

The Libyan Academy
Tripoli – Libya



School of Engineering and Applied Sciences
Department of Mechanical Engineering and Energies

Modelling of MED plant in Zawia plant

A Dissertation Submitted to

The Department of Mechanical Engineering

In Conformity with the Department Requirements as Partial

Fulfilment For the Degree of Master of Engineering Science

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NOV - 2014

قرار لجنة المناقشة للطلاب

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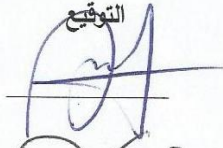
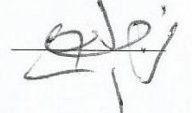
للحصول على درجة الإجازة العالية في قسم الهندسة الميكانيكية والطاقات

قامت اللجنة المشكلة بقرار السيد وكيل الأكاديمية للشؤون العلمية رقم (294) الصادر بتاريخ 26 / 10 / 2014 بمناقشة الرسالة المقدمة من الطالب / فتحي مسعود عمر اللافي لنيل درجة الإجازة العالية (الماجستير) في قسم الهندسة الميكانيكية والطاقات وعنوانها:

Modelling of MED Plant in Zawia Plant

وبعد مناقشة الرسالة علنياً على تمام الساعة العاشرة صباحاً يوم الخميس الموافق 20 / 11 / 2014 م بقاعة المناقشات بالأكاديمية وتقويم مستوى الرسالة العلمي والمنهج الذي اتبعه الطالب في بحثه قررت اللجنة ما يلي:

قبول الرسالة ومنح الطالب / فتحي مسعود عمر اللافي درجة الإجازة العالية (الماجستير) في قسم الهندسة الميكانيكية والطاقات .

أعضاء لجنة المناقشة:	التوقيع	التاريخ
1-الدكتور. المختار محمد عاشور (مشرف)		2015/1/15
2-الدكتور . المحبوب عامر دريدر (عضو ممتحن)		2015/1/18
3الدكتور . فرج المهدي الصغير (عضو ممتحن)		2015/1/25

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2015/2/5




التوقيع: التاريخ:



Acknowledgments

The great thanks for GOD first

This Research would not have been possible without help and support from many people and GECOL Company . I wish particularly to express my greatest gratitude to the following:

- My supervisor, Dr Mukhtar Mohamed Ramadan Ashour for his invaluable advice and support throughout the whole research.
- Staffs of GECOL Company in Zawia power plant.
- Finally ,my parents and my family, who has supported me during this whole period away from home, and my friends, for their patience, encouragement, understanding and continuous support throughout my research.

The full and great thanks to Academy of Graduate Studies.

ABSTRACT

This work focuses on the modeling and thermodynamic analysis of the Horizontal Tube Multi Effect Distillation plant (HT- MED) with parallel feed system ,this plant installed in Zawia Power plant in 2006. The plant has been used as a demonstration unit for seawater desalination in many places in Libya. In this study a computational model for simulation of horizontal-tube evaporators. The model based on the mass and energy balances for the streams flowing through the components (Evaporators ,Pre-heaters, flash tanks , condenser) of the MED system to predict the unit performance, produced distillate, rejected brine and temperature distribution .Visual studio program 2010 was developed based on the mathematical model of the system components. Property values of sea water are taken from Doosan company documents. The results of the model obtained were compared with design and real operation data a good agreement was observed between them. A parametric analysis where conducted to study the effect of different parameter such as seawater temperature, steam temperature and mass of steam .

المخلص

يركز هذا العمل على النمذجة والتحليل الحراري لمحطة ذات نظام أنبوب أفقي متعدد التأثير مع نظام تغذية متوازي ، هذه المحطة مثبتة في محطة توليد الكهرباء الزاوية .في 2006 وقد استخدمت وحدات المتعددة التأثير لتحليه مياه البحر في العديد من الأماكن في ليبيا . في هذه الدراسة تم كتابة نموذج حسابي لمحاكاة المبخر ذو الأنابيب الأفقية و يدرس النموذج الحالة المستقرة للنظام و النموذج مستند على أساس موازين الكتلة والطاقة وكذلك تدفق السريان لكل مكونات النظام (المبخرات، المسخنات ،صناديق التبخير، المكثف) لنظام متعدد التأثيرات، للتنبؤ بأداء الوحدة وكمية الماء المقطرا لمنتج وكمية الملح المرفوض وتوزيع درجات الحرارة وتمت كتابة برنامج حاسوب باستخدام برنامج فجول أستوديو 2010 المتطور مستند على النموذج الرياضي لمكونات النظام. وتم استخدام قيم خصائص مياه البحر مأخوذة من وثائق شركة دوسان. كما تم مقارنة نتائج النموذج المتحصل عليها مع البيانات التصميمية و الحقيقية أو التشغيلية للمحطة. ولوحظ توافق جيد بعد ذلك استخدم النموذج لدراسة تأثير بعض المتغيرات مثل درجة حرارة مياه البحر ودرجة حرارة البخار وكتلة البخار المستخدم.

TABLE OF CONTENTS

	CONTENTS	page
	Acknowledgement	
	Abstract	
	Abstract by Arabic	
	Table of contents	
	List of tables	
	List of figures	
	List of symbols	
CHAPTER 1	INTRODUCTION	1
1.1	Desalination	2
1.2	Desalination status in the world	3
1.3	Desalination Technologies	4
1.3.1	Multi-Stage Flash Distillation	5
1.3.2	Multi-Effect Distillation	8
1.3.3	Reverse Osmosis	13
1.3.4	Electro-Dialysis	14
1.3.5	Freezing	16
1.3.6	Ion exchange	17
1.4	Desalination in Libya	18
1.5	Objective and Scope of Research	19
1.5.1	Objective of Research	19

1.5.2	Scope of Research	20
1.6	Thesis Layout	21
CHAPTER 2	LITERATURE REVIEW	22
2.1	Modeling and Simulation of HT- MED plants	23
CHAPTER 3	ZAWIA MED POWER PLANT SYSTEMS	27
3.1	BASIC DEFINITIONS	27
3.1.1	Types of water	27
3.1.2	Properties of sea water	28
3.1.3	Boiling point elevation	28
3.1.4	Gain Output Ratio (GOR)	30
3.1.5	Inter effect temperature difference	31
3.1.6	The Horizontal Tube (HT) of MED	31
3.1.7	Evaporators and condenser design	33
3.1.8	Condensation inside horizontal tubes	33
3.1.9	Low temperature and brine concentration operation curve	33
3.2	Process Description	35
3.2.1	Desalination plant operating mode	35
3.2.2	Low pressure heating steam process	35
3.3	Definition of the physical system	37
3.3.1	General description of Zawia MED power Plant	37
3.3.2	Feed Water System	38

3.3.3	Heat Rejection System	39
3.2.4	Brine Discharge System	40
3.3.5	Distillate System	41
3.3.6	Plant Data	42
3.3.7	Desalination Unit Performance	47
3.3.8	Main Process Data	47
3.3.9	Technical data	48
3.4	Water salinity analysis	49
CHAPTER 4	MODEL EQUATION	50
4.1	Introduction	50
4.2	Model formulation	51
4.2.1	The main assumption	52
4.2.2	Evaporator model	52
4.2.3	Pre-heater model	58
4.2.4	Flash tank model	59
4.2.5	Condenser model	60
4.3	The Gain Output Ratio	62
4.4	Physical properties	62
4.4.1	Steady state plant analysis algorithm	62
4.4.2	Calculation procedure	63
4.5	ZAWIA MED Power Plant Simulator	66
4.5.1	DESALINATION UNIT SIMULATOR (DUS)	66

CHAPTER 5	RESULTS AND DISCUSSIONS	68
5.1	Validation of the model	68
5.1.1	The comparison of vapour and brine temperature in each effects	74
5.2	Results	76
5.2.1	The influence of steam temperature	77
5.2.2	The effect of the mass of steam on GOR and distillate produced	78
5.2.3	The effect of sea water temperature	79
CHAPTER 6	Conclusions	81
REFERENCES		83
APPENDIXES		86
Appendix A		87
Appendix B		89
Appendix C		90

LIST OF TABLES

Table 1.1	The Libya desalting plants on line in 2011	19
Table 3.1	Conductivity of various waters and equivalent NaCl-content	27
Table 3.2	Composition of standard and synthetic seawater and salt initial weight	28
Table 3.3	The main Process Data	47
Table 3.4	The technical data	48
Table 5.1	Comparison of results from the model with design data and real data in Winter and Summer.	69
Table 5.2	Error analysis between obtained results and design data in Winter and Summer.	69
Table 5.3	Comparison the vapour temperature(T_v) model with real data in Winter and Summer	75
Table 5.4	Comparison the brine temperature(T_b)model with real data in Winter and Summer	75

LIST OF FIGURES

Figure 1.1	Simple scheme of desalination systems	3
Figure 1.2	Classification of desalination processes	4
Figure 1.3	The schematic flow diagram of MSF system	6
Figure 1.4	Typical stage arrangement of a large MSF plant	6
Figure 1.5	Parallel feed Multiple Effect Distillation desalination	9
Figure 1.6	diagram of reverse osmosis operations	14
Figure 1.7	Schematic diagram of electrodialysis desalination process	15
Figure 3.1	Vapour pressure curve or boiling point curve for pure water	29
Figure 3.2	BPE as a function of salt content and temperature	30
Figure 3.3	Typical HT and VT arrangement	32
Figure 3.4	Flow behaviour in the wetting of horizontal tubes	32
Figure 3.5	Brine concentration /temperature curve for LT-MED process Operation range of low-temperature distillation	34
Figure 3.6	MED desalination units (Evaporators) of Zawia power plant	37
Figure 3.7	Feed Flow Seawater Temperature	39

Figure 3.8	Schematic of a Single Effect	43
Figure 3.9	The internal view of Effect in Zawia MED power Plan	44
Figure 3.10	In internal view (A and B)of the Flash box of effect	45
Figure 3.11	The pre-heaters of Zawia MED Power Plant	46
Figure 3.12	The Condenser of Zawia MED Power Plant	46
Figure 3.13	The Sension7 Salinity Device	49
Figure 4.1	Single Effect	52
Figure 4.2	Pre-heaters diagram	58
Figure 4.3	Flash tank schematic diagram	60
Figure 4.4	Condenser schematic diagram	61
Figure 4.5	Calculation procedure of plant analysis	65
Figure 4.6	simulator view of MED in zawia power plant as appear in computer screen	67
Figure 5.1	Bart 1 The Operation data from computer screen of Zawia MED Power plant in Summer at temperature (25.6°C)	70
Figure 5.2	Bart 2 The Operation data from computer screen of Zawia MED Power plant in Summer at temperature (25.6°C)	71
Figure 5.3	Bart 1 The Operation data from computer screen of Zawia MED Power plant in Winter at temp (18.3°C)	72

Figure 5.4	Bart 2 The Operation data from computer screen of Zawia MED Power plant in Winter at temperature (18.3°	73
Figure 5.5	Comparison of vapour temperature in each effects with model results	74
Figure 5.6	Comparison of brine temperature in each effects with model results	75
Figure 5.7	Shows the result data of the model	76
Figure 5.8	Impact the steam temperature on Gain Output Ratio	77
Figure 5.9	Impact the mass of steam on Gain Output Ratio	78
Figure 5.10	Impact the mass of steam on Distillate produced	79
Figure 5.11	Effect of temperature of seawater on Gain Output Ratio(GOR)	80
Figure 5.12	Effect of temperature of seawater on Total Distillate produced (kg/s)	80
C1	Results of med in Zawia power plant for design data at temperature (27°C)	91
C2	Results of med in Zawia power plant for data in Winter at temperature (18.3°C)	96
C3	Results of med in Zawia power plant for data in Summer at temperature (25.6°C)	101

LIST OF SYMBOLS

Symbol	Definition
A	Surface area (m^2)
CP	Specific heat capacity($\text{KJ/Kg } ^\circ\text{C}$)
m	Mass(Kg)
mbf	mass flow of vapour generated by flashing brine stream (Kg/s)
mdf	mass flow of vapour generated by flashing distillate stream (Kg/s)
N	Number of effects
P	Pressure (KPa)
Q	Total heat transfer (Kw)
T	Temperature ($^\circ\text{C}$)
U	Overall heat transfer coefficient ($\text{KW/m}^2\text{ }^\circ\text{C}$)

Greek symbols

λ	latent heat (kJ/Kg)
Δ	difference
Φ	recovery rate or yield

Subscripts

b	Brine
c	Condenser
d	Distillate
e	Evaporator
f	Feed
s	steam
t	Total
pr	Preheater
Fpr	Feed of preheater
vh	Vapour of preheater

Abbreviation

BPE	Boiling point elevation
GOR	Gain output ratio
MED	Multi effect desalination
MSF	Multi stage flash
RO	Reverse Osmosis
ED	Electro dialysis
NEA	Non equilibrium allowance
DTC	Condenser range
PR	Performance Ratio

CHAPTER ONE

INTRODUCTION

Fresh water is an indispensable resource which is vital for human consumption as well as economic development of a country. According to the World Health Organization (WHO) standards, the total dissolved solids (TDS) of drinkable water is less than 500 mg/l. Of all the available water on Earth, 97% is salt water where the TDS of the sea water is about 35000 Parts per million (ppm) and unfit for consumption.[1] Only 2 to 3% of the water resources are locked in lakes, ice caps and glaciers. It is reported that at least one billion of the world's population is presently living in water-stressed regions and this number may further increase to 3.5 billion by the year 2025.[1] Although many countries have constructed ground reservoirs to store water when it becomes available during raining seasons, the problem of global water shortage is a pressing concern for many countries, particularly in the semi-arid and desert regions due to low rainfall. It is recognized that the water shortage is caused by two major factors, first is the rapid increase in the world's population and ,the second is the relentless pursuit of economic development which leads to high water consumption of major sectors of industries. Therefore, the practical solution to solving the global water shortage is not only to develop a cost-effective technology for converting the sea or brackish water to potable water, the solution must also be environmental-friendly. [1]

1.1 Desalination

Simply, the two main methods of removing salt from sea water currently in use for large scale applications are: distillation and reverse osmosis. In distillation, sea water is heated until it is boiled. The salt remains in the water; the vapour is captured and condensed into fresh water.

In reverse osmosis, hydrostatic pressure is applied to separate fresh water from sea water through a semi-permeable membrane, which will reject the salt and the fresh water will be collected as permeate.

In both techniques, the leftover brine is piped back into the sea or ocean. About 6-7 % of that water is salt, compared with 3.5% for regular sea water.[2]

Figure (1.1) shows the desalination systems in a simple scheme, a typical desalination plant consists of a water pre-treatment system, the desalination unit and a post-treatment system. Several desalination processes have been developed but not all of them are reliable and in commercial use. The most important processes are split into two main categories ,thermal (or distillation) and membrane processes.

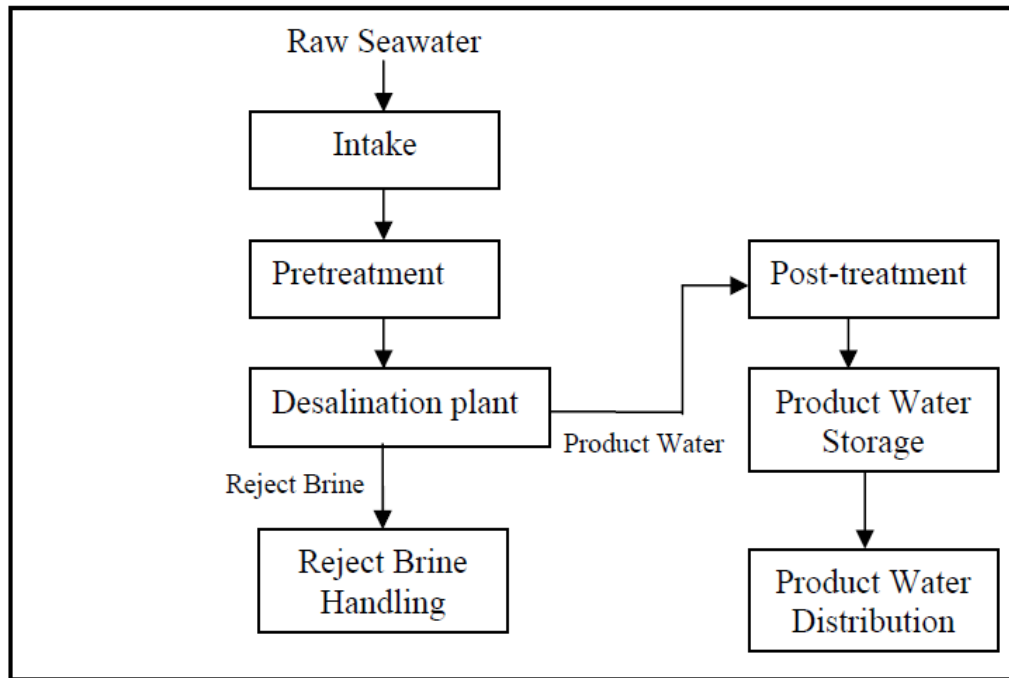


Figure 1.1 Simple scheme of desalination systems[3]

1.2 Desalination status in the world

Desalination has often been perceived as a high cost means of producing water. Desalination processes have so far mostly been built when:

(a) there is no practical alternative , (b) cost of other resources is high , (c) low-cost energy is readily available , and (d) high living standards override the cost factor. These conditions are met in the Arabian Gulf states, and to a lesser extent in the Mediterranean area. In underdeveloped countries there is usually a pressing need for water, but the economic conditions have, so far, excluded desalination.

1.3 Desalination Technologies

A desalination process essentially separates saline water into two parts one that has a low concentration of salt (treated water or product water), and the other with a much higher concentration than the original feed water, usually referred to as brine concentrate or simply as ‘concentrate’. The two major types of technologies that are used around the world for desalination can be broadly classified as either thermal or membrane separation . Both technologies need energy to operate and produce fresh water. Within those two broad types, there are sub-categories (processes) using different techniques, Freezing and Ion Exchange as shown in. [5]. Figure1.2 shows the classification of desalination processes

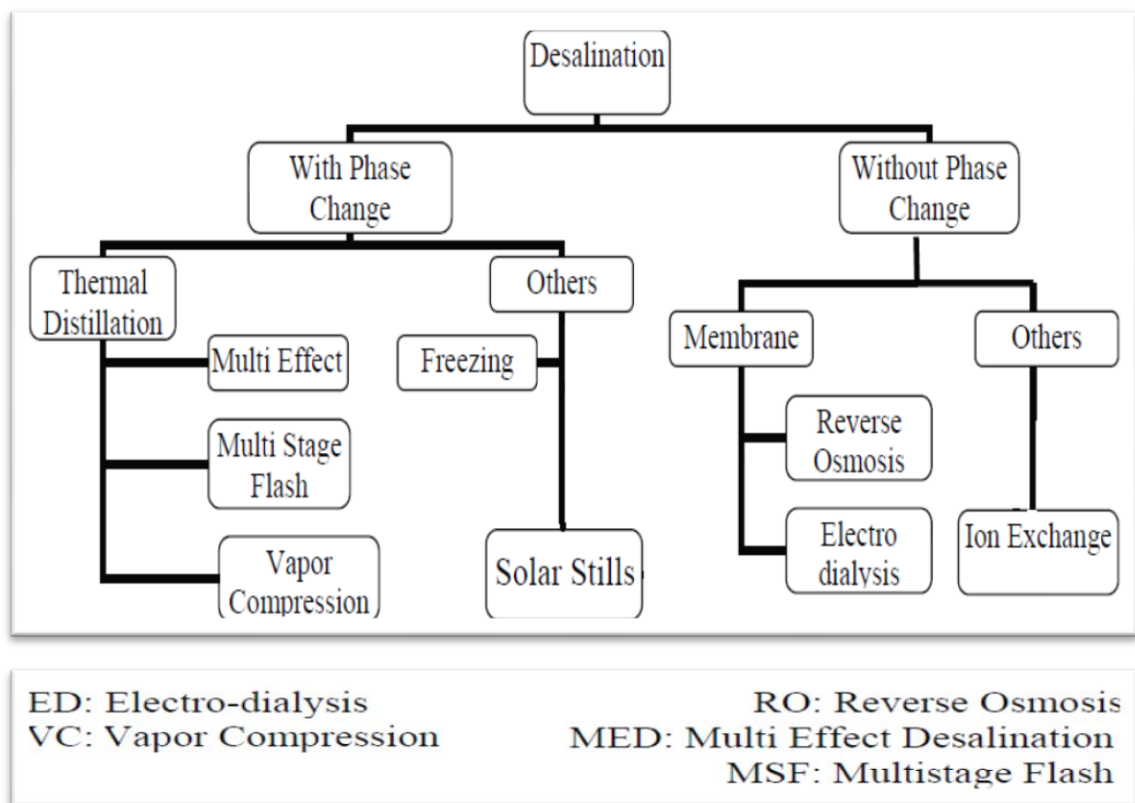


Figure 1.2 Classification of desalination processes

1.3.1 Multi-stage flash distillation

In Multistage Flash (MSF) Desalination, vapour formation takes place within the bulk liquid and not on the surface of hot tubes. The hot and concentrated brine is allowed to flow freely and flash off in a series of chambers, so it is kept away from the tube surfaces during vapour formation.

MSF configurations applied on an industrial scale include the once-through (MSF-OT) arrangement and the brine recirculation (MSF-BR) system. The use of the former is presently limited; most MSF plants rely on brine circulation, a method particularly suited to regions with large daily and seasonal temperature fluctuations. In general, brine recirculation results in a higher conversion ratio, uses smaller amounts of chemical additives and allows better control of feed seawater temperature.

