salt than the feed seawater. This constraint is to ensure that the brine stream does not contain too high concentration of salt.

The energy balance is also one of the constraints in the optimization model. By using the optimized and feasible values of amount of distillate and brine, the energy required by the system is calculated by the optimization model. From the results, the energy required is 6.3377×10^{10} kJ/day or 6.3377×10^{9} kWh/year.

As other auxiliary costs, such as pumping cost are not included in this optimization model, thus the operating cost of the single stage flash chamber depends solely on the cost of energy required. The results in the optimization model shows that the operating cost is 5.108566×10^{7} USD/year.

4.2 Capital Cost

The capital cost of the single stage flash chamber is based on the design of the flash chamber. Table 4.2 below shows the results for the capital cost factors and the capital cost:

Table 4.2: Results for capital cost factors and cost

Variable	**Value**
Vapor flow rate	0.347 m^3/s
Maximum vapor velocity	4.32 m/s
Cross sectional area	0.08 m^2
Diameter of flash chamber	0.32 m
Length of flash chamber	0.96 m
Total surface area	1.125 m^2
Wall thickness of flash chamber	0.005 m
Amount of stainless steel required	0.046 ton
Capital cost	164.209 USD

The vapor flow rate, maximum vapor velocity and cross sectional area are variables which is affected by the amount of distillate. The values for these variables are calculated by the optimization model in order to obtain the diameter and length of the single stage flash chamber. Thus, in other words, the diameter and length of the flash chamber depends on the amount of distillate.

The wall thickness of flash chamber, together with the total surface area of the flash chamber give the amount of stainless steel required for the fabrication of the flash chamber. It is worth to take note that in this optimization model, the flash chamber is assumed to be in cylindrical shape. Ultimately, the capital cost is obtained from the optimization model, which is 164.209 USD.

4.3 Total Annualized Cost

The TAC in this research project is the summation of operating cost and capital cost of the single stage flash chamber of MSF desalination process. The TAC generated by the optimization model is shown in Table 4.3 below:

Table 4.3: Results for total annualized cost

Variable	**Value**
Total annualized cost	5.108583×10^{7} USD/year
Water cost	0.005 USD/m^3

It is worth taken note that the major contribution to the TAC is the operating cost, or to be more specific, the energy cost. In this research project, the operating cost contributes almost entirely to the TAC, with only a small portion of contribution by the capital cost.

By using the linear equation for TAC estimation based on the research work by Tanvir and Mujtaba (2008), the estimated TAC for a MSF process with the production capacity of 30000 m^3/day is 4.0967×10^{7} USD/year.

In comparison with the TAC obtained from the optimization model, it is obvious that the TAC estimated is lower than the TAC obtained. In percentage difference,

$$Percentage\ difference = \left| \frac{Actual - Estimation}{Estimation} \right| \times 100\%$$

$$Percentage\ difference = \left| \frac{5.108583 \times 10^{7} - 4.0967 \times 10^{7}}{4.0967 \times 10^{7}} \right| \times 100\%$$

$$Percentage\ difference = 24.7\%$$

In other words, the actual TAC obtained from the optimization model is 24.7% higher than the estimated TAC. The comparison between the actual TAC and estimated TAC is shown in Figure 4.1.

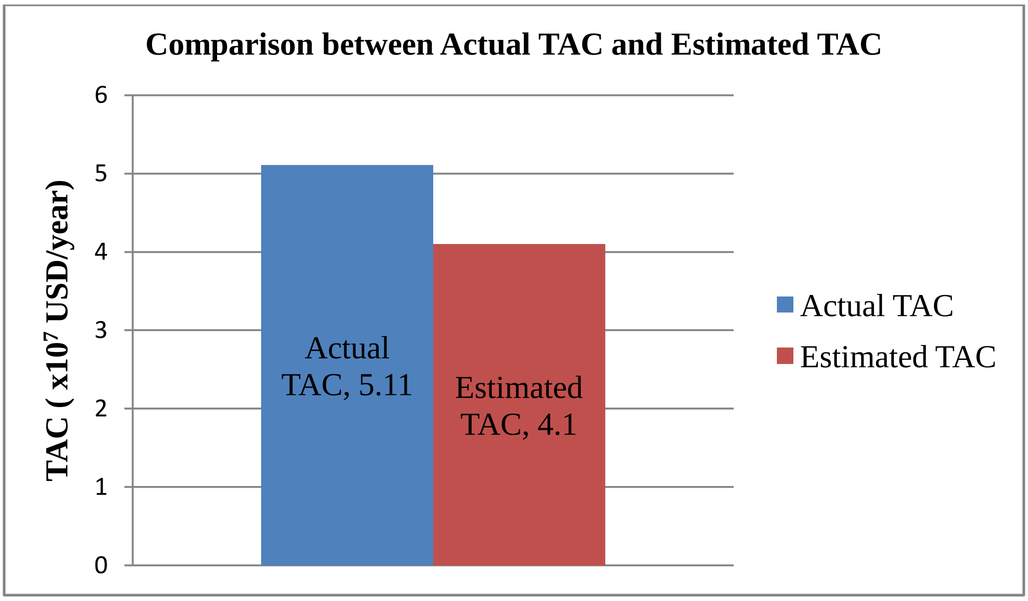

Figure 4.1: Comparison between actual TAC and estimated TAC

There are several factors which may lead to the higher TAC generated by the optimization model proposed in this research project.