Figure (1.3) and Figure (1.4) shows the schematic flow diagram of an MSF system and typical stage arrangement of a large MSF plant respectively. [7]

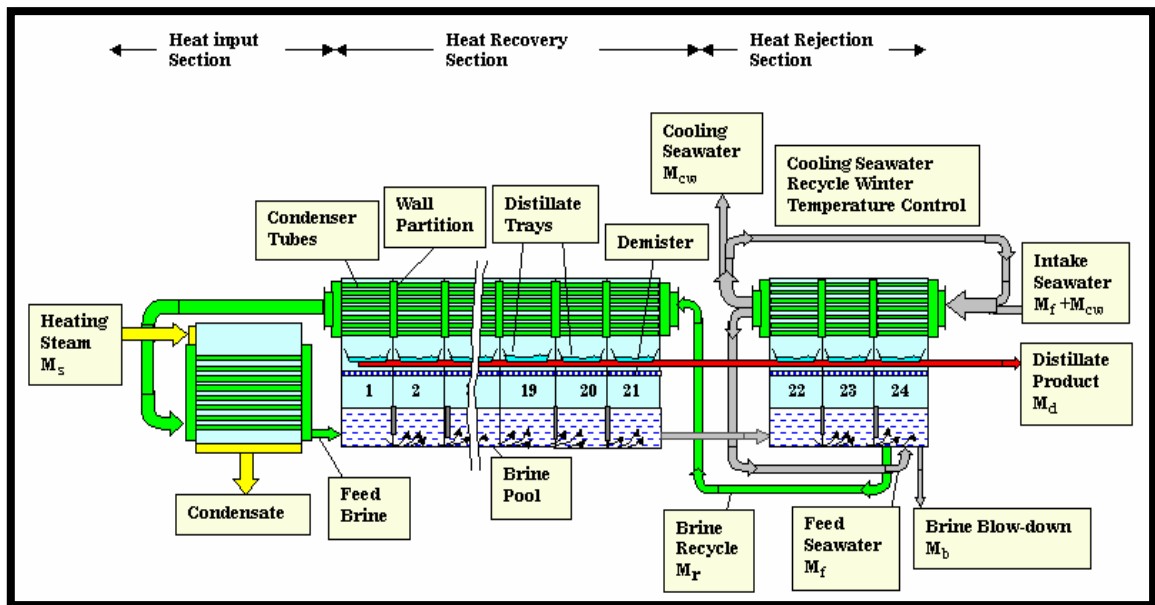


Figure 1.3 The schematic flow diagram of MSF system[7]

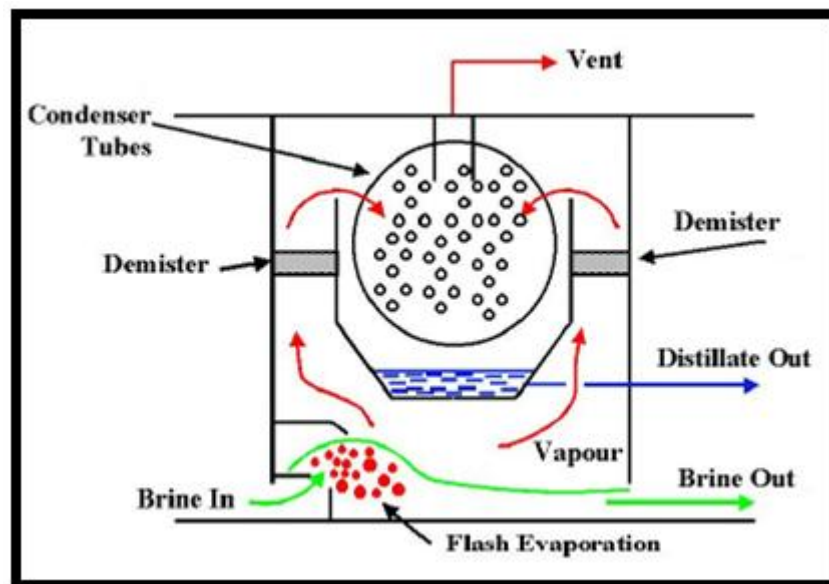


Figure 1.4 Typical stage arrangement of a large MSF plant[7]

In the MSF process, water boils at temperatures below the saturation boiling temperature; when intruded to lower pressure. This is referred to as the “flashing effect”. Feed water is heated in a vessel called the brine heater, before being allowed to flow into a series of chambers, known as “stages”, that constitutes the “evaporator” in the MSF unit. Most stages are

maintained at reduced pressure, relative to atmospheric pressure, so that the sudden introduction of heated feed water into these chambers causes rapid boiling, or “flashing”. Steam generated by flashing is converted into fresh water by condensation on tubular heat exchangers that exist above the brine Pool in each stage. After passing through dim ester to remove any saline water droplets entrained by vapour. The tubes are cooled down by incoming feed water on its way to the brine heater. This has the effect of warming up the feed water so that the amount of thermal energy needed to raise its temperature in the brine heater is reduced. Distillate flowing from stage to stage is taken out as product water from the last stage. It may then be chemically treated to adjust its pH and hardness prior to storage or usage. Most MSF plants operate in a dual-purpose or cogeneration mode, incorporating both power generation and water desalination. Waste or extracted heat produced in electricity generation units is used to preheat feed water, resulting in high thermal efficiencies and improved process economics. Top brine temperature (TBT), the temperature of brine leaving the brine heater) is a major factor in the design of an MSF unit. Increasing TBT values offers to the designer a clear thermodynamic advantage by raising the overall efficiency of the MSF distillation process; however, a high TBT has important implications for the durability of the unit components There is a direct link between TBT and the measures applied to reduce or prevent corrosion and control scale. If scale inhibition is based

on the use of organic polymers and sponge ball circulation for tube cleaning, a TBT limit of about 110°C is imposed to minimize alkaline scale formation.[5] When acid dosing is used for scale control, the TBT may be allowed to reach 120°C .[7]

1.3.2. Multi-Effect Distillation

Multiple Effect Distillation (MED) is the oldest process used for seawater desalination. Reliable design, technological maturity, high quality of distillate, good operating records and high unit capacity are its main benefits. MED occurs in a series of effects and uses the principle of reducing the ambient pressure in the various effects; this permits the feed water to undergo multiple boiling without supplying additional heat after the first effect. MED reuses the heat of vaporisation by replacing evaporators and condensers in series and is based on the principle that vapor produced by evaporation can be condensed in a way that uses the heat of vaporization to heat brine at a lower temperature and pressure in the following chamber. Figure (1.5) shows MED plants typically comprise units producing 2,000 to 13,000 m³/d.[7]

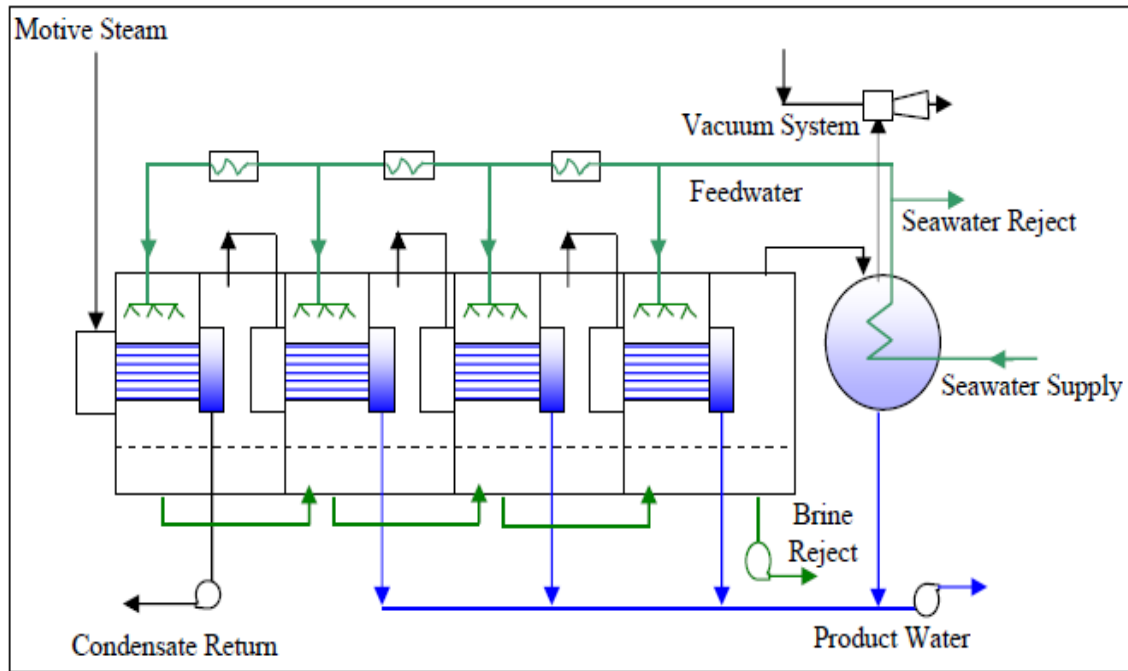


Figure 1.5 Parallel feed Multiple Effect Distillation desalination [7]

Between 8 and 16 effects are generally used in large MED plants; performance ratios are typically around 12, internal electric power consumption is about $2\text{--}3 \text{ kWh/m}^3$, and production ratios range between 45 and 120 per cent of nominal capacity.[7] The incoming feed water is pumped into the plant through a number of pre-heaters located in each effect in order to raise its temperature. After passing through the last of these, feed enters the top effect where the heating steam from a boiler or other source raises its temperature to the saturation one for the effect pressure. The second and the subsequent effects use steam produced by the previous effects. Even for higher concentration factors, the once-through principle can be realised as a consequence, concentrated brine will be in contact only with the effects operating at lower temperature. Because the distillate production, between maximum capacity and zero production, is

totally controlled by the amount of live steam, the mode of operation appears relatively simple.

Moreover, high overall heat transfer coefficients can be achieved permitting small temperature differences between the effects at economical rates of heat transfer. In some cases, the feed water is sprayed onto the surface of the evaporator tubes to form a thin film to promote rapid boiling and evaporation. The vapour produced then goes, in part, to heat the incoming feed and in part to provide the heat supply for the second effect which is at lower pressure and receives its feed from the brine of the first effect. From the second effect the vapor itself is condensed while giving up heat to evaporate a portion of the remaining feed water in the next effect. This process is repeated downstream the plant. The effects are gradually operated at lower temperatures; this is accomplished by maintaining the effects at gradually lower pressure.[7]

Depending on the combinations of heat transfer configurations and the flow-sheet variations (horizontal/vertical, recirculation, once-through, etc.) that may be selected independently, a number of MED variants exists:

- Multi-Effect Submerged Tube.
- Multi-Effect Vertical Tube Climbing-Film.
- Multiple-Effect Vertical Tube Falling-Film.
- Multi-Effect Horizontal Tube Falling-Film.

Horizontal falling-film type multiple-effect distillers are considered the

workhorse of desalination concerning the MED technology. Recent MED plants have been built to operate with top temperature (in the first effect) below 70°C in order to minimise corrosion and scaling problems; this leads to an efficient use of low grade heat.[7]

Advantages of low temperature MED plants (LT-MED) are the most efficient thermal desalination processes currently in use. It incorporates technological advances which have resulted in reliable, durable and economical desalination plants producing high purity product water.[8]

These advances are as follows:

1. Development of a unique design of a falling film horizontal tube evaporator/condenser with high heat transfer coefficient, utilizing only latent-heat transfer, avoiding sensible heat pick-up.
2. Superior thermodynamic efficiency and very low pressure drops at high volumetric vapor flows, as prevailing in low temperature operation. This enabled the optimization of the process for operation at a maximum brine temperature of 70°C.
3. The low temperature operation aided by a comprehensive multi-disciplinary development and design approach has made possible the utilization of economical and durable materials of construction such as aluminum alloy for heat transfer tubes, plastic process piping and epoxy-painted carbon steel shells which show a better resistance when matched with aluminum alloy or titanium.

4. The economy of using aluminum tubes for heat transfer as compared with copper alloy tubes, which are essential for higher temperature plants (used by other distillation manufacturers), enables the increase of the heat transfer area per ton of water produced in the desalination plant for the same investment costs.
5. The significant increase in heat transfer area, in addition to the thermodynamic superiority of MED over the MSF process, results in a very low temperature drop per effect (1.5 , 2.5°C), enabling the incorporation of a large number of effects (10–16) even with a maximum brine temperature as low as 70°C, consequently resulting in very high economy ratios (product to steam).
6. Possibility of using low-cost/low-grade heat available through cogeneration schemes to minimize the energy cost component.
7. Minimal requirements for intake and pretreatment systems.

The practical experience with commercial plants using the above mentioned advances has shown the remarkable stability, flexibility and reliability of the low temperature process in comparison with others. Continuous research and development improved further the advantages of the low temperature process by increasing the unit's capacities; decreasing the energy consumption; and lowering levels of scaling and corrosion, both being significantly reduced due the low temperature operation[8].

1.3.3. Reverse Osmosis

Reverse osmosis (RO) (Figure 1.6) is a membrane separation process that recovers water from a saline solution pressurized to a point greater than the osmotic pressure of the solution. The United States ranks second worldwide in desalination capacity, primarily relying on RO to treat brackish and surface water [9].

In essence, the membrane filters out the salt ions from the pressurized solution, allowing only the water to pass. RO post-treatment includes removing dissolved gasses (CO_2), and stabilizing the pH via the addition of Ca or Na salts.

Pressurizing the saline water accounts for most of the energy consumed by RO. The osmotic pressure, and hence the pressure required to perform the separation is directly related to the salt concentration. RO is often the method of choice for brackish water, where only low to intermediate pressures are required. The operating pressure for brackish water systems ranges from 15-25 bar and for seawater systems from 54 to 80 bar (the osmotic pressure of seawater is about 25 bar). Since the pressure required to recover additional water increases as the brine stream is concentrated, the water recovery rate of RO systems tends to be low. A typical recovery value for a seawater RO system is only 40%. [10].

Most of energy losses for RO result from releasing the pressure of the concentrated brine, large scale RO systems are now equipped with devices

to recover the mechanical compression energy from the discharged concentrated brine stream with claimed efficiencies of up to 95% [11].

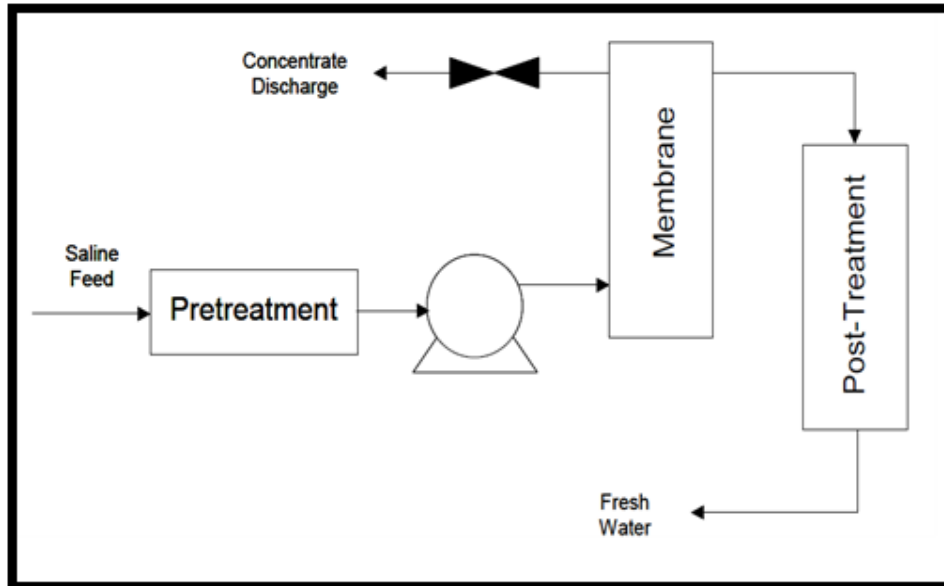


Figure 1.6 Diagram of reverse osmosis operations[7]

1.3.4. Electro-Dialysis

Electrodialysis (ED) utilizes a direct current source and a number of flow channels separated by alternating anion and cation selective membranes to achieve the separation of water and dissolved salts (Figure 1.7) [12]. The driving force for the separation is an electric field, ED is only capable of removing ionic components from solution, unlike RO or distillation.

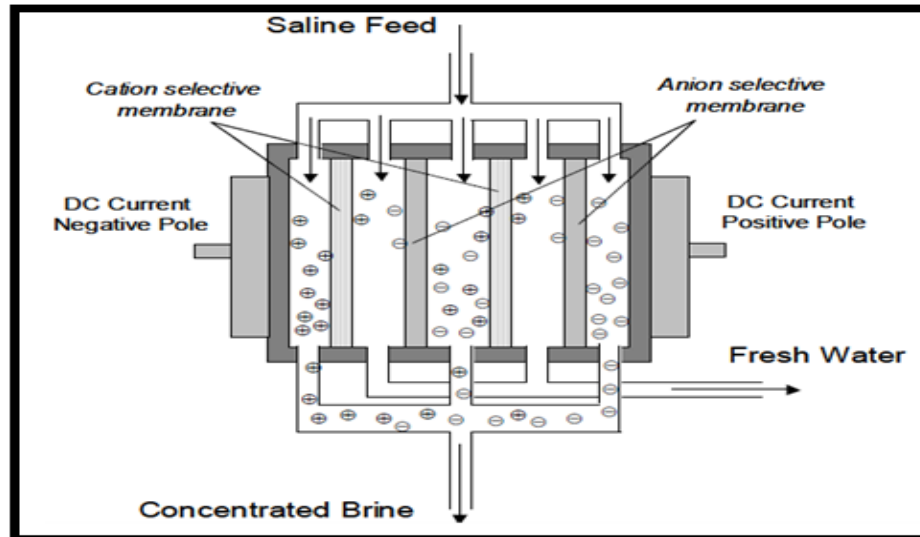


Figure 1.7 Schematic diagram of electrodialysis desalination process[12]

In the ED process, saline water is fed in parallel to each of the separate channels. Cations and anions then migrate in opposite directions in response to the applied voltage. Due to the charge selectivity of the membranes, the ion concentration increases and decreases in alternating channels of the apparatus. A single membrane stack may consist of hundreds of these alternating channels. Since the resistance in the stack changes from top to bottom, the separation is typically carried out in a series of small steps. This makes the process more economical and easier to control.

Like RO, the energy required to separate the ions from solution increases with concentration, thus ED is generally limited to brackish waters containing only a few thousand ppm of dissolved solids.[10]

The membrane of ED units are subject to fouling, and thus some

pretreatment of the feed water is usually necessary. Precipitation of scale can be facilitated in the ED process by changes on pH that occur near the membranes as a result of the transport of H^+ and OH^- ions [10].

However, since there is not a flux of water through the membranes, ED can treat water with a higher level of suspended solids than RO. Also, since nonionic solids, e.g. silica, are not concentrated by the process, these components are of less concern. The electrodialysis reversal (EDR) process was developed to help eliminate membrane fouling. In the EDR process, the membrane polarity is reversed several times an hour. This has the effect of switching the brine channels to freshwater channels, and the freshwater channels to brine channels, and breaks up and flushes out deposits [11].

1.3.5. Freezing

In 1786, Anton Maria Lorgna gave the first description of working method for desalination by freezing [13] Extensive work was done in this area in the 1950s and 1960s. Desalination by freezing process is carried out by cooling the salt water to form ice crystals under controlled conditions and then melting the ice to get pure water. It has several advantages such as low energy consumption, minimal potential for corrosion and low scaling probability. However, it involves handling of ice and water mixtures. In practice, the surface of the ice remains coated with brine, resulting in high salinity of the product water. To avoid this, the ice is washed with pure

water. The amount of pure water required for washing is large. Several pilot plants have been built by using freezing for desalination. However, it has still not met commercial success.[12]

1.3.6. Ion Exchange

Ion exchangers are organic or inorganic solids that are capable of exchanging one type of cation (or anion) immobilized on the solid for another type of cation (or anion) in solution. For example, Na^+ ions in solution can be replaced with H^+ by a cation exchanger and Cl^- can be subsequently be replaced with OH^- by an anion exchanger resulting in the complete “demineralization” of a NaCl solution. The process can be reversed by regenerating the cation exchanger with an acid, and the anion exchanger with a base. A typical ion exchanger is capable of exchanging about 5 mill equivalents of ions for each gram of solid material.

In concept ion exchangers appear attractive for desalination, since they focus on the minor component, the salt, rather than the major component, the water. In practice, ion exchange is a useful process for completely demineralizing water in applications where high purity is required, e.g. high pressure boilers.[14]

1.4 Desalination in Libya

Libya is sited in North Africa with total area of (1,775,000 km²), with coast extends for (1950 km), along the southern coast of the Mediterranean Sea starting from the border with Tunisia at the west to Egypt at the east. The IDE worldwide desalting plants inventory (2011), informs on 285 desalting plants installed in Libya since 1962. The total capacity of desalinated water production was served by 747.906m³/day.[15]The desalination plants projects is considered as one of the important projects as it is the other choice for many Arab countries including Gulf and North Africa countries. Table 1.1 shows the Libya desalting plants on line in 2011..

Table 1.1 The Libya desalting plants on line in 2011.[15]

Denomination /Location	Method	Production m³/day	Raw Water	Year	Use
Osmo 10 AM	RO	240	Brackish	2007	Industry
Osmo 12 AM	RO	500	Brackish		Industry
Osmo 60 AM	RO	2.000	Brackish		Tourism
Osmo 40 AN	RO	1.500	Brackish		Tourism
Ras Lanuf	MSF	8.400	Seawater	1993	Municipal
Ras Tajura	ED	1.870	Brackish	1996	Industry
Shahhat	MSF	13.500	Seawater		Municipal
Soussa	MED	10.000	Seawater	1997	Municipal
Soussa	MED	26.000	Seawater	2010	Industry
Subrata	RO	1.000	Brackish	2000	Municipal
Subrata	RO	500	Brackish	2001	Municipal
Surman City	RO	500	Brackish	2001	Municipal
Tobruk	MED	750	Seawater	1995	Industry
Tobruk	RO	500	Seawater	2000	Tourism
Tobruk II	MED	40.000	Seawater	2000	Municipal
Trípoli	RO	650	Seawater	1995	Municipal
Trípoli	RO	376	Seawater		Industry
Trípoli	RO	1.000	Seawater	1995	Municipal
Trípoli	RO	2.500	Seawater	1995	Municipal
Trípoli	MED	10.000	Seawater	1997	Industry
Trípoli	MED	1.200	Seawater	2002	Power
Trípoli	RO	288	Brackish	2009	Industry
Trípoli	RO	120	Brackish	2009	Industry
Trípoli	RO	48	Brackish	2009	Industry
Trípoli Project	RO	240	Brackish	2008	Industry
Wafa	RO	2.000	Seawater	2002	Industry
Wafa	MSF	16.000	Seawater	2003	Industry
Zanzur	RO	130	Brackish	1996	Discharge
Zawia	MED	5.000	Seawater	2005	Power
Zawia	MED	80.000	Seawater	2010	Municipal
Zliten	MSF	30.000	Seawater	1998	Municipal
Zuara	MED	40.000	Seawater	2005	Municipal

1.5 Objective and Scope of Research

1.5.1 Objective of Research

The desalination plant which is the subject of study assembled and operated in the vicinity of the Zawia Combine Cycle Power Plant complex.

The plant consists of two MED desalination units with a design capacity of 2,500 m³/day . The study will be aimed to the following :

1. To developed steady state mathematical model of MED in Zawia power

Plant.

2. Write a computer programme which can be used to train MED operators to give them deep understanding of the system ,where the cause and effect relationships between variable cab be seen more clearly.

1.5.2 Scope of work

Theoretical research was conducted to study the above-described concept and develop a practical model based on this concept. The following tasks were performed:

1. A thermodynamic analysis of each component of the whole process was conducted. A mathematical model was developed that can be used in evaluating the performance of the system, by solving a set of equations from the thermodynamic model.
2. Validate the simulation results with the design data and compared with existing real operation data gathered from Desalination plant

.

1.6 Thesis Layout

Based on the objective described earlier. Chapter2 discuss some relevant literature and past works. Chapter 3 basic definitions in desalination technology and description of multi-Effect Desalination system will be highlighted. Some theoretical related subjected will also be included.

Chapter 4 deals with development of one region model for HT-MED plant .All the component of the MED plant are described in details .The mathematical formulation and computational method employed are given in details.

In chapter 5 validation of model , results and discussion from the model development earlier will be presented .Comparisons between result of model and design data and real operation data.

Finally the conclusion of the work carried out in this study is given in chapter 6.

CHAPTER TWO

LITERATURE REVIEW

With the increase of the world population, the need for sustainable water is increasing day by day. Desalination is an effective method to obtain fresh water from seawater and brackish water. Desalting techniques have captured attention as an option to increase the range of water resources available for use by a community. However, as desalination of seawater and brackish water is an energy intensive process, the need for economically feasible fresh water increases steadily. Therefore, cost effective desalination system is always an attractive option to meet the water crisis. Of the different desalination techniques, evaporative technology is still popular and competing with non-evaporative processes, such as, reverse osmosis (RO). Among different thermal desalination techniques, multi-effect distillation (MED) is technically and economically a good option. MED can use low temperature and low pressure steam as its heating source. Over the years, waste heat utilization in MED is growing increasingly popular. The challenges, however, exist in the improvement of performance of MED with efficient evaporator design and control of fouling.

A research effort has been undertaken in this context to build an efficient thermal desalination system utilizing waste heat in the form of hot water.

2.1 Modelling and Simulation of HT- MED plants

El- Dessouky et al. [16] considered a steady state analysis of multiple Effect Evap Desalination process was developed to determine the important design and operating variables controlling the cost of producing fresh water. The model assumed constant heat transfer areas for the evaporators and feed preheaters in all effects. The variation in thermodynamic losses from one effect to another, the dependence of the physical properties of water on salinity, temperature and the influence of no condensable gases on the heat transfer coefficients in the evaporators and the feed preheater. Simulation results showed that the heat transfer coefficients increase with the boiling temperature. The plant performance ratio was found to be nearly independent of the top brine temperature and strongly related to the number of effects. The required specific heat transfer areas at a top brine temperature of 100°C were 30.3% and 26% of that required at 60°C when the number of effects were 6 and 12, respectively.

Khairy Agha, et al ,[17] presented a steady state thermal and hydraulic behavior of thermal desalination plants (MSF and MED) along with the necessary subsystems namely brine heat and steam ejector were developed and discussed. APROS simulation model for the coupling unit utilizing two different concepts has also been developed. Considering the results of this model for each set of circumstances. APROS Simulation Models for MSF and MED were developed and compared against Design Data. A model for

the coupling unit taking into account the pressure reversal condition was developed. The APROS Simulation Model were validated and assessed against a real design data of 1200m³/day capacity.

* APROS is a process simulation software which is capable of modeling thermal hydraulic processes and the APROS is one of the well advanced soft wares available.

El-Methnani [18] presented an unsteady state model for single HT-evaporator based on two regions ,a top regions for the vapour and a bottom one for the brine . It is basically a one pressure , two temperature model .Correlations of heat transfer coefficient are used for single tube and assuming laminar flow conditions.

Mukhtar Ashur [19] presented two mathematical models to simulate Horizontal Tube Multi- Effect Desalination (HT-MED) based on tow regions; 1st one rejoin model and 2nd.The first is designated for water vapour and the second is for brine bool .The models also consider the designs of the condenser and the flash tanks .In first model overall heat transfer coefficients are derived using simple curve fitting equation based on published data . For the second model , the overall heat transfer coefficient is based on correlations for heat transfer coefficients of the outer and inner surfaces of the evaporator tubes.

Hanbury [20] provided a steady-state solution to the performance equations of an MED plant. The simulation was based on a linear decrease in boiling

heat-transfer coefficient, unequal inter-effect temperature differences, and equal effect thermal loads from the second effect down.

Husain et al [21] described the modeling and simulation of a multistage desalination plant with 15 recovery stages and three rejection stages. The study was based on both steady-state and dynamic simulations; the study was carried out using a FORTRAN program for the steady-state simulation and also through a SPEEDUP package.

Hamed [22] investigated the thermal performance of a ME desalination system. An analytical solution was developed to verify the impact of different process variables on the performance of the MED system as number of effects, TBT, inlet seawater and the amount of product. The dependence of the water production cost on the performance of the plant was also studied. The results showed that the performance ratio is highly dependent on the number of effect, and both the inlet seawater temperature and TBT are slightly affected on the plant performance ratio.

Khademi [23], focuses on the development of a steady-state model for the multi-effect evaporator desalination system. The paper presents the model equations, method of solution, optimization of operating conditions by sequential simplex method, optimization of operating time, and the effect of feed mass flow rate, feed temperature and condenser pressure on GOR, consumed steam and produced distilled water. The paper presents new plant data which is interesting for industrials.

As seen from the previous review computer simulation programs are widely used in the desalination industry, a lot of work has been reported so far on the modeling of the MED plants .

Most of the model discussed previously are not complete or describing a very specific plant. In this research the MED parallel feed desalination in Zawia power plant has been modelled at steady state.

CHAPTER THREE

ZAWIA MED POWER PLANT SYSTEMS

3.1 BASIC DEFINITIONS

3.1.1.Types of water

The earth contains about $1.4 \times 10^9 \text{ km}^3$ of water, which covers approximately 70% of the planet surface area; the percentage of salt water in this large amount is 97.5%. The remaining 2.5% is fresh water with 80% of this amount frozen in the icecaps or combined as soil moisture..Both forms are not easily accessible for human use. The remaining quantity, about 0.5%, is believed to be adequate to support all life on Earth. Table (3.1) gave the conductivity of various waters and equivalent NaCl- conten [24]

Table:3.1Conductivity of various waters and equivalent NaCl- conten. [25]

Type of water	Conductivity [$\mu\text{S}/\text{cm}$]	Equivalent NaCl content [mg/l]
Deionised water	0.2	≈ 0.1
Distillate	5	≈ 3
Drinking water	400	≈ 205
Rain water	60-100	$\approx 30-50$
Ground water	100-2,500	$\approx 50-1,280$
Mineral water	500-2,500	$\approx 260-1,280$
Seawater	45,000	$\approx 23,000$

3.1.2 Properties of sea water

Typical values for the conductivity of various types of water are summarised in Table (3.2) shows the composition of standardised seawater, synthetic seawater and the relevant salt initial weight to produce this synthetic water.

Table 3.2 Composition of standard and synthetic seawater and salt initial weight . [25]

Substance	Standard seawater	Synthetic seawater	Salt initial weight
	[g/l]	[g/l]	[g/l]
SO_4^{--}	2.7689	2.7683	28.050 NaCl
Cl^-	19.8406	20.0305	5.980 $\text{MgCl}_2 \cdot 6 \text{H}_2\text{O}$
Ca^{++}	0.4182	0.4182	6.299 $\text{MgSO}_4 \cdot 7 \text{H}_2\text{O}$
Mg^{++}	1.3297	1.3295	1.534 $\text{CaCl}_2 \cdot 2 \text{H}_2\text{O}$
Na^+	11.0348	11.0347	0.618 K_2SO_4
K^+	0.3972	0.3972	0.021 KHCO_3
Sr^{++}	0.0139	-	0.214 KCl
HCO_3^-	0.1460	0.0125	-
Br^-	0.0675	-	-
F^-	0.0013	-	-
H_3BO_3	0.0272	-	-
Salt	36.0453	35.9909	
Water	963.9547	964.0091	

3.1.3 Boiling point elevation

The generated vapour is at the saturated temperature , which corresponds to the pressure in the evaporator vapour space. This temperature is less than

the boiling temperature by boiling point elevation (BPE), where the BPE at a given pressure is the increase in the boiling temperature due to the salt dissolved in the water. Figure(3.1) show the Vapour pressure curve or boiling point curve for pure water. BPE has a direct effect on the rate of the heat transfer , that is because of the reduction in the temperature potential for heat transfer , where the temperature difference is reduced by the amount of BPE .Figure(3.2) shows BPE as a function of salt content and temperature. In the MED plants where evaporator are connected to each other in series the BPE of the individual effects are cumulative, this fact is important factor in the determination of the optimum number of effect for a given MED plant [26]. Correlation for (BPE) of seawater is given in Appendix(A).

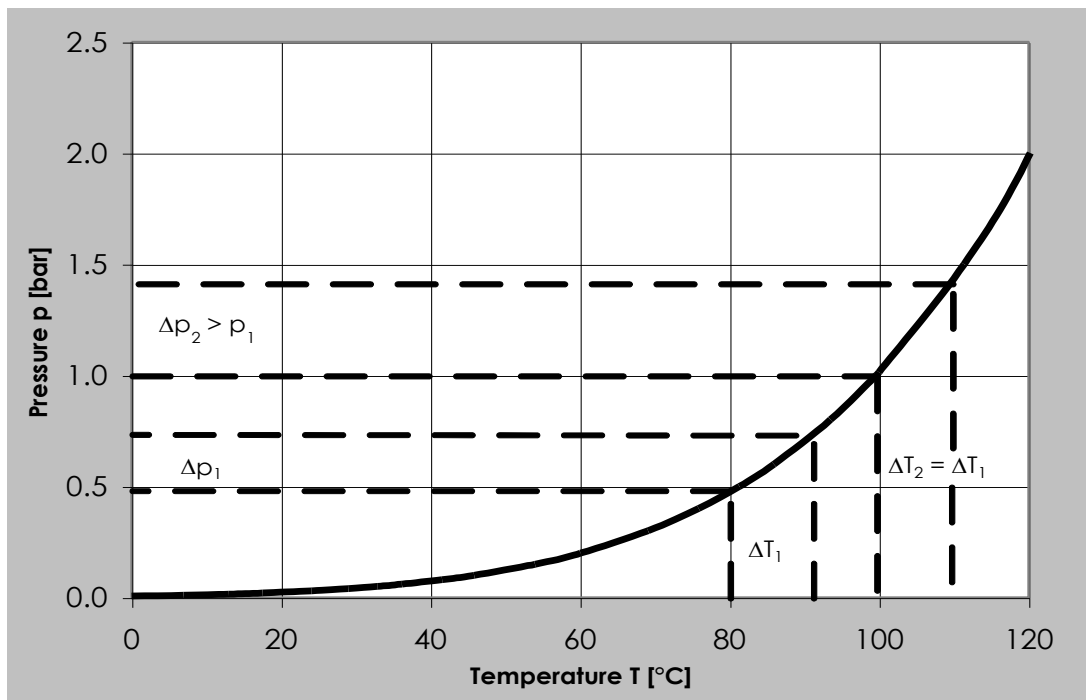


Figure 3.1 Vapour pressure curve or boiling point curve for pure water[25]

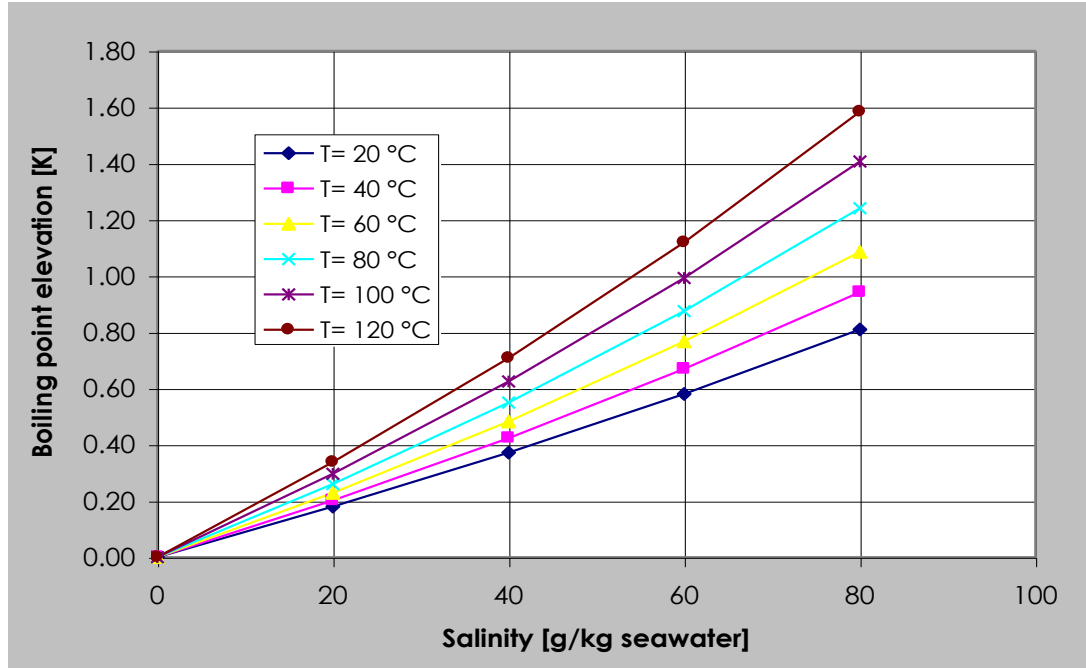


Figure 3.2 BPE as a function of salt content and temperature[25]

3.1.4 Gain Output Ratio (GOR)

The gain output ratio is defined as the mass of fresh water produced from the plant divided by the mass of steam used to produce this fresh water; hence it is a measure of the thermal efficiency of the distillation plant which is in the range of 6 to 10 for current commercial multi-stage flash (MSF) distillation plants and up to 20 for multi effect distillation (MED) plants. However, the GOR does not account for the steam temperature and therefore does not assign a thermodynamic value to the steam. Thus, comparing different distillation plants by means of the GOR is only useful if the temperature difference between the heating steam and the seawater is the same, a distillation plant with a GOR of 8 requiring steam of 70°C, for instance, is thermodynamically superior to a distillation plant with the same GOR but requiring steam of 120°C [16]

3.1.5 Inter effect temperature difference

The inter effect temperature difference is the temperature difference between one effect and the next. It is made up of contribution from a number of sources, which is boiling point elevation (BPE), temperature drop due to film evaporation of the brine, temperature loss due to pressure drop as the vapour generated in effect has to be transported to the second effect. In the tubes of the second effect the vapour will condense forming distillate film across which the latent heat must be conducted with temperature drop. The latent heat is then conducted through the tube wall to the boiling brine on the other side with tube wall temperature drop. By keeping the inter effect temperature difference as minimum as possible the number of effects in a given temperature range will increase. Temperature range is limited by scale formation and temperature of available steam as a top temperature and bottom temperature is limited by sea water temperature which could rise up to 30°C in summer. Also inter effect temperature difference depends on material of the heat transfer surface, vapour transportation path and type of demister used. [27]

3.1.6 The Horizontal Tube (HT) of MED

Most modern large MED plants have a horizontal tube (HT) configuration, with the feed water film sprayed over the outside of the tubes and condensation inside. The horizontal tube design uses a system of spray nozzles or perforated trays to distribute the feed water evenly over the heat

transfer tubes. In the vertical tube (VTE) configuration this is reversed, with the feed water flowing down the inside of the tubes. The two arrangements are shown in Figure(3.3).[27] Figure(3.4) shows the flow behaviour in the wetting of horizontal tubes.

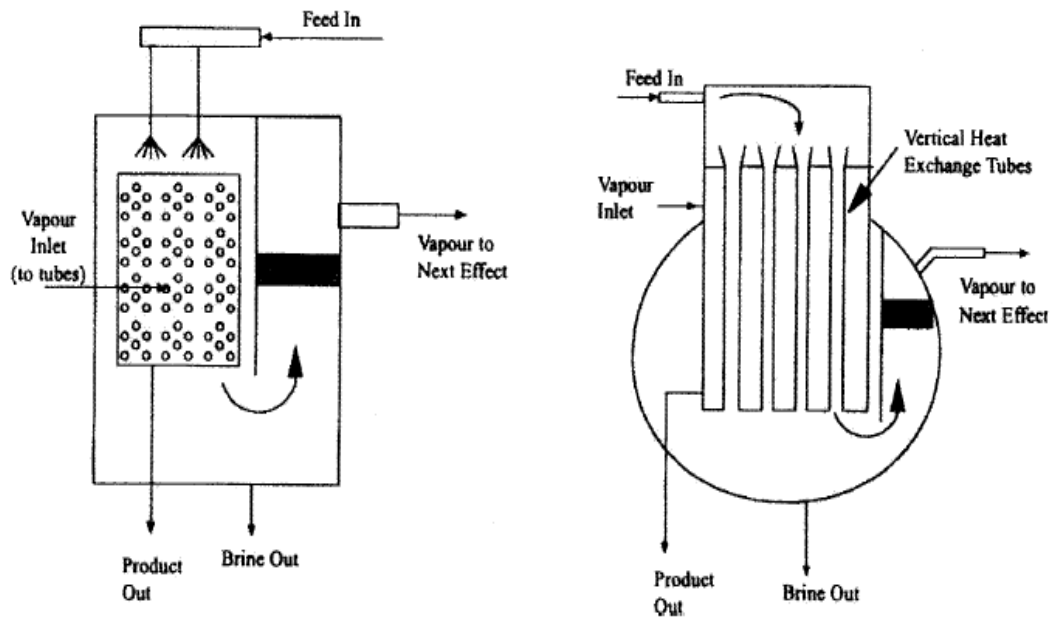


Figure 3.3 Typical HT and VT arrangement[25]

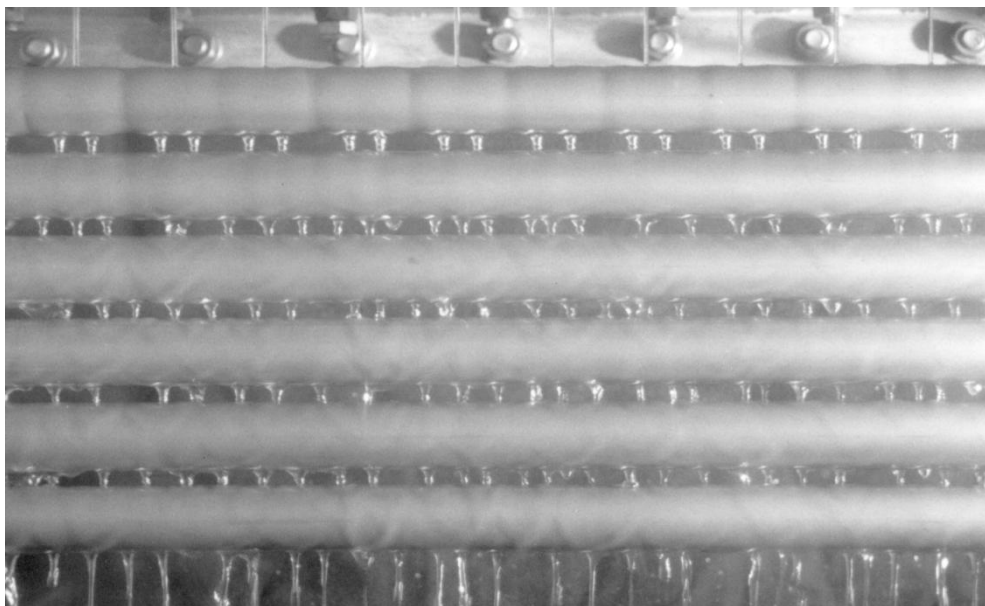


Figure 3.4 Flow behaviour in the wetting of horizontal tubes[25]

3.1.7 Evaporators and condenser design

Evaporation is the removal of solvent as vapor from a solution. It is the operation which is used for concentration of solution. There could be single effect evaporator or multiple effect evaporators. With addition of each effect steam economy of the system also increases.

Evaporators are the main part of desalination units. Therefore, it is very important to design them more cost effective and more efficient. To design a shell and tube evaporators, there are some parameters should be calculated such as tube size (diameter and length), number of pass and number of tubes[28]

3.1.8 Condensation inside horizontal tubes

The seawater flows over the outside of the tubes, while the vapour condenses inside the tubes. The heat of condensation is transferred through the wall into the water film on the outside of the tube and leads to a partial evaporation of the water streaming down.

The condensation of the vapour inside the tube is comparable in principle with the condensation of the vapour on the outside of the tube.[28]

3.1.9 Low temperature and brine concentration operation curve

The operating temperatures are well below the saturation limits of problematic scalants found in sea and ground waters (Fig 3.5). Scale is reduced to an insignificant level, enabling plants to operate for long

periods—5 years in some cases—between chemical cleanings. Low cost polyelectrolyte feed pre-treatment is convenient. Descaling simply consists of mild acid recirculation, using the plant's own recirculation pumps.. The reduced corrosiveness of seawater, at the low operating temperature and vacuum conditions (deaerated feed water), allows safe and economic use of corrosion proof materials and coatings both for piping and for vessels lining, as well as the use of aluminum for heat transfer tubing and vessel internals[8]

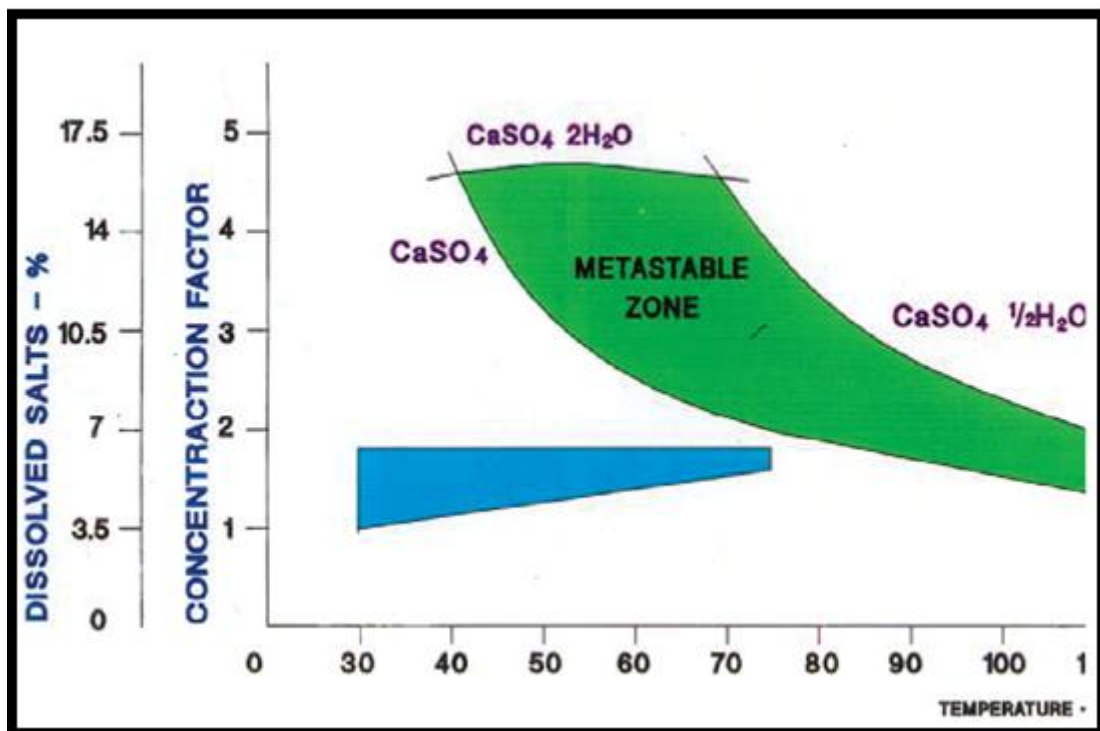


Figure 3.5. Brine concentration/temperature curve for LT-MED process.

Operation range of low-temperature distillation[8]

3.2 Process Description

The Zawia power plant has two MED , each evaporator has eight effect , the Process description of this plant will be done in next titles.

3.2.1 Desalination plant operating mode

The desalination plant is assembled and operated in the vicinity of the Zawia combine cycle power plant complex. The steam for desalination plant heating is provided from the auxiliary low pressure (LP) steam header of the power plant with the pressure range of (2.6 - 2.8) bar.[29]

3.2.2 Low pressure heating steam process

The MED desalination unit of a multi-effect evaporator consists of eight effects, one final condenser. In principle, MED utilizes the latent heat released by condensation of the vapor on tube side of each effect to boil the feed water. The vapor generated in one effect is passed to the next effect, where the same procedure is repeated. The multiple effect is achieved in a series of horizontal-tube, falling film type evaporative condensers. These serve as heat recovery section, wherein, through successive steps of evaporation and condensation at equilibrium vacuum conditions, a multiple amount of distillate is produced from a given quantity of low pressure steam. Also, in order to maintain the heat balance of the plant, there must be a heat input section and a rejection section. The first effect is the heat input section of the desalination system while the one integral final condensers serve as the heat rejection section.

The feed water is distributed in parallel (parallel feed concept) to the eight effects and evenly sprayed over the top rows of the horizontal tubes of each effect. The feed water descends as a thin film over the outside of tubes arranged horizontally. Vapor flows inside the tubes, where it condenses to form the distillate. As the vapor condenses inside the tubes, it heats up the feed water film on the outside of the tubes and causes the film to boil, thereby generating additional vapor which is introduced to the next effect via vapor duct. Each successive effect works at a lower saturation temperature than the one before. The condensation/evaporation process is repeated from the first to the eighth effect. With the partial vaporization of the feed water, its concentration is increasing toward the bottom of the effect. This concentrated feed water, called brine, flows to the next effect through a weir gate driven by the pressure difference between effects.

The distillate produced at each effect flows to the next effect in a same manner. As brine and distillate water is passing through the weir or U-loop seal, it is flashing down to the saturation pressure of the next effect, providing extra steam .

The brine accumulated in the eight effects is discharged as blow down. The distillate water accumulated at the (last) effect, where it flashes down to its final discharge temperature. The desalination plant operates under high vacuum condition. The vacuum is maintained by the venting system, which removes non-condensable gasses released from feed water in the effect

chambers and air in-leakage through various connections. Since the operation temperature is kept rather low temperature (below 69°C), the risk of corrosion of the plant components and the risk of scaling in the effect tube bundle is minimized. Anti-scale additive is dosed to the feed water to minimize the scaling at the outer surface of effect tubes.[29]

3.3 Definition of the physical system

3.3.1 General description of Zawia MED power Plant

The plant consists of two (2) MED desalination units (Evaporators) with a design capacity of 2,500 m³/day . as shown in (Figure 3.6) [29].