The first factor is that the number of flashing stages is not being considered as a variable in this research project. As mentioned by Rosso, Beltramini, Mazzotti and Morbidelli (1996), in designing a MSF plant, the number of stages selected is a compromise between the fixed costs and variable costs, where the fixed costs increase while the variable costs decrease with increasing number of stages. In addition, as the energy cost contributes mainly to the TAC and the number of flashing stages definitely affects the energy cost, it is justified to mention that the TAC can be further optimized by taking number of flashing stages as one of the variables.

Besides that, the error in the TAC estimation by using the linear equation from research work by Tanvir and Mujtaba (2008) may also causes the actual TAC to be higher than the estimated TAC. Linear equation which only involves the power of 1, gives lower accuracy in estimation in comparison with exponential or polynomial equations. Due to the lack of data from the literature reviews, the cost estimation in this research project can only be done by using the linear equation.

Lastly, the method of calculating the energy cost being used in this research project defers with other researches also may lead to the higher actual TAC by the optimization model. In many researches, such as Tanvir and Mujtaba (2008), the energy cost is calculated by using the steam cost; while in this research project, the energy cost is calculated by using the electricity cost. As the energy cost contributes the most to the operating cost, thus, with a different calculation method, it may cause significant difference in the operating cost, and ultimately the capital cost.

Chapter 5 : Conclusion and Recommendations

5.1 Conclusion

An automatic synthesis for optimization of the total annualized cost of a single stage of MSF desalination process is successfully being developed. The optimization model formulated includes the constraints such as amount of distillate and brine, concentration of water in brine, energy required for the system, dimensions, i.e., diameter and length and wall thickness of the single stage flash chamber.

From the proposed optimization model for the optimization of single stage MSF desalination process with the production capacity of 30000 m^3/day, the total annualized cost (TAC) obtained is 5.108583×10^7 USD/year. The TAC includes 5.108566×10^7 USD/year of operating cost, which is mainly contributed by the cost of energy required by the process and 164.209 USD of capital cost, which is the fabrication of the single stage flash chamber.

In comparison the TAC obtained from the optimization model is 24.7% higher than the TAC estimated from previous research work by Tanvir and Mujtaba (2008). The higher TAC may be due to the number of flashing stages are not taken into account in this optimization model, which focuses on a single stage flash chamber. Besides that, the inaccuracy of using the linear equation based on the data in Tanvir and Mujtaba (2008)'s work may be another factor. Lastly, the energy cost calculation which used the cost of electricity in this research project may lead to the higher TAC as well.

5.2 Recommendations

Several recommendations have been made for the future improvement and development of this research project:

1. The number of flashing stages can be included as one of the variables in the optimization model as it can affects both the operating and capital costs and ultimately, the TAC of the MSF desalination process.

2. As the energy cost contributes the most to the TAC of the MSF desalination process, future optimization study of MSF desalination process should focus on optimization of the energy cost.

3. The optimization model proposed is developed based on the general configuration of MSF desalination process. Similar optimization models can also be developed for other configurations of MSF desalination process, such as single stage MSF, once through MSF, brine mixing MSF, MSF with brine circulation and conventional MSF.

4. Similar research project should be done on other desalination methods, especially reverse osmosis (RO) desalination process which is the most widely used worldwide nowadays.

5. The optimization study of desalination process can be done on the energy efficiency of the process.

6. A detailed technical assessment can be done to measure the technical feasibility for the implementation of the optimization model.

Chapter 6 : References

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Appendices

Appendix I: Gantt Charts

First Semester of Final Year Project:

No.	Details/Week	1	2	3	4	5	6	7	8		9	10	11	12	13	14
1	Literature reviews on related articles									Mid Sem Break						
2	Learning of GAMS programming language															
3	Research on the operating variables and cost factors															
4	Performing MEB analysis on single flashing stage															
5	Analyzing the operating variables and cost factors															
6	Integrating the operating variables and operating cost factors in the MEB analysis															
7	Developing GAMS programming for the single stage MEB analysis															
8	Analyzing the design equations related to operating variables and operating cost factors															
9	Implementing the design equations in the GAMS programming															

Second Semester of Final Year Project:

No.	Details/Week	1	2	3	4	5	6	7	8		9	10	11	12	13	14
1	Analyzing the design equations related to operating variables and operating cost factors									Mid Sem Break						
2	Implementing the design equations in the GAMS programming															
3	Analyzing the capital cost factors															
4	Analyzing the flash chamber design equations related to capital cost factors															
5	Implementing the design equations in the GAMS programming															
6	Solving all operating variables and cost factors by running the GAMS programming															

Appendix II: Calculation of Basis Data

Calculation of concentration of water in distillate:

At the pressure of 1 bar, the saturation temperature of water is 99.6°C.