Figure 3.6 MED desalination units (Evaporators) of Zawia power plant

3.3.2 Feed Water System

The feed water system consists of one feed water control valve, two Seawater filters at the entrance of the feed water system, six Pre-heaters between the 8th~7th , 7th~6th , 6th~5th , 5th~4th ,4th~3rd and 3rd~2nd effects, eight feed distribution headers with spray nozzles on each effect and instrumentation. The seawater heated through the final condenser is partially used as feed water, while the remaining is discharged. The feed water flow shall be maintained at the constant value of 370 ton/hr regardless of the change of the plant production range within 50~100% and seasonal temperature variation. The feed water flow is equal to the distillate plus brine blow down flow. The feed water is heated up further through Pre-heaters and inter/after condenser using a part of the vapor generated in the effects or ejector discharge steam in order to give moderate temperature difference between the feed and the vapor. The feed water flow rate is controlled by the feed water control valve and flow transmitters installed at the inlet to each effect. The flow rate of feed water to each effect shall be tuned finely during the commissioning period by adjusting manual valves installed at the inlet to each effect. The initial setting of the flow to each effect will be based on the process flow diagram. It is very important to supply adequate and consistent quantity and quality of seawater on tubes of the effect. Also, even distribution of the feed water throughout the tubes is critical to the reliable operation of the MED units.[29]

3.3.3 Heat Rejection System

The heat rejection system, which consists of final condenser , a piping system, the necessary control valves and instrumentation, to facilitate a proper operation system, maintains a proper balance of heat in the MED unit. This is accomplished by exchanging latent heat between the vapor and the cold seawater in the final condensers. The seawater from the intake facility will be heated flowing through the final condensers successively. A portion of the heated seawater will be forwarded to the MED unit as a feed stream and the remainder will be returned to the open sea. The feed temperature needs be maintained at 37.0°C throughout the year. Therefore, the seawater flow rate will be adjusted automatically by throttling the cooling water control valve properly. The cooling water control valve will be modulated by the temperature of the feed water at the downstream of the final condenser as a input signal. as shown in (Figure 3.7),[29]

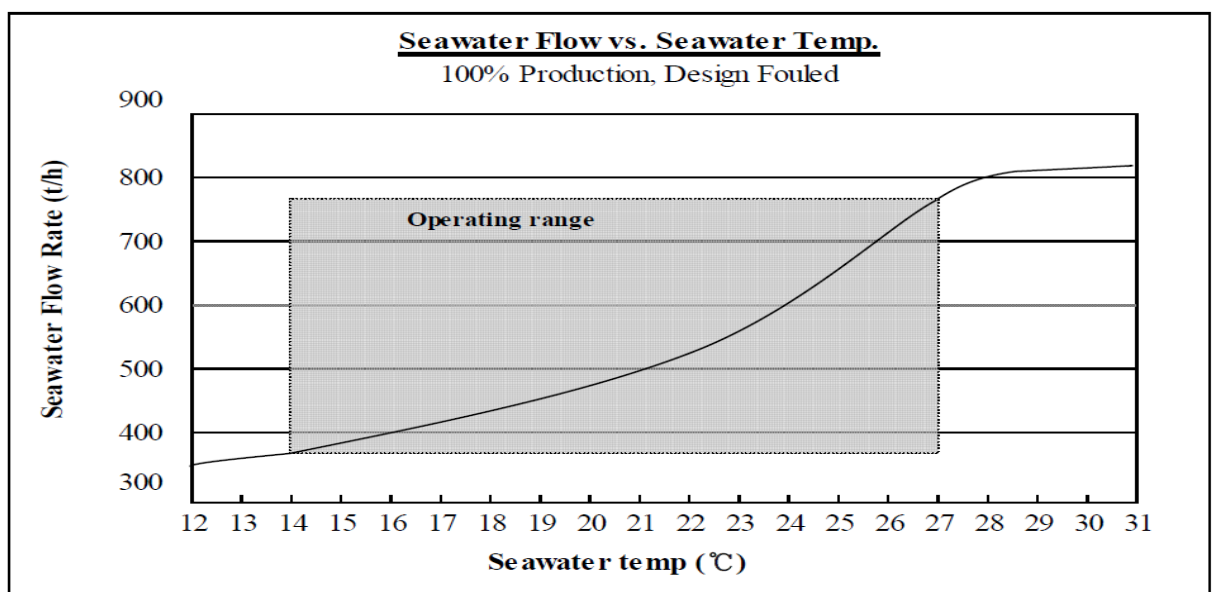


Figure 3.7 Feed Flow Seawater Temperature[29]

The final condenser is combined in one shell. A portion of the vapor in the last effect will be delivered to the final condenser, which is a shell and tube type heat exchanger. Seawater from the final condenser will be introduced to tube side of the final condenser to remove latent heat of the vapor. However, the final condenser is designed to handle a portion of the vapor from 8th effect.[29]

3.3.4 Brine Discharge System

The brine discharge system consists of two (2) brine discharge pumps, two (2) brine level control valve and other valves and instrumentation per each unit. The feed water evenly sprayed over the top rows of tubes flows down the tubes in a thin film absorbing heat. When it reaches at a saturation temperature, a part of the feed begins to evaporate. The remaining and concentrated feed water is collected at the bottom of the effect while the generated vapor flows to the next effect. This concentrated feed water at the bottom of effect is called “brine”. The brine is transferred to the brine chamber of the next effect through weir gates by the pressure difference between the two effects. By repeating this process, the brine is accumulated at the brine chamber of the 8th effect. The brine in the brine chamber of the 8th effect is pumped out by the brine discharge pump and disposed to the disposal system. Weir gates are to be fully submerged by the brine to prevent any vapor blow through. The height of the brine weir gate is adjusted properly during the commissioning period. When the brine enters

the next effect, the brine is cooled down the brine reaches at the new saturation temperature by giving off a portion of the brine through vapor that is called “flash-down”. The vapor produced by the flash is combined with the vapor generated at the tubes by the evaporation and then the combined vapor will be directed to the tube side of the next effect. The effect brine level transmitter is used for the brine level control. The output signal of the transmitter is received by the level controller, which is in turn responsively modulate the brine level control valve through pneumatic converter. [29]

3.3.5 Distillate System

The system consists of a piping system to transfer the distillate from the final condenser to the distillate water tank. The purpose of the distillate system is to transfer the produced distillate water from the final condenser to the distillate water tanks which is provided by Client. The vapor generated at the first effect flows to the second effect through the mist eliminator. The mist eliminator removes any brine particles entrained during the evaporation. The clean vapor is condensed by giving off their latent heat to the feed water through tubes and collected to the distillate chamber. The distillate produced at an effect is transferred to the next through the internal transfer piping. The temperature of the transferred distillate is higher than that of the next effect. Therefore, the distillate flashes down by the temperature difference between the two effects. In this

way, the distillate produced from the effects is collected to the distillate chamber of the eighth effect. Then, the collected distillate is delivered to the final condenser and further flash down makes the temperature of the distillate below 40°C. The flow rate corresponding to the production rate is automatically controlled by means of level controller of the distillate sump in the final condenser with related level control valve.[29]

3.3.6 Plant data

The desalination unit is designed to be in continuous steady operation at the production rate of between 50 up to 100% whereby [29] the operation parameters may vary in a wide range under various operation conditions.

Therefore the MED plant is made up of four main components which has to be modeled and shown in Figures (3.8,3.9,3.10,3.11,3.12,... respectively

- a) Evaporator unit(effects).
- b) Flash tanks .
- c) Pre-heaters
- d) Condenser unit

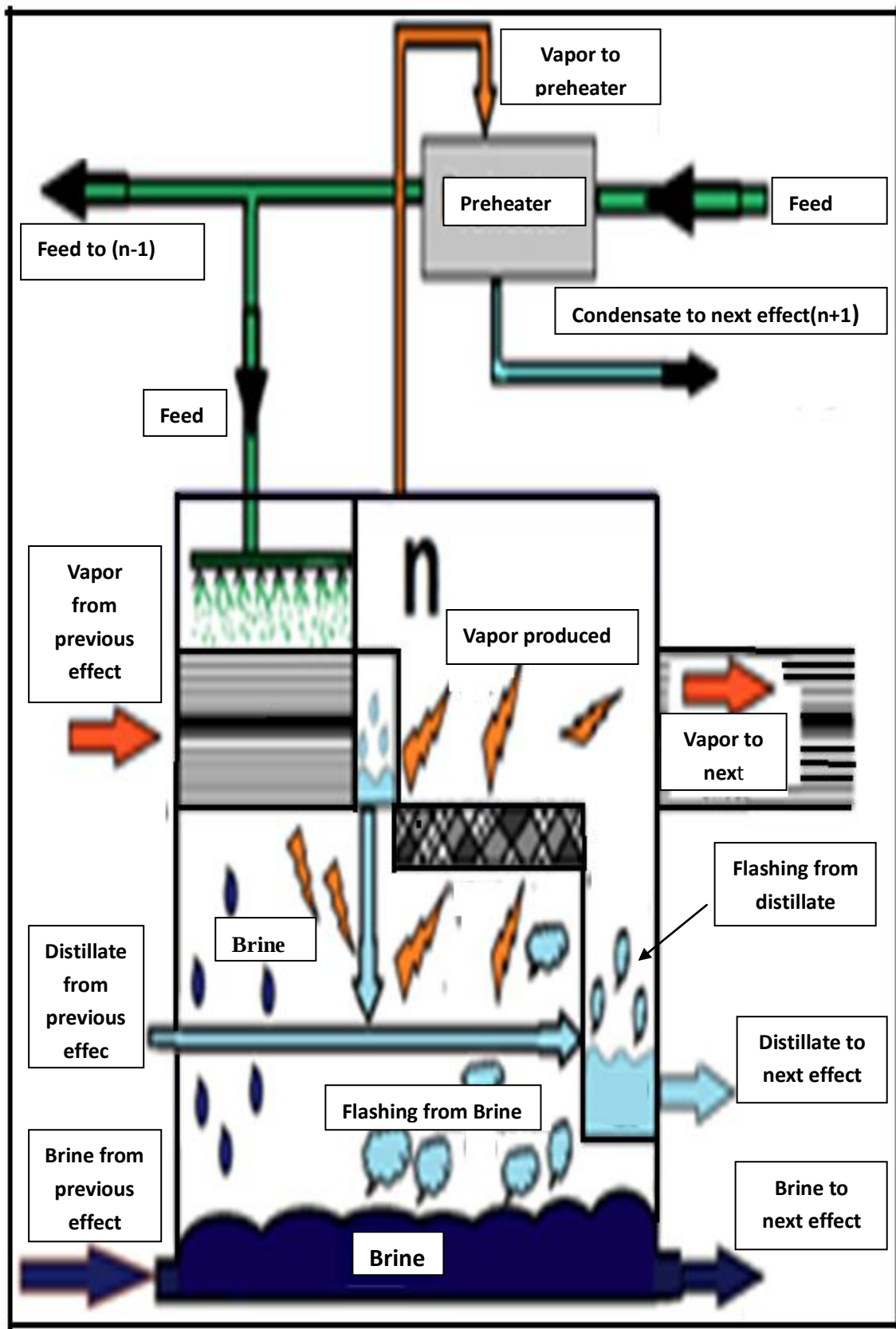


Figure 3.8 Schematic of a Single Effect

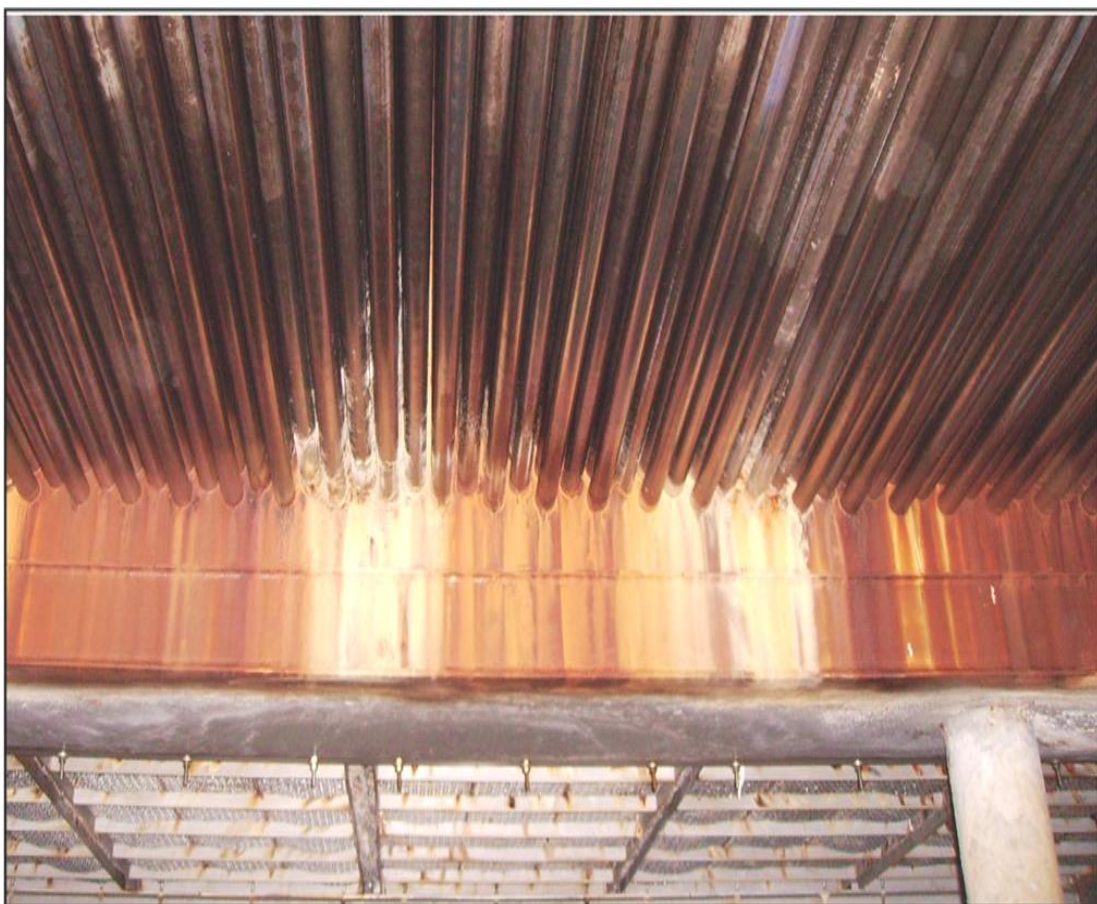


Figure3.9 The internal view of Effect in Zawia MED power Plant[29]

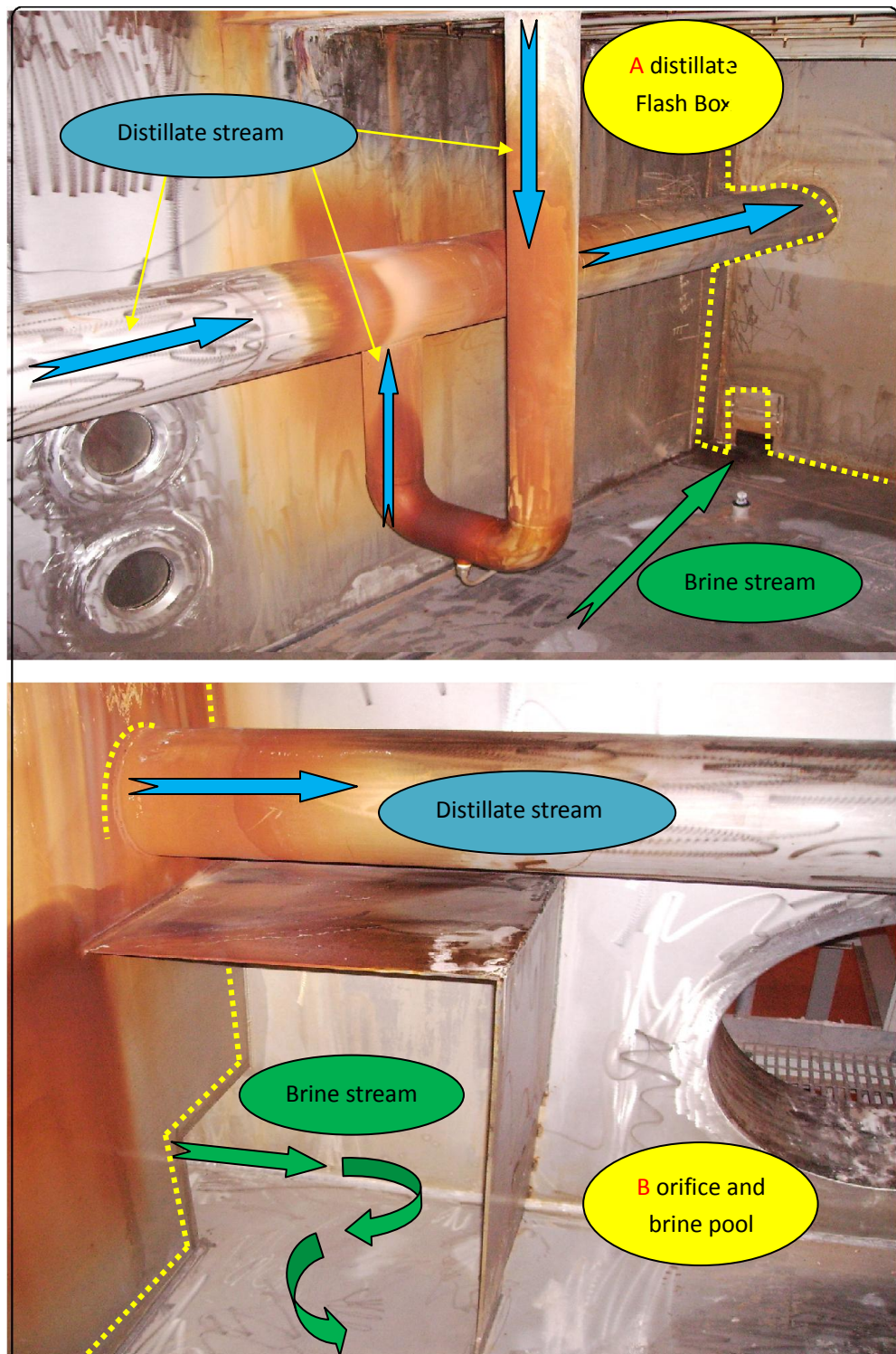


Figure3.10 In internal view (A and B)of the Flash box of effect[29]



Figure3.11 The pre-heaters of Zawia MED Power Plant[29]

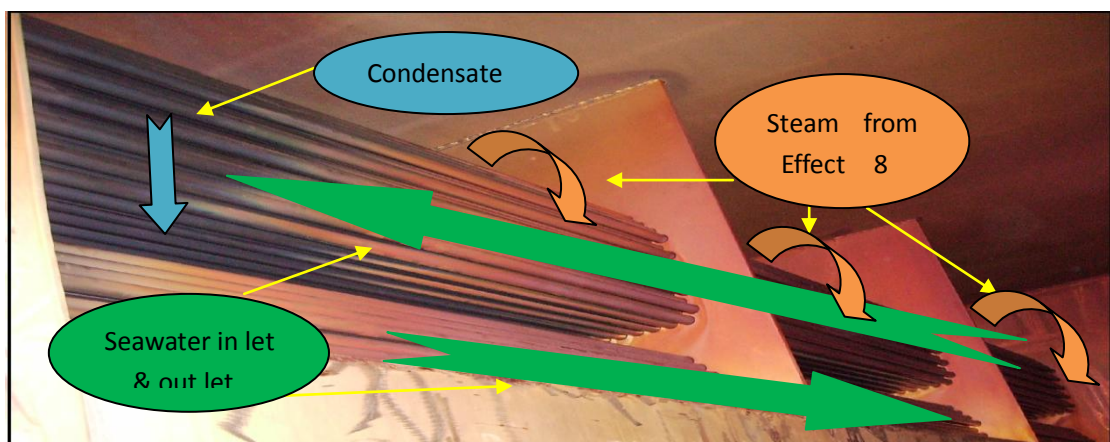


Figure3.12 The Condenser of Zawia MED Power Plant[29]

3.3.7 Desalination Unit Performance

Rated

- Net water production 2,500(m³/day)
- Distillate purity (ppm as TDS at 25°C) less than 4
- Design Seawater temperature (14- 27)°C

3.3.8 Main Process Data

The main process data relating to the above desalination unit performance are listed in Table(3.3)[29]

Table(3.3) The main Process Data

Steam supply	1.6 - 1.8 bar , saturation
Product water	104 t/h
Condensate return	13.6 t/h , 71°C from the first effect
Seawater supply	760 t/h , at 27°C
Seawater discharge	390 t/h , max 37°C
Brine discharge	266 t/h , 41°C and 54 g/kg TDS
Brine concentration ratio	1.47

3.3.9 Technical data

Table 3.4 The technical data [29]

Effects	
Item	Data/Unit
Distillate output	2,500 m ³ /day
Maximum temperature	65°C seawater temperature of 27°C
No. of effects	Eight (8)
No. of tubes per effect	5,040
Tube size × thickness	25.4 mm OD × 0.4 mm th'k
Tube material	(effect 1 to 4) Titanium
Tube length	2,100 mm
Final Condenser	
No. of condenser	One (1)
No. of tubes/condenser	3,344
No. of tubes/path	418
Tube size × thickness	19.05 mm OD × 0.4 mm th'k
Tube material (stages 1 to 4)	Titanium
Tube length	3,200 mm
Operating pressure of final Condenser	
Shell side	0.1 bar
Tube side	4.5 bar
Operating temperature	40 °C

3.4 Water salinity analysis

The real salinity of intake sea water ,distillate produced and reject brine of the plant are analysed using portable TDS meter. Figure (3.13) this device measure the salinity in gram per litter (g/l) and conductivity (μS). The intake seawater salinity is 37.3 g/l , salinity of distillate produced is 2.8 g/l , reject brine is 49.8 g/l . shows Sension7 Salinity Device [29]



Figure3.13 The Sension7 Salinity Device[29]

CHAPTER FOUR

MODEL EQUATIONS

4.1 Introduction

Modeling refers to formulating a set of equations that describe mathematically an industrial process under consideration. In the simulation phase, the formulated model is solved by using a suitable solution procedure, as well as by entering the values of independent process variables. This typically done with the aid of a computer, which is termed as computer-aided simulation. The goals of modeling and simulation in the process industry include improving and optimizing designs, developing better insight into the process, and ultimately leading to the optimal operation and control of the process.

In simulating a plant, represent all the individual units in the process should be presented separately, and couple them together in accordance with physical interconnections to give the model for the whole plant. The basic approach is to utilize mathematical relationships among process variables. These relationships define the physical laws, material, energy and momentum balances, etc. in the form of a set of generally nonlinear, algebraic and differential equations. For the preliminary checking of material and heat balances of a plant, a simplified model may be sufficient. For in-depth studies, rigorous models will be required.

According to the time-dependence, there are two kinds of process models: steady-state and dynamic models. The steady-state model is time-invariant. It describes the process through a set of algebraic equations. Design models of continuous processes are always steady-state models. Different steady-state models are used for design purposes as well as for parametric studies of existing plants for performance evaluation and operational optimization.

Dynamic model are time-dependent, and they contain differential equations and supporting algebraic equations. The model of the process must be connected to the model of the control system to accomplish the simulation of the whole process. In this study, a steady state model has been developed to study the performance and operational optimization of Zawia MED Power plant to produce distillate water for the power plant.

4.2 Model formulation

When developing a mathematical model it is necessary to make a compromise between being too simplistic to obtain good results and too complex to obtain an exact solution. for this reason some assumption had to be made but these have been kept to a minimum.

The assumption made in modelling the MED plant are the common assumptions which are already justified to be used in thermal desalination plants

4.2.1 The main assumption

- The vapour and the brine are at saturation temperature .
- No condensate holds up in the tube side .
- All vapour and liquid phases are mixed .
- Each affect has the same mass of feed flow .
- All the steam entering the tube side will condense at saturation temperature
- Constant specific heat.
- Product leaving any effect is salt free .
- Heat losses from evaporator (effect) surfaces are negligible.
- Constant heat transfer area in all evaporators (effects).
- Shell side temperature is equal to brine outlet temperature .

4.2.2 Evaporator model

Based on Figure 4.1 material and energy balance can be stated as .

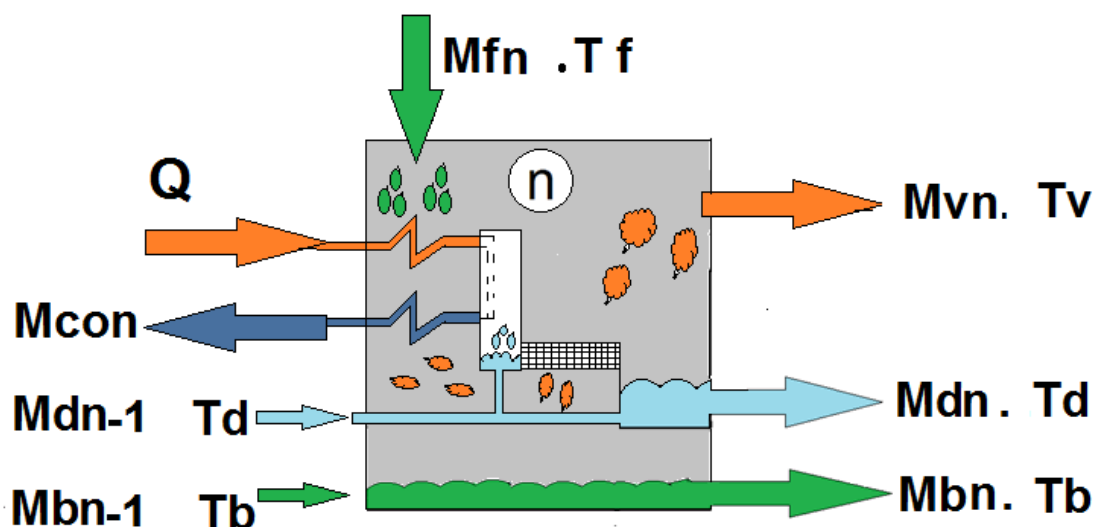


Figure 4.1 Single Effect

Considering the shell side :

Mass Balance:

Rate of change of mass = $\sum$ mass rates inlet - $\sum$ mass rates outlet

$$\frac{dM(n)}{dt} = M_f(n) - M_v(n) - M_b(n) \quad (4.1)$$

Material Balance:

The rate of change of salt content equals the rate of change of feed salt content minus rate of change of brine salt content.

$$\frac{dM X_b(n)}{dt} = M_f(n) * X_f(n) - M_b(n) * X_b(n) \quad (4.2)$$

For steady state the amount of distillate produced in a plant with n stages(M_{dn}) is the sum of the distillates of the individual stages

$$M_{dn} = M_{d1} + M_{d2} + \dots \quad (4.3)$$

$$M_v(n) = M_f(n) - M_b(n) \quad (4.4)$$

Brine concentration factor, the maximum salt content in the brine is determined by the solubility of the salts. In order to avoid salt precipitation leading to encrustations and blockages, the concentration in the brine must be limited. The concentration factor (CF) being defined as the ratio of brine concentration(X_b) to feed concentration (X_f) is usually used as the parameter which gives the limitation of the evaporation, i.e. concentration process :

$$CF = \frac{X_b}{X_f} = \frac{M_f}{M_b} \quad (4.5)$$

It may immediately be derived from the salt balance that the concentration factor can also be interpreted as the ratio of the feed mass flow(M_f) to the brine mass flow(M_b). As the capacity of a seawater desalination plant is defined by the distillate production, it is helpful for our analysis to introduce the yield or recovery rate Φ . Starting with the definition of the concentration factor (CF), (Eq 4.5) the recovery rate (Φ) can be expressed as :

$$CF = \frac{M_f}{M_b} = \frac{M_f}{M_f - M_d} = \frac{1}{1 - \frac{M_d}{M_f}} = \frac{1}{1 - \Phi} \quad (4.6)$$

$$\Phi = \frac{CF - 1}{CF} \quad (4.7)$$

For steady state

Salt produced by each effect can be written by (Eq.4.8)

$$X_{be(n)} = \frac{(M_f(n) * X_f(n))}{M_b(n)} = \frac{(M_f(n) * X_f(n))}{(M_f(n) - M_v(n))} \quad (4.8)$$

Considering the distillate as salt free ,the mass of the salt in the feed and the brine streams are equal.

For $X_b(n) \dots$ from $(n) = 2$ to 8

$$X_b(n) = \left[\frac{M_b(n-1) * X_b(n-1) + M_{be(n)} * X_{be(n)}}{M_b(n)} \right] \quad (4.9)$$

Energy Balance :

Rate of heat change = $\sum$ heat added to the system $-\sum$ heat exiting from the system

The heat added to the evaporator is the latent heat of condensing vapour in the tube plus the total enthalpy of the feed , while the heat exiting the evaporator is the enthalpy of the generated vapour plus brine enthalpy .

$$[\sum M_s * C_p] \frac{dT_b(n)}{dt} = Q(n) + [M_f(n) * C_p * T_f(n)] - [M_v(n) * \lambda_v(n)] - [M_b(n) * C_p * T_b(n)] \quad (4.10)$$

For steady state

The heat added to the evaporator equal to the heat taken by the generated vapour plus the heat carried by the brine minus the heat content of the feed .

$$Q(n) = M_v(n) * \lambda_v(n) + M_b(n) * C_p * T_b(n) - M_f(n) * C_p * T_f(n) \quad (4.11)$$

Considering the tube side as our control volume .

Mass Balance:

$$M_v - M_d = 0 \quad \text{or} \quad M_v = M_d \quad \text{As assumption} \quad (4.12)$$

Energy Balance :

Heat added to tube side = Latent heat of condensation of the total vapour entering the tube side, which are the rate of evaporation in the previous effect plus the vapour generated by flashing from distillate and brine streams ,but the vapour generated by flashing from brine start from effect number 2 until 8 and the vapour generated by flashing from distillate start from effect number 3 until the condenser .

$$Q(n) = [M_d(n-1) + M_{df}(n-1) + M_{bf}(n-1)] * \lambda(n-1) \quad (4.13)$$

For the first effect($n = 1$) the total heat added is the latent heat of the supplied steam.

$$Q(1)=M_s*\lambda_s = M_f(n)*C_p*[T_b(n) - T_f(n)] + M_v(n)*\lambda_v(n) \quad (4.14)$$

The mass of brine produced in effect 1 :

$$M_b(1)=M_f(n) - M_v(n) \quad (4.15)$$

For second effect($n=2$) the heat added is the latent heat of condensing vapour from previous effect($n-1$).

$$Q(2)=M_f(n)*C_p*[T_b(n)-T_f(n)] + M_v(n)*\lambda_v(n) \quad (4.16)$$

$$Q(2)=M_v(n-1)*\lambda_v(n-1) \quad (4.17)$$

The mass of brine and distillate produced in effect 2 :

$$M_d(2)_t=M_v(n-1) \quad (4.18)$$

$$M_b(2)_t=M_b(n-1)-M_{bf}(n) + M_f(n) - M_v(n) \quad (4.19)$$

For third effect ($n = 3$) the heat added is the latent heat of condensing vapour from previous effect($n-1$) plus the vapour generated by brine flashing .

$$Q(3)=M_f(n)*C_p*[T_b(n)-T_f(n)] + M_v(n)*\lambda_v(n) \quad (4.20)$$

$$Q(3)=[M_v(n-1)-M_{vh}(n-2)]*\lambda_v(n-1)+M_{bf}(n-1)*\lambda_b(n-1) \quad (4.21)$$

The mass of brine and distillate produced in effect 3 :

$$M_d(3)_t=M_d(n-1)+[[M_v(n-1)-M_{vh}(n-2)]+M_{bf}(n-1)]-M_{df}(n) \quad (4.22)$$

$$M_b(3)_t=M_b(n-1)-M_{bf}(n) + M_f(n) - M_v(n) \quad (4.23)$$

For effect (n=4) the heat added is the latent heat of condensing vapour from previous effect (n-1) plus the vapour generated by brine flashing and distillate flashing .

$$Q(4) = M_f(n) * C_p * [T_b(n) - T_f(n)] + M_v(n) * \lambda_v(n) \quad (4.24)$$

$$Q(4) = [M_v(n-1) - M_{vh}(n-2)] * \lambda_v(n-1) \\ + [M_{bf}(n-1) + M_{df}(n-1)] * \lambda_b(n-1) \dots \quad (4.25)$$

In general equation for effects from (n= 4 to 8):

$$Q(n) = M_f(n) * C_p * [T_b(n) - T_f(n)] + M_v(n) * \lambda_v(n) \quad (4.26)$$

$$Q(n) = [M_v(n-1) - M_{vh}(n-2)] * \lambda_v(n-1) \\ + [M_{bf}(n-1) + M_{df}(n-1)] * \lambda_b(n-1) \quad (4.27)$$

In general equation for effects from (n= 4 to 8) the mass of brine and distillate produced are :

$$M_d(n)_t = M_d(n-1) + [[M_v(n-1) - M_{vh}(n-2)] + [M_{bf}(n-1) + M_{df}(n-1)]] \\ - M_{df}(n) \quad (4.28)$$

$$M_b(n)_t = M_b(n-1) - M_{bf}(n) + M_f(n) - M_v(n) \quad (4.29)$$

Heat transfer across tube wall in effects:

$$Q(n) = U(n) * A * [T_b(n-1) - T_b(n)] \quad (4.30)$$

4.2.3 Pre-heater model

A sea water pre-heater is a conventional shell-and-tubes heat exchanger, in the tubes flow the make-up of all the cells hotter than the one where the pre-heater is installed, including the make-up of the cell itself. In the shell condenses part of the steam that is produced in the cell where it is installed. when heating the make-up in pre-heaters, the necessary heat is taken from cells by mass $M_{vh}(n)$ at lower temperature and the total production rate is increased with the same heat input to the evaporator. The Pre-heater schematic diagram shown in Figure 4.2

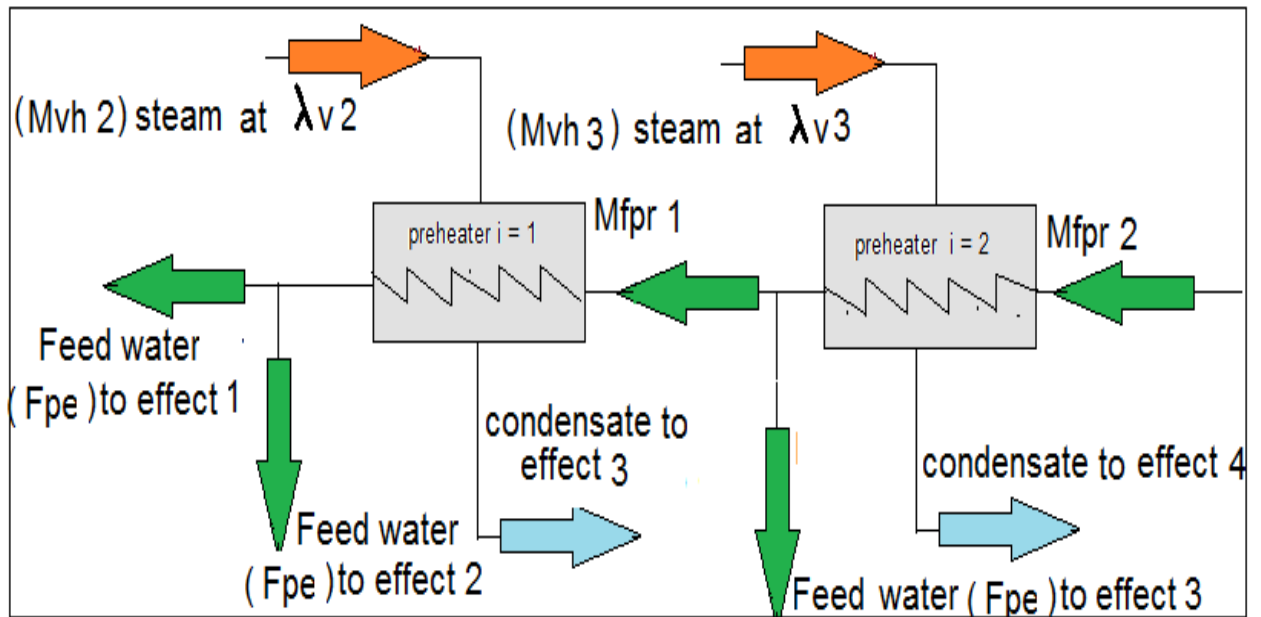


Figure 4.2 Pre-heaters diagram

Feed of each effect(F_{pe}) = Total feed(M_{ft}) per number of unit (n)

$$F_{pe} = \frac{M_{ft}}{n} \quad (4.31)$$

The mass of seawater inter to preheater =

$$\text{Total feed} - (\text{number of effect} - 2) * \text{Feed of effect}$$

$$M_{fpr(i)} = M_{ft} - (n - 2)F_{pe} \quad (4.32)$$

Where $(n= 3 \text{ to } n=8)$

Where $i = 7 \text{ to } 1$ i decrease from 7 to 1

$$M_{vh(n-2)} = \left[\frac{[M_{fpr}(n - 2) * C_p [T_f(n - 1) - T_f(n)]]}{[\lambda_v(n-1)]} \right] \quad (4.33)$$

4.2.4 Flash tank model

In order to demonstrate this we have to follow the path of both of these flows through the plant. In a multiple-effect plant, the brine and the distillate are fed from stage to stage. Both brine and distillate are leaving the stage as “boiling water”. The boiling temperature and the boiling pressure are determined by the conditions in each stage. To transfer the heat in the condenser/evaporator unit a driving temperature difference must be present. This means that both brine and distillate enter a chamber in which a lower pressure and a lower boiling point temperature exist. The only possibility to release this surplus energy present under these conditions is through spontaneous vapour production. This process is known as “flash evaporation” or “flashing”. The vapour produced by flashing is added to the vapour produced by evaporation, so that the total distillate produced in a stage comprises some parts as shown in Equations (4.21 ,4.25)

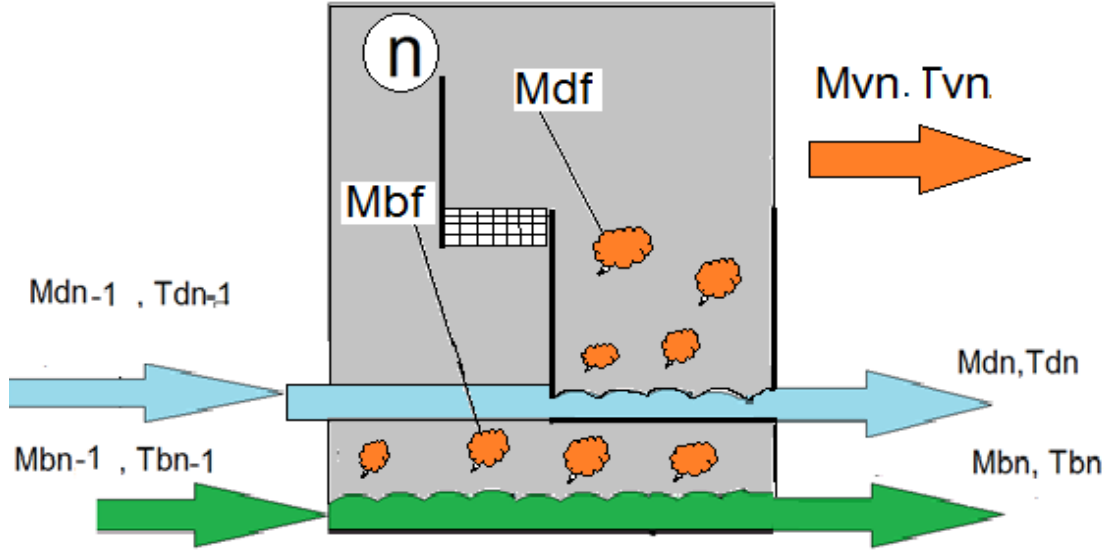


Figure 4.3 Flash tank schematic diagram

The heat content of the vapour formed by flashing the incoming stream

$M(b,d) f$ = the amount of heat lost from the stream

The brine flashing(Mbf)starting from effect ($n=2$ to $n=8$)

$$Mbf(n) = \left[\frac{M_b(n-1) \cdot C_p \cdot (T_b(n-1) - T_b(n))}{\lambda_b(n)} \right] \quad (4.34)$$

The distillate flashing(Mdf) starting from ($n=3$ to $n= 8$)

$$Mdf(n) = \left[\frac{M_d(n-1) \cdot C_p \cdot (T_b(n-1) - T_b(n))}{\lambda_d(n)} \right] \quad (4.35)$$

4.2.5 Condenser model

the main parameter in condenser is the temperature of the exiting sea water(T_{sw}) from condenser ,which will be the feed temperature of last effect taking the energy balance on the control volume of Figure 4.4

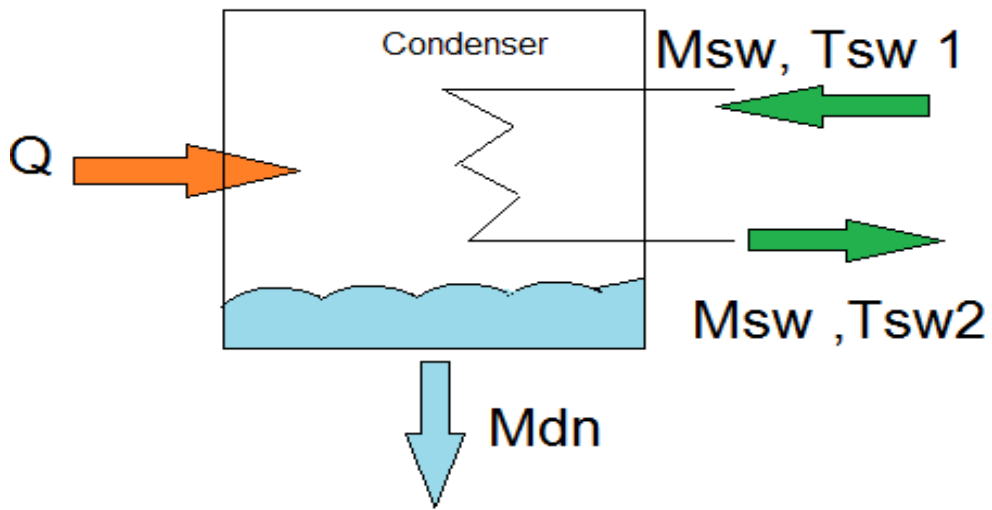


Figure 4.4 Condenser schematic diagram

The rate of heat change = $\sum$ heat added to the condenser - $\sum$ heat exiting the condenser

For steady state

$$Q_c = M_{cw} * C_p * (T_{sw2} - T_{sw1}) \quad (4.36)$$

$$Q_c = M_{cw} * C_p * (DTC) \quad (4.37)$$

In the same time the heat added to the condenser is the latent heat of condensation of the total vapour entering the condenser , which is the vapour generated in the last effect (n=8) plus the vapour generated by flashing the distillate and brine streams in the last flash tank of effect eight.

$$Q_c = [M_v(n) + M_{df}(n) + M_{bf}(n)] * \lambda_v(n) \quad (4.38)$$

$$Q_c = A_c * U_c * (LMTD)_c \quad (4.39)$$

$$(LMTD)_c = \frac{T_f - T_{sw}}{\ln \left[\frac{[T_v(n) - T_{sw}]}{[T_v(n) - T_f]} \right]} \quad (4.40)$$

4.3 The Gain Output Ratio

The gain output ratio is defined as the mass of fresh water produced from the plant divided by the mass of steam used to produce this fresh water. It is used when steam is available as a heat source, it can be defined mathematically:

$$GOR = \left[\frac{[M_d(n)]}{[M_s]} \right] \quad (4.41)$$

It is an important parameter since it is an indication of the efficiency of the plant, in other words the GOR indicates how effectively heat is utilized in the secondary circuit (MED) plant to produce a certain amount of product water.

4.4 Physical properties

In order to proceed with calculation the physical properties of the system as to be defined as a correlation function of temperature salinity for brine, overall heat transfer coefficient is predicted using a correlation function [30]. Details of the correlations used are presented in appendix (A).

4.4.1 Steady state plant analysis algorithm

To analyze an existing plant using the developed model, the input data required are; sea water temperature (T_{sw}), salinity (X_f), available steam temperature (T_s), number of effects (N), condenser range (DT_c), area of effect (A_e), area of condenser (A_c), mass of steam (M_s). The output will be the distillate product (M_{dn}), temperature distribution (T_v, T_b), heat load for each effect (Q), total rejected brine and Performance ratio (PR). or. (GOR).

4.4.2 Calculation procedure

The calculation procedures can be summarized in flow chart shown in Figure(4.5).The calculation procedures, can be pointed as:

1. The solution starts by assuming a condenser range (DT_c) to find temperature of last effect .
2. Calculate the feed flow of each effect .
3. Using the average inter effect temperature difference calculate brine temperature distribution.
4. Calculate feed flow temperature of each effect .
5. Calculate boiling point elevation (BPE) of each effect .
6. Calculate vapour temperature distribution of each effect .
7. Using the correlation to find the latent heat .
8. From energy balance calculate the mass of steam produced from effect one and two .
9. Calculate the mass of vapour for pre-heater number one.
- 10.Using the correlation to find the specific heat .
- 11.From salt balance calculate the salinity of each effect .
12. From energy balance calculate the mass of steam produced from effect (start from 3to 8)and calculate the mass of vapour for pre-heater which it takes from effects number(2 to 6).
- 13.Calculate the distillate produced of each effect and Calculate the brine rejected .

14. Solve energy balance to find heat load by the equation

$(Q = U \cdot A \cdot \Delta T)$ and calculate the distillate produced for each effect .

15. Iterate on the mass of distillate produced until converge to constant value.

16. Find condenser heat load . calculate the condenser temperature range if it is different from the first guess within the given tolerance , calculate the Gain Output Ratio (GOR) .

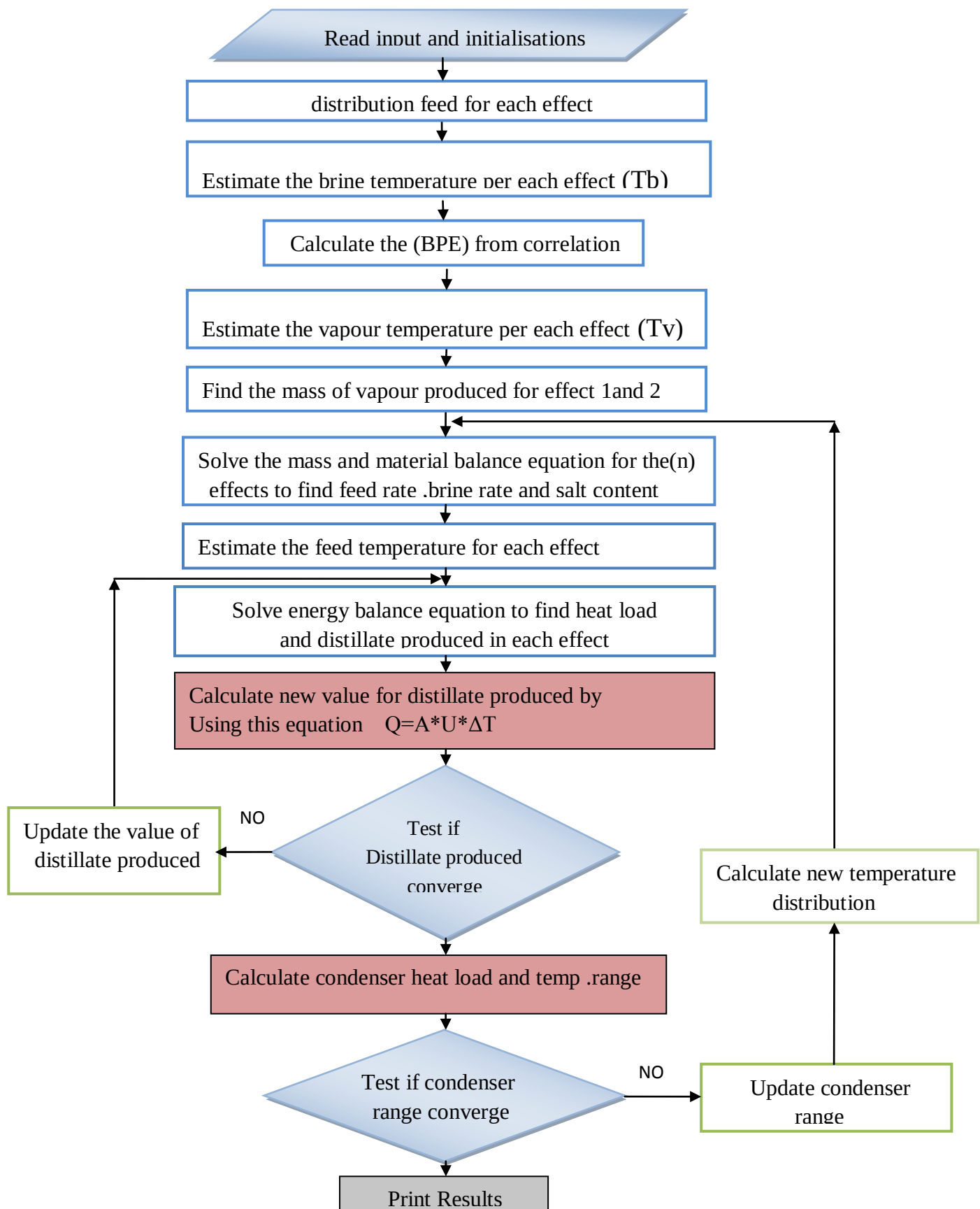


Figure 4.5 Calculation procedure of plant analysis.