Calculate the vapor pressure by using Antoine's Equation,

$$log_{10}p^* = A - \frac{B}{T + C}$$

$$log_{10}p^* = 7.96681 - \frac{1668.210}{99.6 + 228}$$

$$p^* = 749.1933mmHg$$

Applying Raoult's Law,

$$p^* = y_i P$$

$$y_i = \frac{749,1933}{750.0617}$$

$$y_i = 0.9988$$

Thus, the concentration of water in distillate is 0.9988

Appendix III: NLP Optimization Model of Single Stage Flash Chamber

Scalar
f amount of feed seawater in kg per day/50000000/

**assumption: based on inlet pressure of 2 bar*
hf enthalpy of feed seawater in kJ per kg /504.7/
xf concentration of water in feed in mass fraction /0.964/
dmin minimum demand of distillate in kg per day /30000000/

**assumption: based on outlet pressure of 1 bar*
hd enthalpy of distillate in kJ per kg /2675.4/
hb enthalpy of brine in kJ per kg /417.5/
xd concentration of water in distillate in mass fraction /0.998842/

**costing*
cm cost of 308 stainless steel in USD per ton /3565/
cp cost of power in USD per kwh /0.0080606/

**properties*
dl density of liquid water in kg per cubic meter /958.204/
dv density of vapor water in kg per cubic meter /0.5875/
k values in meter per second /0.107/

**flash chamber calculation*
ld ratio of length over diameter /3/
s maximum allowable working stress in kPa /72400/
mp maximum allowable internal pressure in kPa /1000/

**assumption: maximum allowable pressure is 10 bar*
ej efficiency of joints expressed as a fraction /1/
cc allowance for corrosion in meter /0.003/
dcs density of carbon steel in ton per cubic meter /7.83/

**operation: assume 360 days operations*
rn operating hours per year /8640/
;

Parameter vm maximum vapor velocity;
    vm = k*(sqrt((dl-dv)/dv));

Positive Variables

b	amount of brine in kg per day
xb	concentration of water in brine in mass fraction
vfr	vapor flow rate
dia	diameter of drum in meter
l	length of drum in meter
a	cross sectional area in meter squared
th	wall thickness of vessel
ta	total area of vessel
we	weight of stainless steel
qr	energy required in kWh per year

;

Free Variables

cap	capital cost
op	operating cost
q	energy required in kJ per day
d	amount of distillate in kg per day
tac	total annualized cost
wc	water cost in USD per cubic meter

*obj

;

Equation

total	objective function
const1	constraint #1
const2	constraint #2
waterbalance	water balance
diameter	diameter of vessel
length	length of vessel
csa	cross sectional area
vaporflowrate	vapor flow rate
area	total area
thickness	wall thickness
weight	weight of steel
capital	capital cost
energy	energy balance
energyreq	energy required
operating	operating cost
mass	overall mass balance
watcost	water cost

```
*objf
;

*mass and energy balance
mass.. f-b-d =e= 0;
const1.. d =g= dmin;
const2.. xd*d =l= 0.8*xf*f;
waterbalance.. (xf*f) - (xd*d) - (xb*b) =e= 0;

*capital cost
diameter.. dia =e= sqrt((4*a)/3.1416);
length.. l =e= ld*dia;
csa.. a =e= vfr/vm;
vaporflowrate.. vfr =e= d/86400000;
area.. ta =e= (3.1416*dia*l)+((3.1416/2)*(sqr(dia)));
thickness.. th =e= ((((mp)*(dia/2))/((s*ej)-(0.6*mp)))+cc);
weight.. we =e= ((ta*th)*(dcs));
capital.. cap =e= (we*cm);

*operating cost
energy.. (hf*f)+q =e= (hd*d)+(hb*b);
energyreq.. qr =e= ((q/86400)*rn);
operating.. op =e= (qr*cp);

*objective function
total.. tac =e= (op+cap);

watcost.. wc*(d*360) =e= tac

*objf.. obj =e= 1;
*;

Model capital1
/all/
;

Solve capital1 using nlp minimizing tac;
display d.l, b.l, xb.l, q.l, qr.l, op.l, cap.l, vm, vfr.l, a.l, dia.l, l.l, ta.l, th.l, we.l, tac.l,
wc.l;
```

Appendix IV: Estimation of Total Annualized Cost

The TAC for a desalination process with the production capacity of 30000 m3/day or 30000000 kg/day is estimated based on the research by Tanvir and Mujtaba (2008).

$$TAC = 1 \times 10^{-7}(Production\ capacity) + 1.0967$$

$$TAC = 1 \times 10^{-7}(30000000) + 1.0967$$

$$TAC = 4.0967 \times 10^{7} USD/year$$

Appendix V: Journals

Selected journals which are reviewed and referred in this research project are attached:

i. Optimization of design and operation of MSF desalination process using MINLP technique in gPROMS by Tanvir M.S. and Mujtaba I.M.
ii. Water desalination cost literature: review and assessment by Karagiannis I.C. and Soldatos P.G.