4.5 ZAWIA MED Power Plant Simulator

4.5.1 DESALINATION UNIT SIMULATOR (DUS)

The program was written by visual studio 2010 ,Figure 4.6 simulator view as appear in computer screen, the elevation of the simulator has view of MED of Zawia power plant ,and the icon for RUN the simulator ,icon of input date ,and icon of result as DTC($^{\circ}\text{C}$) ,GOR, Mdcon [mass of distillate from condenser] (kg/s), and icon for results of each cell .

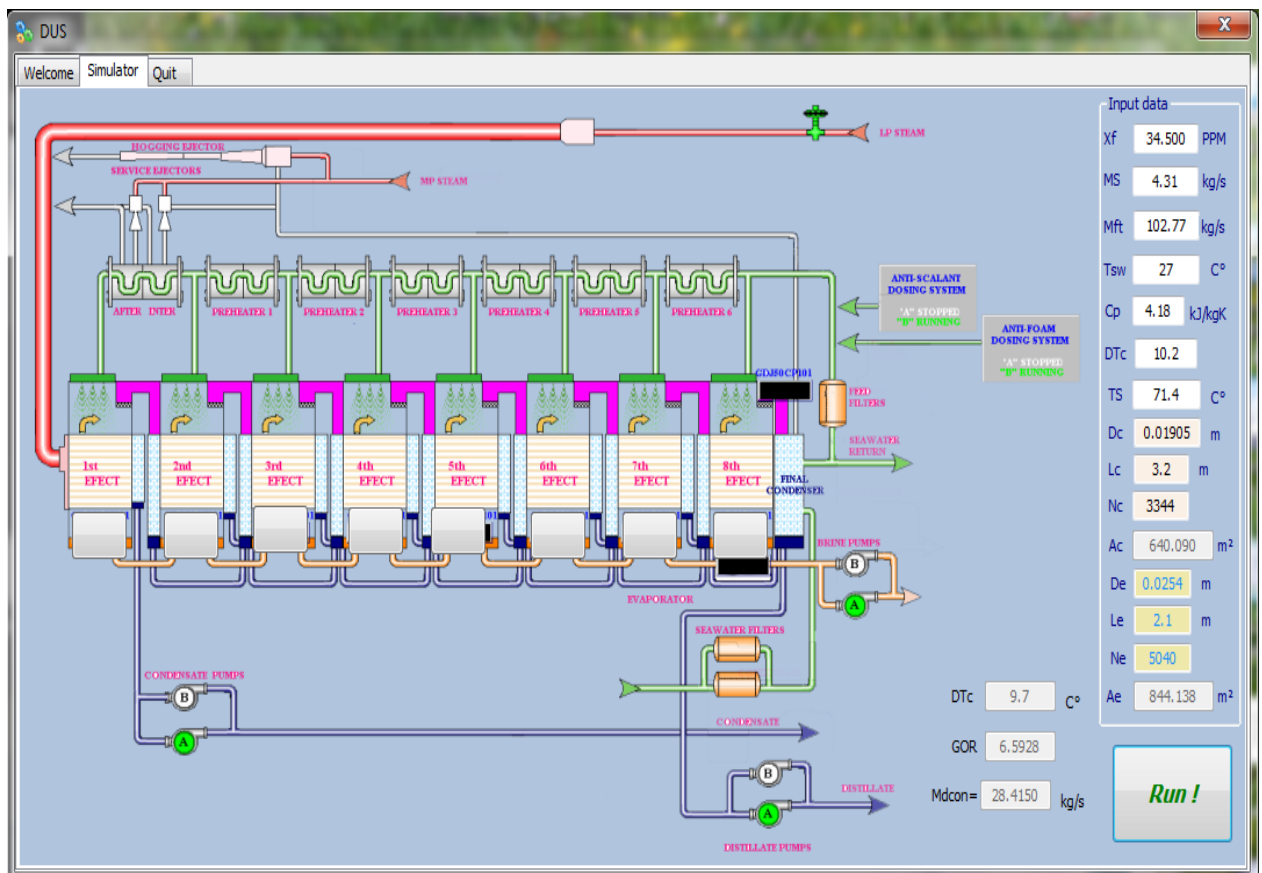
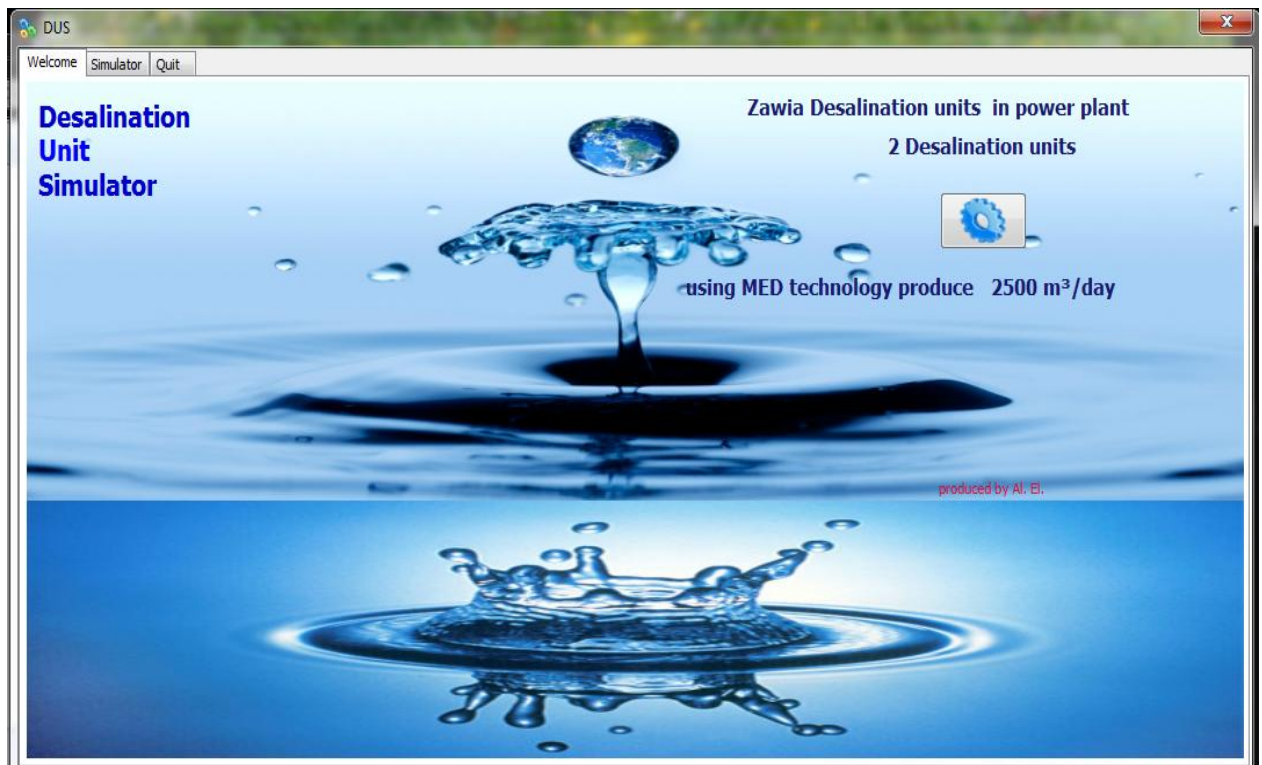


Figure 4.6 simulator view of MED in Zawia power plant as appear in computer screen

CHAPTER FIVE

RESULTS AND DISCUSSION

5.1 Validation of the model

The results of the developed model described in chapter 4 have been compared with design and real data.

The model is based on a one region where the effects was treated as a region contains saturated mixture of liquid and vapor. It was applied to a parallel feed system MED plant , the equation derived in the one region effect model has been programmed in the visual basic studio (2010).The program has one option ,which is used to analyse an existing plant. The details of the program input and output are given in Appendix(B).

The validation procedure proves that the model simulates accurately and consistently the plant. However, some discrepancies between results and expected responses can be observed.

The variation on the total distillate produced in Winter was 10.53% less. while the variation on the total distillate produced in Summer was 6.36% less, the variation on the total distillate produced which it take from design data was 1.672% less.

This reduction of distillate produced between Winter and summer refer to the reason which was a long time between the data ,the Figures (5.1, 5.2, 5.3, 5.4) shows the operation data in Summer and Winter of Zawia MED.

Total distillate produced in Summer and Winter are give good results

compared with the data obtained from design data. In the Table: 5.1 give the Comparison of results from the model with design data and real data in Winter and Summer, and in Table 5.2 give the error analysis between obtained results and design data in Winter and Summer.

The programme results in Summer and Winter are shown in Apendexe (C)

Table5.1 Comparison of results from the model with design data and real data in Winter and Summer.

Parameter	Design data	The model	Real data In Winter	Real data In Summer
Gain Output Ratio	6.71	6.5	6.429	6.615
Total distillate produced (kg/s)	28.888	28.405	25.846	27.05

Table 5.2 Error analysis between obtained results and design data in Winter and Summer.

Design data	Gain Output Ratio	Total distillate produced (kg/s)
	3.129 %	1.672 %
Real data in Winter	4.18%	10.53%
Real data in Summer	1.45 %	6.36 %

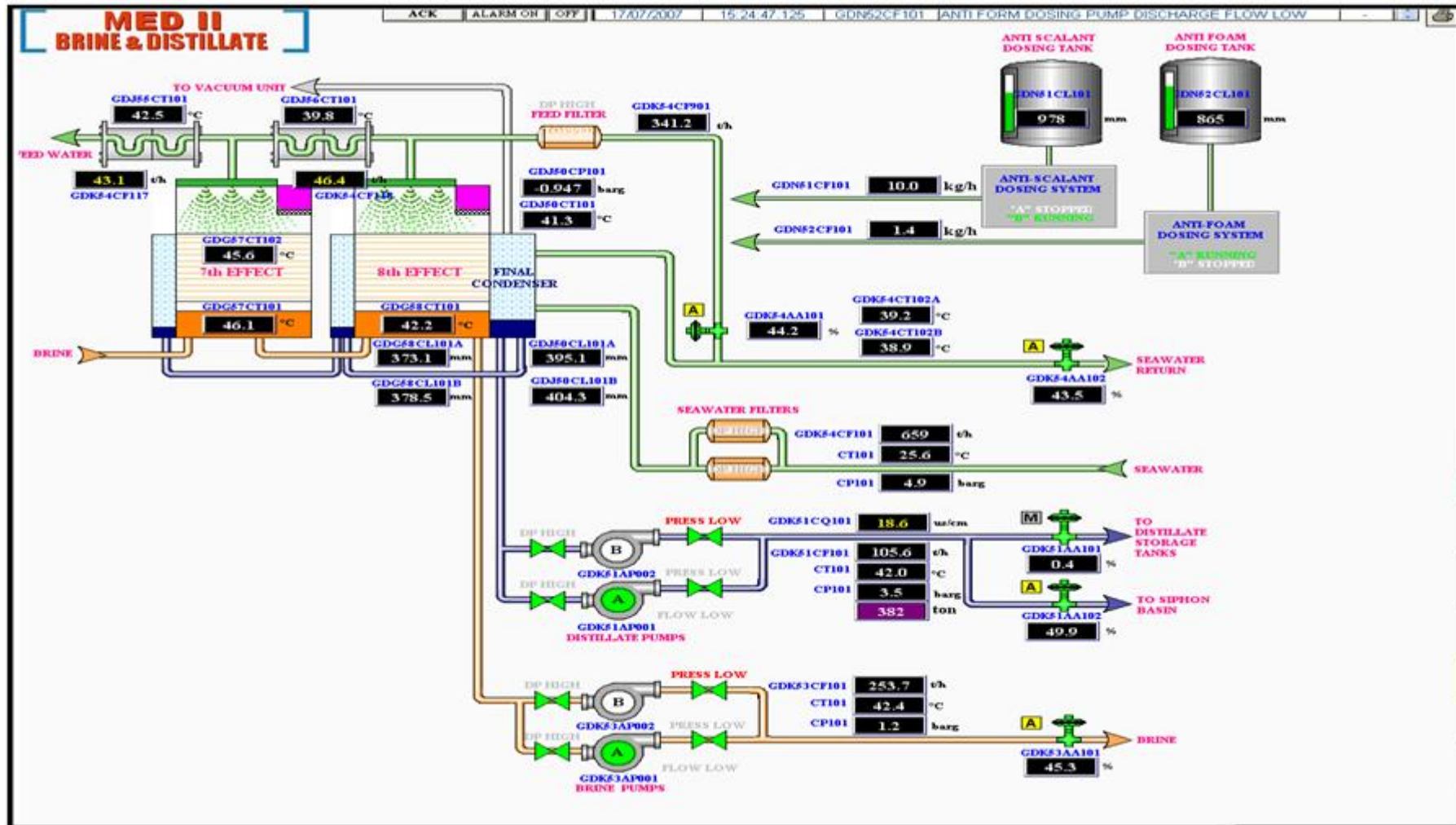


Figure 5.1 Bart 1 The Operation data from computer screen of Zawia MED Power plant in Summer at temperature (25.6 °C)

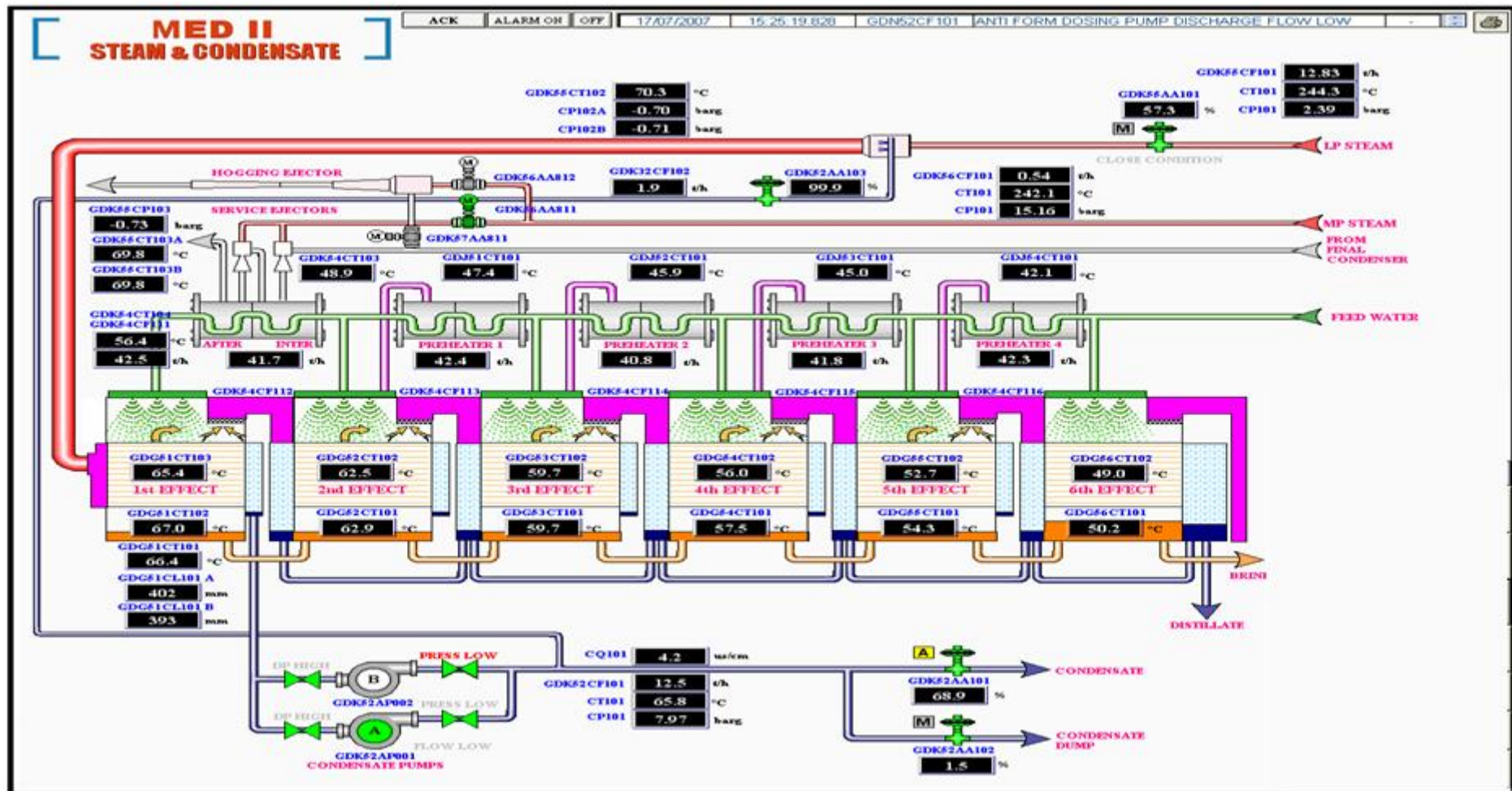


Figure 5.2 Bart 2 The Operation data from computer screen of Zawia MED Power plant in Summer at temperature (25.6°C)

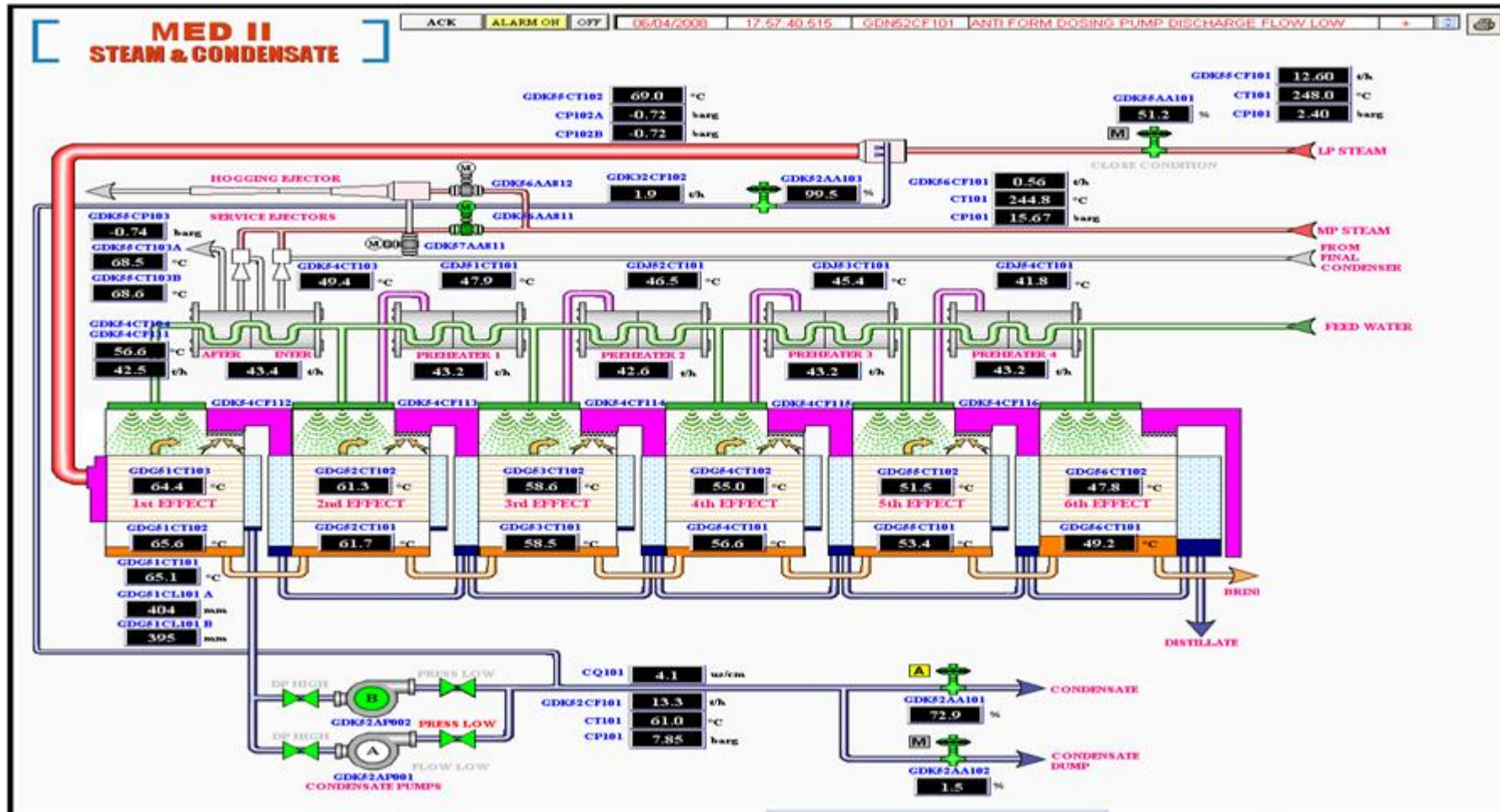


Figure 5.4 Bart 2 The Operation data from computer screen of Zawia MED Power plant in Winter at temperature (18.3°C)

5.1.1 The comparison of vapour and brine temperature in each effects

Figure 5.5 show the relationship between vapour temperature (T_v) and the number of effects for real operating data of Zawia MED power plant .Table 5.3 show the error analysis (%)for (T_v) for winter and summer season. While Figure 5.6 show the brine temperature(T_b) for each effect in both winter and summer .The comparison between the real and simulated temperature (T_m) data shown in both Figures. Obtained (T_m) shows that there is find good agreement real and calculated temperature .Table 5.4 shows the error analysis (%) for (T_b) for winter and summer.

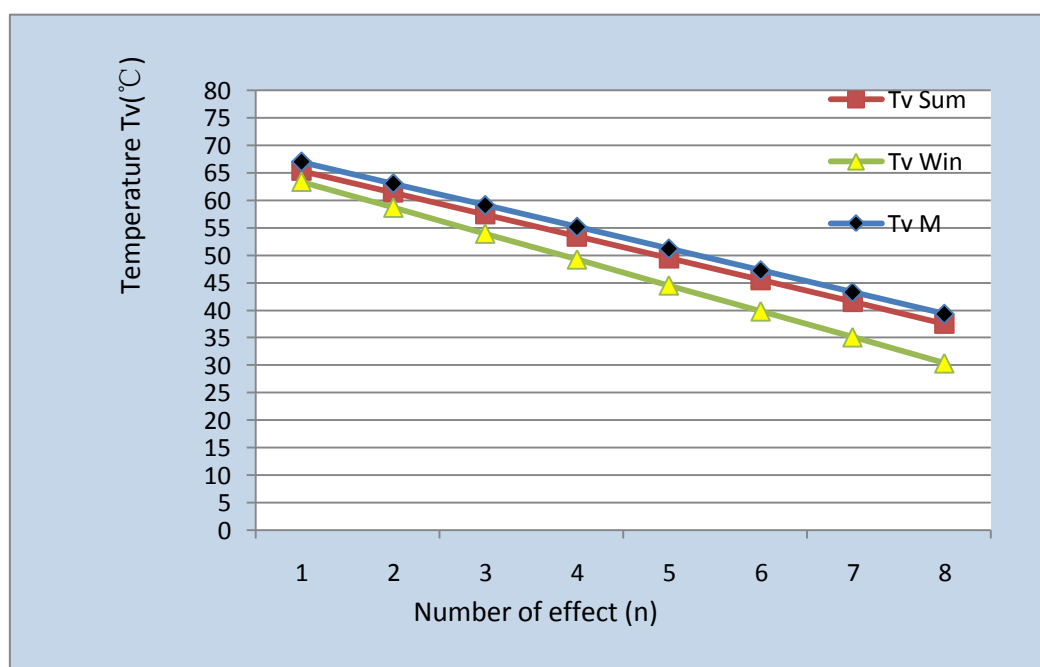


Figure 5.5 Comparison of vapour temperature in each effects with model results

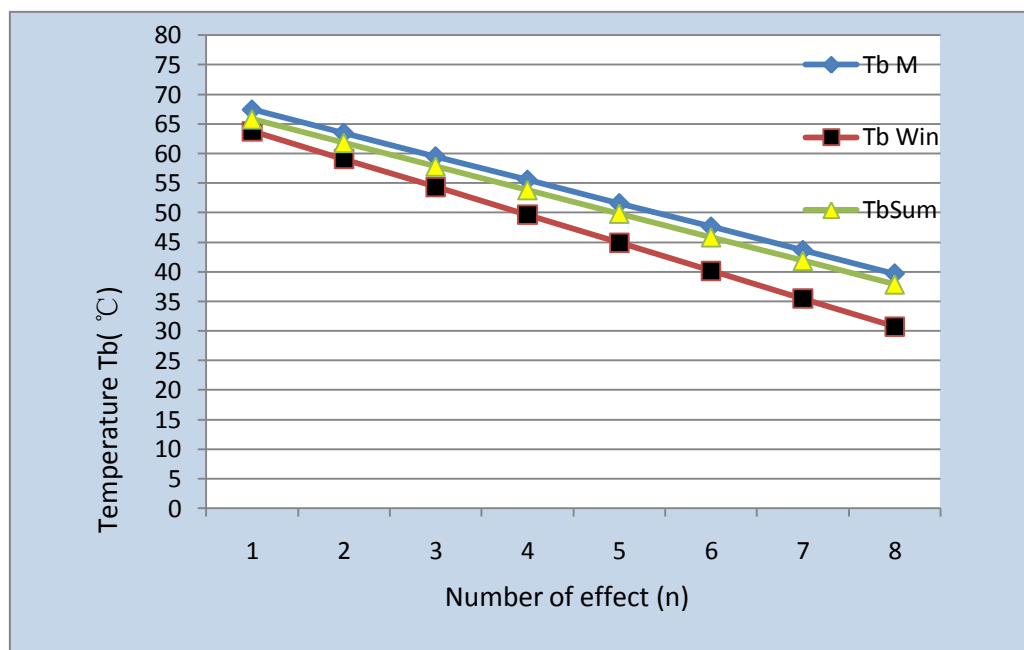


Figure 5.6 Comparison of brine temperature in each effects with model results

Table 5.3 Comparison the vapour temperature(T_v) model with real data in Winter and Summer

Number of Effect	T_v ,Model	T_v ,in Winter	Error%	T_v ,in Summer	Error%
1	67.022	63.37	5.44%	65.39	2.43%
2	63.074	58.66	6.99%	61.42	2.62%
3	59.125	53.95	8.19%	57.44	2.85%
4	55.177	49.25	10.74%	53.46	3.11%
5	51.228	44.54	13.05%	49.49	3.39%
6	47.278	39.83	15.75%	45.51	3.73%
7	43.329	35.13	18.92%	41.53	4.15%
8	39.379	30.40	22.80%	37.55	4.64%

Table 5.4 Comparison the brine temperature(T_b) model with real data in Winter and Summer

Number of Effect	T_b , Model	T_b ,in Winter	Error%	T_b ,in Summer	Error%
1	67.44	63.77	5.44%	65.81	2.41%
2	63.40	59.05	6.86%	61.82	2.52%
3	59.52	54.53	8.38%	57.83	2.83%
4	55.56	49.62	10.6%	53.84	3.10%
5	51.60	44.90	13.01%	49.86	3.37%
6	47.65	40.18	15.67%	45.86	3.75%
7	43.69	35.47	18.81%	41.88	4.14%
8	39.73	30.73	22.6%	37.89	4.63%

5.2 Results

The model results compared with data in Winter and Summer are shown in the Figures (5.7). The Figures shows the window of MED Simulator, It is shows some values as mass of brine and mass of distillate, latent heat, temperatures ,salinity , heat load ,physical properties of seawater (poilling point elevation and Specific heat of seawater).The results show good agreement between the model and design or vendor values

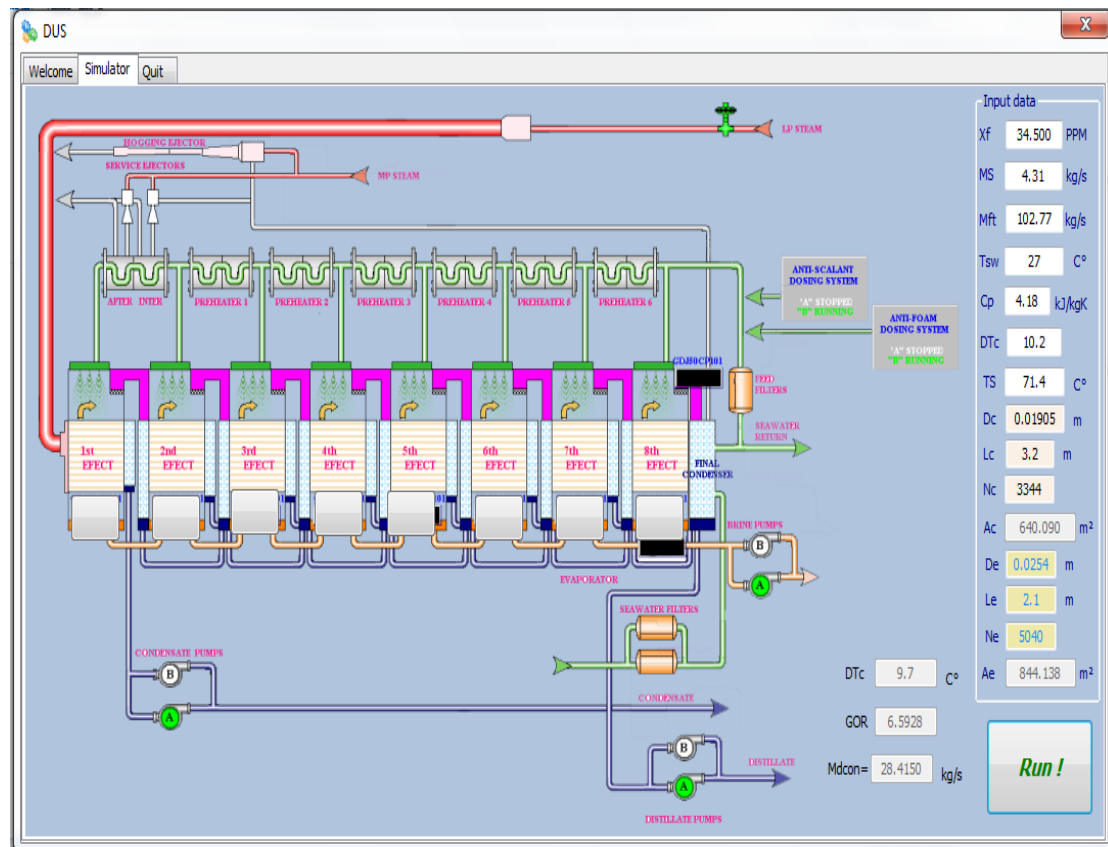


Figure 5.7 shows the result data of the model

5.2.1 The influence of steam temperature

The relation between steam supply temperature and Gain Output Ratio (GOR) for Zawia MED power plant are shown in Figure (5.8). The GOR decrease due to the increase in the steam temperature. This probably due to the increase in the inter effect temperature difference as the steam temperature increase. Then the decrease in the Gain Output Ratio at higher steam temperature is due to the decrease in the amount of latent heat at higher steam temperature which is the available amount of heat to be transferred to the evaporating seawater on the outer surface of tube.

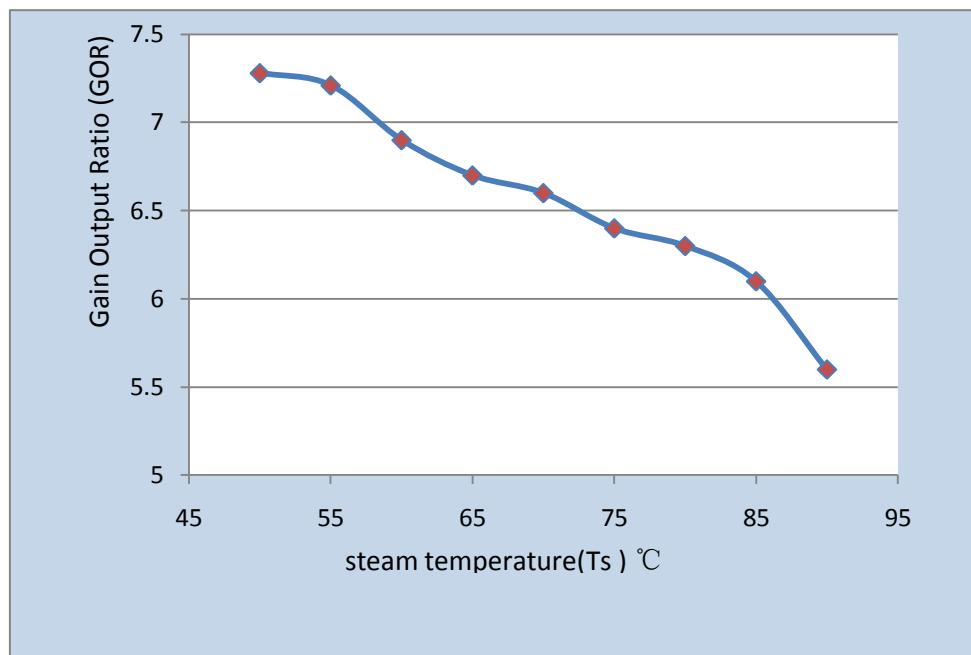


Figure 5.8 impact the steam temperature on Gain Output Ratio

5.2.2 The effect of the mass of steam on GOR and distillate produced

The mass of steam flow rate (M_s) has a very strong effect on the Gain output ratio(GOR) and distillate produced. When the mass of steam supply increase at constant temperature ,the GOR increasing due to increase of latent heat available, and distillate produced , Figures (5.9 and 5.10) shows the relation between the GOR and Distillate produced with mass of steam respectively .It can be seen from Figure 5.9 that the increase of mass flow will leads to sharp increase of GOR while in Figure 5.10 ,linear relationship has been observed.

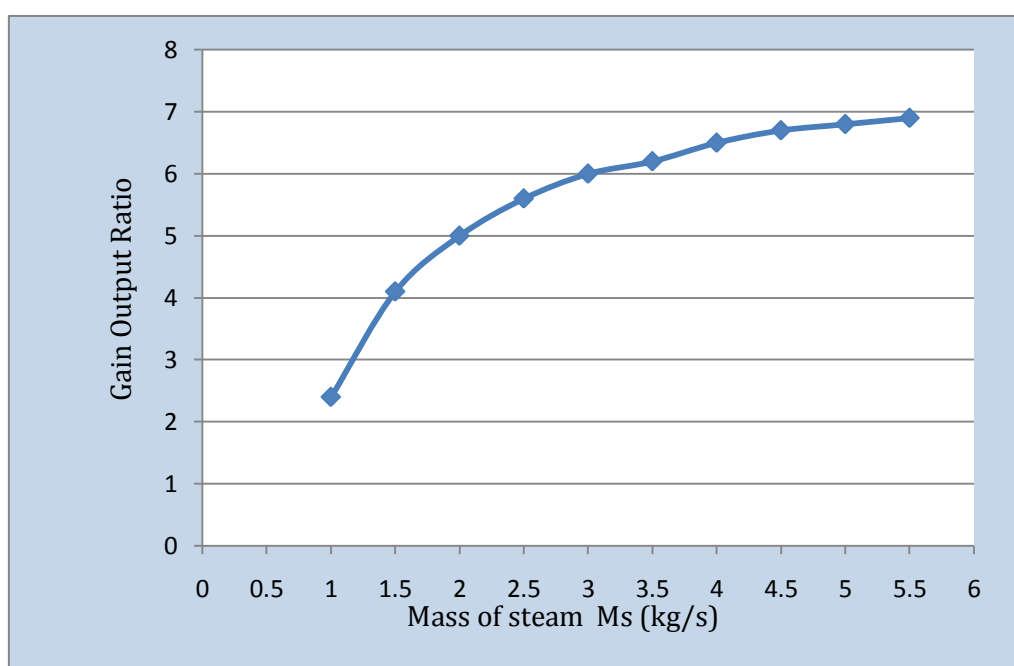


Figure 5.9 impact the mass of steam on Gain Output Ratio

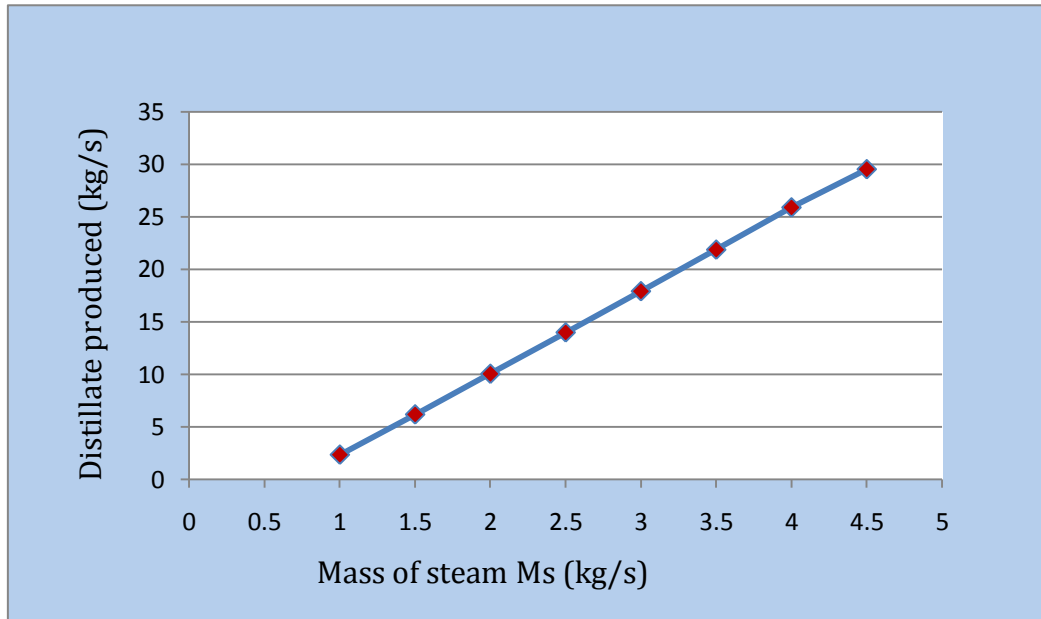


Figure 5.10 impact the mass of steam on Distillate produced

5.2.3 The effect of seawater temperature

The total temperature range of the MED plant is the difference between supply steam temperature and the seawater temperature, which represents the thermal energy transfer in the effects and the condenser. Seawater temperature varies with seasonal changes from 14.0°C in winter to above 30°C in summer. The effect of the inlet seawater temperature on Gain output ratio is shown in Figure (5.11) at fixed steam supply temperature of design temperature (71.4°C), in Figure (5.12) the total distillate produce increase with increasing seawater temperature that is due to increase in heat content in the feed, so less preheating required.

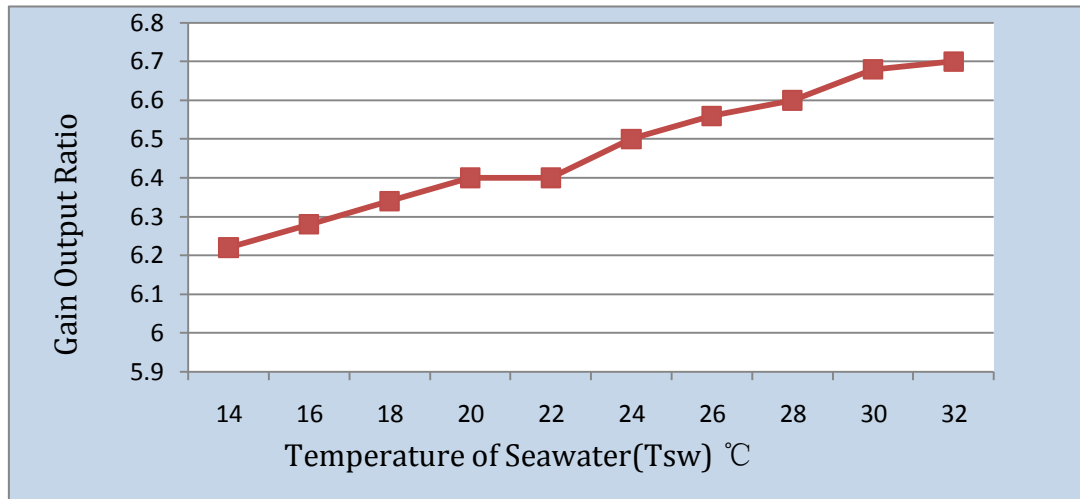


Figure 5.11 Effect of temperature of seawater on Gain Output Ratio(GOR)

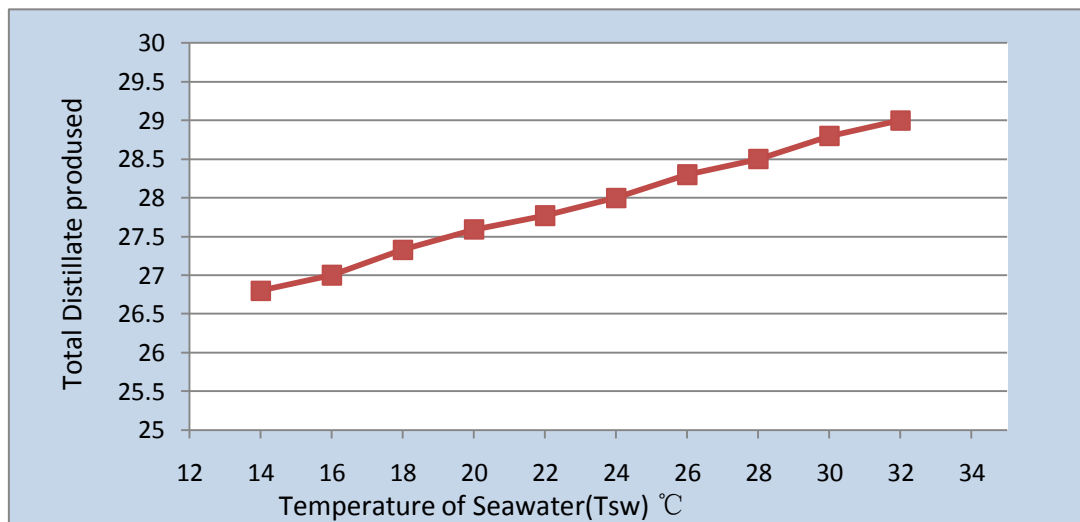


Figure5.12 Effect of temperature of seawater on Total Distillate produced (kg/s)

CHAPTER SIX

Conclusion

The model developed in this work, can simulate efficiently the performance of the Zawia MED power plant, which is based on the detailed mass and energy balances of the streams flowing in each cell. Visual Basic Studio program (2010) was used to written programme which characterized by simplicity .

The important conclusions which can be drown from this study can be stated as follows:

- 1.A steady state model is useful in the study of process variables interaction, troubleshooting ,safety and reliability studies.
- 2.The results obtained are comparable with those given in the literature.
3. The results of the model compared with design data and real data in Summer and Winter and show good agreement.
- 4.The steam require mass in Summer was less than that in Winter because the temperature of seawater inlet in Summer is higher.
- 5.The total distillate produced in Winter was 10.53% less than that design value. While the variation on the total distillate produced in Summer was 6.36%less ,and the variation on the total distillate produced which it take from design data was 1.672% less.

6. The comparison of vapour and brine temperature in each effects from operation data of the Zawia plant , were excellent agreement between the real and calculated temperatures.

7. The GOR decrease due to the increase in the steam temperature, while it is gradually increasing when the mass of steam increasing.

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APPENDIXES

APPENDIX A

Physical properties of seawater

A.1.seawater specific heat at constant pressure

The seawater specific heat at constant pressure is given by the following correlation

$$C_p = (A + BT + CT^2 + DT^3) \times 10^{-3}$$

The variables A, B, C and D are evaluated as a function of the water salinity as follows:

$$A = 4206.8 - 6.6197 s + 1.2288 \times 10^{-2} s^2$$

$$B = -1.1262 + 5.4178 \times 10^{-2} s - 2.2719 \times 10^{-4} s^2$$

$$C = 1.2026 \times 10^{-2} - 5.3566 \times 10^{-4} s + 1.8906 \times 10^{-6} s^2$$

$$D = 6.8777 \times 10^{-7} + 1.517 \times 10^{-6} s - 4.4268 \times 10^{-9} s^2$$

where C_p in $\text{kJ/kg}^\circ\text{C}$, T in $^\circ\text{C}$, and s is the water salinity in mg/kg .

The above correlation is valid over salinity and temperature ranges of $20000 < X < 160000$ ppm and $20 < T < 180$ $^\circ\text{C}$, respectively.

A.2. BOILING POINT ELEVATION OF SEAWATER

The correlation for the boiling point elevation of seawater is

$$\text{BPE} = AX + BX^2 + CX^3$$

$$A = (8.325 \times 10^{-2} + 1.883 \times 10^{-4} T + 4.02 \times 10^{-6} T^2)$$

$$B = (-7.625 \times 10^{-4} + 9.02 \times 10^{-5} T - 5.2 \times 10^{-7} T^2)$$

$$C = (1.522 \times 10^{-4} - 3 \times 10^{-6} T - 3 \times 10^{-8} T^2)$$

where T is the temperature in °C and X is the salt weight percentage. The above equation is valid over the following ranges: $1 < X < 16\%$, $10 < T < 180^\circ\text{C}$.

Physical properties of water

A.3. Latent Heat of Water Evaporation [35]

The correlation for latent heat of water evaporation is given by:

$$\lambda = 2501.897149 - 2.407064037 T + 1.192217 \times 10^{-3} T^2 - 1.5863 \times 10^{-5} T^3$$

In the above equation, T is the saturation temperature in °C and λ is the latent

heat in kJ/kg. The percentage errors for the calculated versus the steam table

values are less than 0.026%.

Condensers and Evaporators Overall Heat Transfer Coefficient

A.4. Fouled condenser

$$U_c = 1 \times 10^{-3} (1617.5 + 0.1537 T + 0.1825 T^2 - 0.00008026 T^3)$$

A.5. Fouled evaporator

$$U_e = 1 \times 10^{-3} (1939.4 + 1.40562 T - 0.0207525 T^2 + 0.0023186 T^3)$$

APPENDIX B

PROGRAM INPUT DATA DESCRIPTION

For steady state analysis of an existing MED plant

FIRST INPUT

Steam mass in kg/s, steam temperature in °C, steam latent heat in kJ/kg.

SECOND INPUT

Condenser cooling water in kg/s, Feed rate in kg/s.

THIRD INPUT

Seawater temperature in °C, seawater salt concentration in kg/kg.

FOURTH INPUT

Number of effects [N] , Number of Pre-heaters=[N-2] First guess of condenser range °C.

FIFTH INPUT

Effect area (A_e) m^2 , Outer diameter of tube for effect (D_e) m ,Length of tube for effect (L_e)m, Number of tube for effect (N_e) .

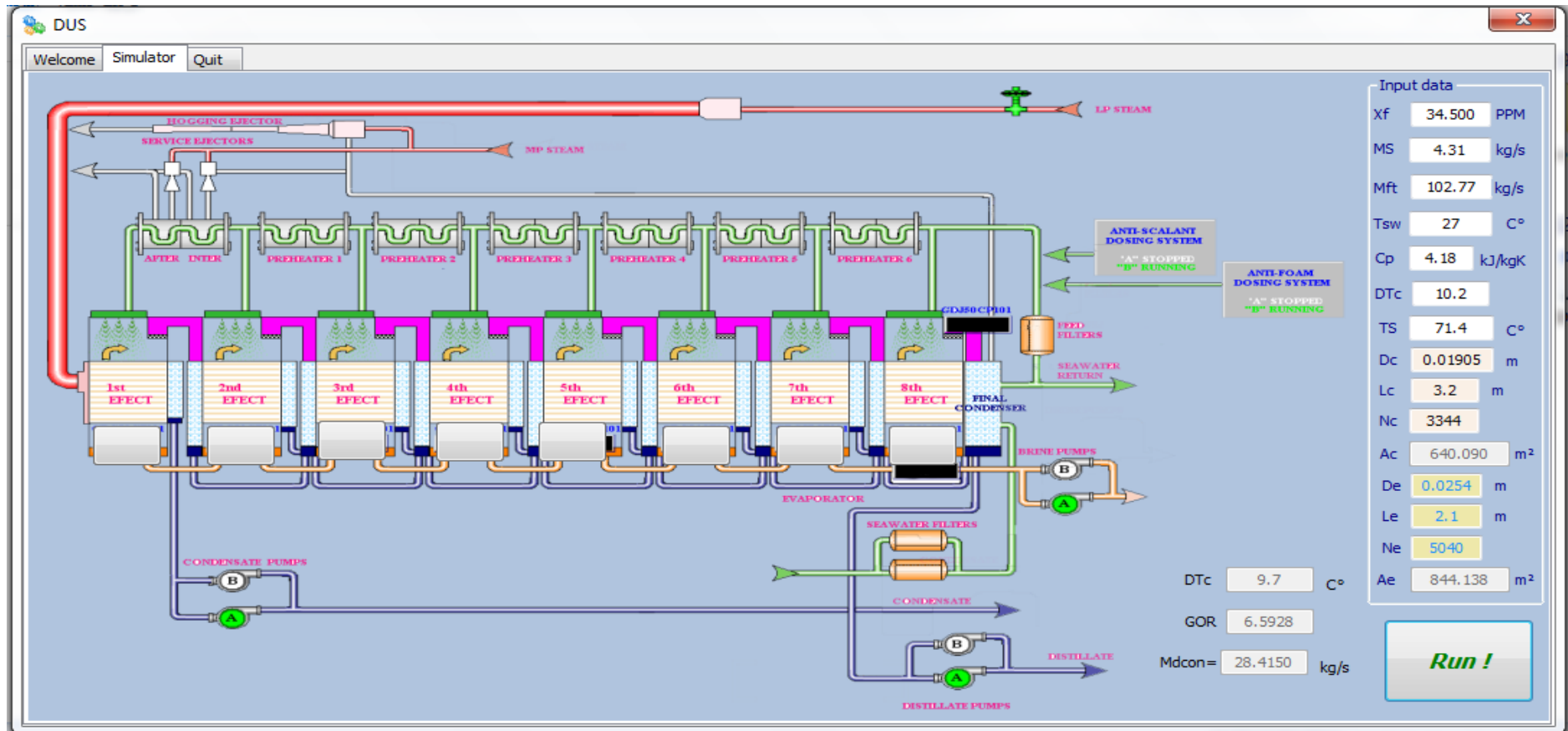
Condenser area (A_c) m^2 , Outer diameter of tube for Condenser (D_c) m

Length of tube for condenser (L_c) m, Number of tube for condenser (N_c)

APPENDIX C

DESALINATION UNIT SIMULATOR RESULTS OF MED ZAWIA POWER PLANT

Figure C.1 Results of med in Zawia power plant for design data at temperature (27°C)



CELL NO 1

Characteristic	Symbol	Value	Unit
Boiling-point elevation	BPE	0.419	°C·kg/mol
Vapor Temperature	T _v	67.022	C°
Latent heat at brine temperature	λ at T _b	2340.391	kJ/kg
Latent heat at vapor temperature	λ at T _v	2341.445	kJ/kg
Latent heat at motive steam	λ_s	2330.397	kJ/kg
Mass of vapor which formed by boiling	M _v	4.221	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.000	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.000	kg/s
Specific heat at constant pressure	C _{pt}	3.899	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	8.625	kg/s
Mass of brine stream leaving effect	M _{bn}	8.625	kg/s
Total distillate flow rate	M _{dn}	0.000	kg/s
Salt concentration inside effect	X _{ben}	51.383	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	51.383	g/kg
Thermal Load	QT	10044.010	kW

CELL NO 2

Boiling-point elevation	BPE	0.409	°C·kg/mol
Vapor Temperature	T _v	63.074	C°
Latent heat at brine temperature	λ at T _b	2350.313	kJ/kg
Latent heat at vapor temperature	λ at T _v	2351.334	kJ/kg
Latent heat at motive steam	λ_s	2330.397	kJ/kg
Mass of vapor which formed by boiling	M _v	4.135	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.000	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.057	kg/s
Specific heat at constant pressure	C _{pt}	3.900	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	8.712	kg/s
Mass of brine stream leaving effect	M _{bn}	17.280	kg/s
Total distillate flow rate	M _{dn}	4.221	kg/s
Salt concentration inside effect	X _{ben}	50.874	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	51.295	g/kg
Thermal Load	QT	9882.918	kW

CELL NO3

Boiling-point elevation	BPE	0.399	°C·kg/mol
Vapor Temperature	T _v	59.125	C°
Latent heat at brine temperature	λ at T _b	2360.163	kJ/kg
Latent heat at vapor temperature	λ at T _v	2361.152	kJ/kg
Latent heat at motive steam	λ_s	2330.397	kJ/kg
Mass of vapor which formed by boiling	M _v	3.926	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.030	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.113	kg/s
Specific heat at constant pressure	C _{pt}	3.901	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	8.921	kg/s
Mass of brine stream leaving effect	M _{bn}	26.088	kg/s
Total distillate flow rate	M _{dn}	8.202	kg/s
Salt concentration inside effect	X _{ben}	49.682	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	50.965	g/kg
Thermal Load	QT	9429.838	kW

CELL NO 4

Boiling-point elevation	BPE	0.390	°C·kg/mol
Vapor Temperature	T _v	55.177	C°
Latent heat at brine temperature	λ at T _b	2369.940	kJ/kg
Latent heat at vapor temperature	λ at T _v	2370.898	kJ/kg
Latent heat at motive steam	λ_s	2330.397	kJ/kg
Mass of vapor which formed by boiling	M _v	3.714	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.057	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.170	kg/s
Specific heat at constant pressure	C _{pt}	3.903	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	9.132	kg/s
Mass of brine stream leaving effect	M _{bn}	35.050	kg/s
Total distillate flow rate	M _{dn}	11.943	kg/s
Salt concentration inside effect	X _{ben}	48.533	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	50.579	g/kg
Thermal Load	QT	8967.658	kW

CELL NO 5

Boiling-point elevation	BPE	0.380	°C·kg/mol
Vapor Temperature	T _v	51.228	C°
Latent heat at brine temperature	λ at T _b	2379.645	kJ/kg
Latent heat at vapor temperature	λ at T _v	2380.574	kJ/kg
Latent heat at motive steam	λ_s	2330.397	kJ/kg
Mass of vapor which formed by boiling	M _v	3.501	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.083	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.228	kg/s
Specific heat at constant pressure	C _{pt}	3.906	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	9.345	kg/s
Mass of brine stream leaving effect	M _{bn}	44.168	kg/s
Total distillate flow rate	M _{dn}	15.443	kg/s
Salt concentration inside effect	X _{ben}	47.424	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	50.172	g/kg
Thermal Load	QT	8494.931	kW

CELL NO 6

Boiling-point elevation	BPE	0.371	°C·kg/mol
Vapor Temperature	T _v	47.278	C°
Latent heat at brine temperature	λ at T _b	2389.279	kJ/kg
Latent heat at vapor temperature	λ at T _v	2390.178	kJ/kg
Latent heat at motive steam	λ_s	2330.397	kJ/kg
Mass of vapor which formed by boiling	M _v	3.284	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.107	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.286	kg/s
Specific heat at constant pressure	C _{pt}	3.908	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	9.562	kg/s
Mass of brine stream leaving effect	M _{bn}	53.444	kg/s
Total distillate flow rate	M _{dn}	18.701	kg/s
Salt concentration inside effect	X _{ben}	46.348	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	49.756	g/kg
Thermal Load	QT	8010.230	kW

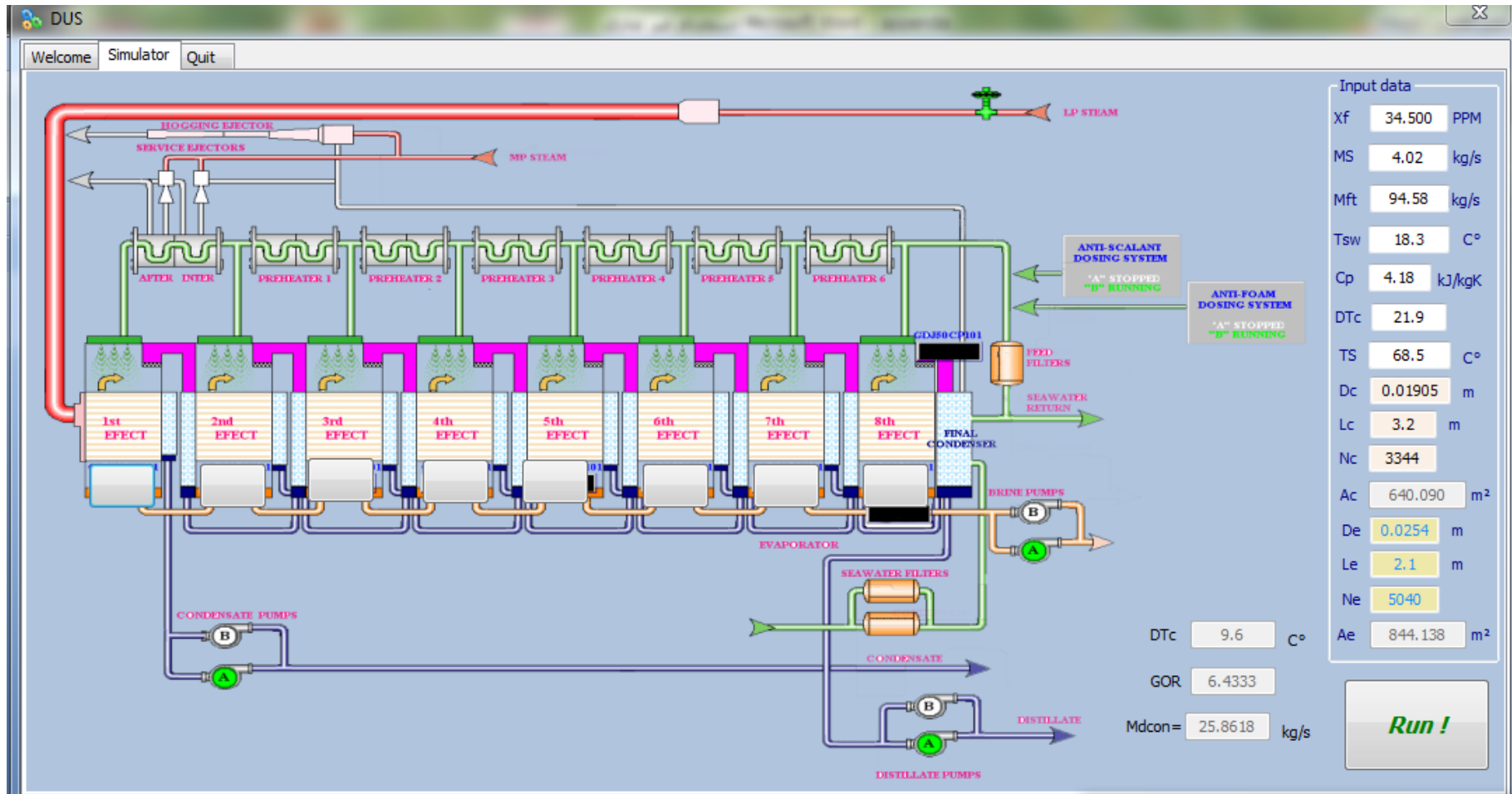
CELL NO 7

Boiling-point elevation	BPE	0.362	°C·kg/mol
Vapor Temperature	T _v	43.329	C°
Latent heat at brine temperature	λ at T _b	2398.839	kJ/kg
Latent heat at vapor temperature	λ at T _v	2399.710	kJ/kg
Latent heat at motive steam	λ_s	2330.397	kJ/kg
Mass of vapor which formed by boiling	M _v	3.063	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.129	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.345	kg/s
Specific heat at constant pressure	C _{pt}	3.910	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	9.783	kg/s
Mass of brine stream leaving effect	M _{bn}	62.883	kg/s
Total distillate flow rate	M _{dn}	21.715	kg/s
Salt concentration inside effect	X _{ben}	45.303	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	49.336	g/kg
Thermal Load	QT	7512.153	kW

CELL NO 8

Boiling-point elevation	BPE	0.353	°C·kg/mol
Vapor Temperature	T _v	39.379	C°
Latent heat at brine temperature	λ at T _b	2408.328	kJ/kg
Latent heat at vapor temperature	λ at T _v	2409.172	kJ/kg
Latent heat at motive steam	λ_s	2330.397	kJ/kg
Mass of vapor which formed by boiling	M _v	2.838	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.149	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.404	kg/s
Specific heat at constant pressure	C _{pt}	3.912	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	10.008	kg/s
Mass of brine stream leaving effect	M _{bn}	72.486	kg/s
Total distillate flow rate	M _{dn}	24.482	kg/s
Salt concentration inside effect	X _{ben}	44.285	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	48.914	g/kg
Thermal Load	QT	6999.319	kW

Figure C.2 Results of med in Zawia power plant for data in Winter at temperature (18.3°C)



CELL NO 1

Characteristic	Symbol	Value	Unit
Boiling-point elevation	BPE	0.410	°C·kg/mol
Vapor Temperature	T _v	63.385	C°
Latent heat at brine temperature	λ at T _b	2349.535	kJ/kg
Latent heat at vapor temperature	λ at T _v	2350.559	kJ/kg
Latent heat at motive steam	λ_s	2337.726	kJ/kg
Mass of vapor which formed by boiling	M _v	3.935	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.000	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.000	kg/s
Specific heat at constant pressure	C _{pt}	3.897	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	7.888	kg/s
Mass of brine stream leaving effect	M _{bn}	7.888	kg/s
Total distillate flow rate	M _{dn}	0.000	kg/s
Salt concentration inside effect	X _{ben}	51.712	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	51.712	g/kg
Thermal Load	QT	9397.657	kW

CELL NO 2

Boiling-point elevation	BPE	0.398	°C·kg/mol
Vapor Temperature	T _v	58.691	C°
Latent heat at brine temperature	λ at T _b	2361.242	kJ/kg
Latent heat at vapor temperature	λ at T _v	2362.228	kJ/kg
Latent heat at motive steam	λ_s	2337.726	kJ/kg
Mass of vapor which formed by boiling	M _v	3.853	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.000	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.061	kg/s
Specific heat at constant pressure	C _{pt}	3.898	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	7.970	kg/s
Mass of brine stream leaving effect	M _{bn}	15.796	kg/s
Total distillate flow rate	M _{dn}	3.935	kg/s
Salt concentration inside effect	X _{ben}	51.178	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	51.643	g/kg
Thermal Load	QT	9249.403	kW

CELL NO 3

Boiling-point elevation	BPE	0.387	°C·kg/mol
Vapor Temperature	T _v	53.997	C°
Latent heat at brine temperature	λ at T _b	2372.847	kJ/kg
Latent heat at vapor temperature	λ at T _v	2373.796	kJ/kg
Latent heat at motive steam	λ_s	2337.726	kJ/kg
Mass of vapor which formed by boiling	M _v	3.637	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.033	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.122	kg/s
Specific heat at constant pressure	C _{pt}	3.900	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	8.186	kg/s
Mass of brine stream leaving effect	M _{bn}	23.860	kg/s
Total distillate flow rate	M _{dn}	7.620	kg/s
Salt concentration inside effect	X _{ben}	49.826	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	51.284	g/kg
Thermal Load	QT	8780.728	kW

CELL NO 4

Boiling-point elevation	BPE	0.376	°C·kg/mol
Vapor Temperature	T _v	49.303	C°
Latent heat at brine temperature	λ at T _b	2384.350	kJ/kg
Latent heat at vapor temperature	λ at T _v	2385.264	kJ/kg
Latent heat at motive steam	λ_s	2337.726	kJ/kg
Mass of vapor which formed by boiling	M _v	3.418	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.063	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.184	kg/s
Specific heat at constant pressure	C _{pt}	3.902	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	8.404	kg/s
Mass of brine stream leaving effect	M _{bn}	32.080	kg/s
Total distillate flow rate	M _{dn}	11.054	kg/s
Salt concentration inside effect	X _{ben}	48.533	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	50.857	g/kg
Thermal Load	QT	8301.975	kW

CELL NO 5

Boiling-point elevation	BPE	0.365	°C·kg/mol
Vapor Temperature	T _v	44.608	C°
Latent heat at brine temperature	λ at T _b	2395.751	kJ/kg
Latent heat at vapor temperature	λ at T _v	2396.631	kJ/kg
Latent heat at motive steam	λ_s	2337.726	kJ/kg
Mass of vapor which formed by boiling	M _v	3.197	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.091	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.246	kg/s
Specific heat at constant pressure	C _{pt}	3.904	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	8.625	kg/s
Mass of brine stream leaving effect	M _{bn}	40.460	kg/s
Total distillate flow rate	M _{dn}	14.238	kg/s
Salt concentration inside effect	X _{ben}	47.290	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	50.405	g/kg
Thermal Load	QT	7811.271	kW

CELL NO 6

Boiling-point elevation	BPE	0.355	°C·kg/mol
Vapor Temperature	T _v	39.913	C°
Latent heat at brine temperature	λ at T _b	2407.050	kJ/kg
Latent heat at vapor temperature	λ at T _v	2407.897	kJ/kg
Latent heat at motive steam	λ_s	2337.726	kJ/kg
Mass of vapor which formed by boiling	M _v	2.973	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.116	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.309	kg/s
Specific heat at constant pressure	C _{pt}	3.907	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	8.850	kg/s
Mass of brine stream leaving effect	M _{bn}	49.000	kg/s
Total distillate flow rate	M _{dn}	17.171	kg/s
Salt concentration inside effect	X _{ben}	46.090	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	49.944	g/kg
Thermal Load	QT	7306.775	kW

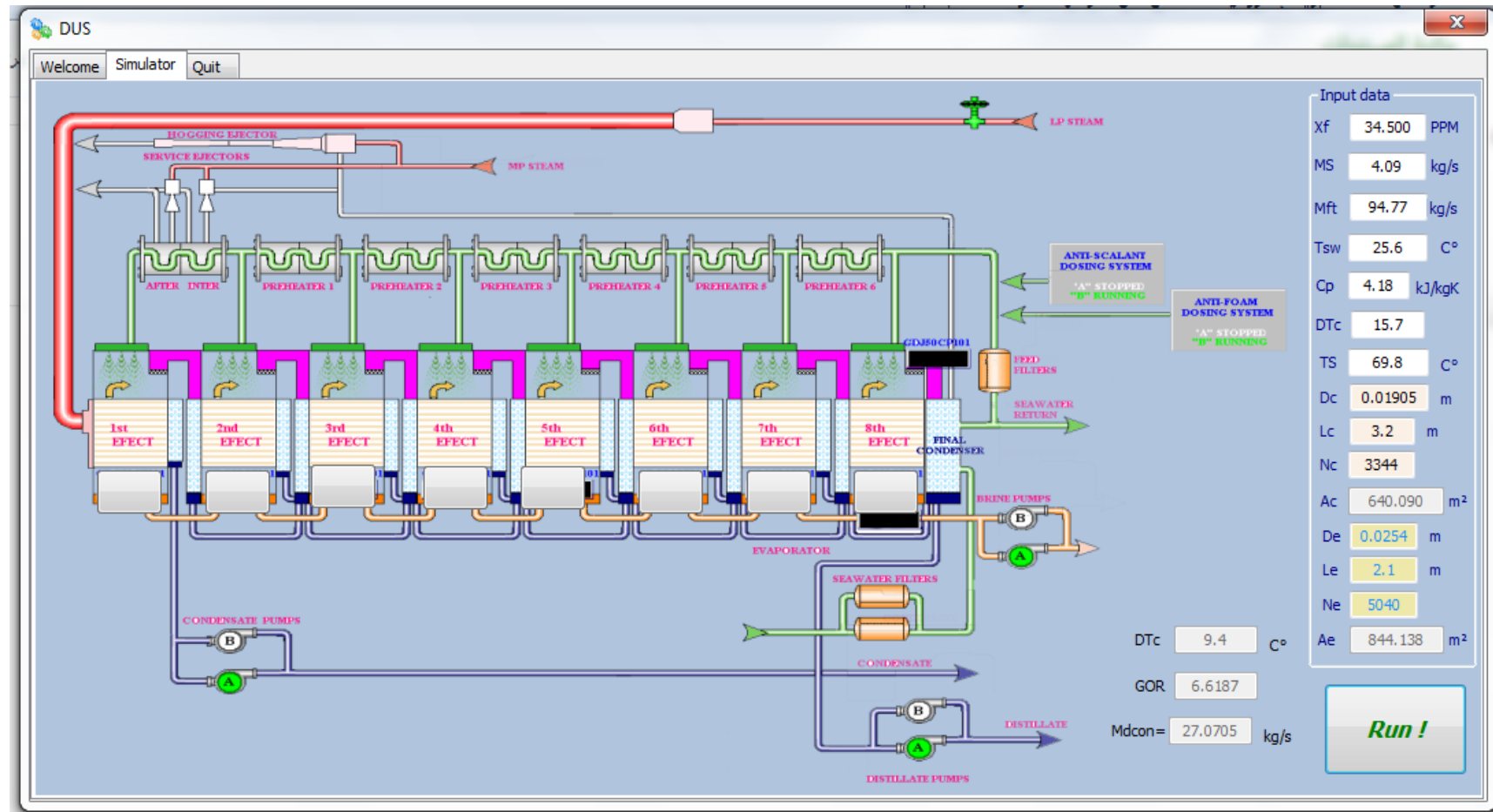
CELL NO 7

Boiling-point elevation	BPE	0.344	°C·kg/mol
Vapor Temperature	T _v	35.217	C°
Latent heat at brine temperature	λ at T _b	2418.247	kJ/kg
Latent heat at vapor temperature	λ at T _v	2419.063	kJ/kg
Latent heat at motive steam	λ_s	2337.726	kJ/kg
Mass of vapor which formed by boiling	M _v	2.744	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.140	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.373	kg/s
Specific heat at constant pressure	C _{pt}	3.909	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	9.078	kg/s
Mass of brine stream leaving effect	M _{bn}	57.706	kg/s
Total distillate flow rate	M _{dn}	19.850	kg/s
Salt concentration inside effect	X _{ben}	44.929	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	49.477	g/kg
Thermal Load	QT	6786.681	kW

CELL NO 8

Boiling-point elevation	BPE	0.335	°C·kg/mol
Vapor Temperature	T _v	30.522	C°
Latent heat at brine temperature	λ at T _b	2429.342	kJ/kg
Latent heat at vapor temperature	λ at T _v	2430.128	kJ/kg
Latent heat at motive steam	λ_s	2337.726	kJ/kg
Mass of vapor which formed by boiling	M _v	2.511	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.161	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.437	kg/s
Specific heat at constant pressure	C _{pt}	3.912	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	9.312	kg/s
Mass of brine stream leaving effect	M _{bn}	66.581	kg/s
Total distillate flow rate	M _{dn}	22.273	kg/s
Salt concentration inside effect	X _{ben}	43.801	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	49.008	g/kg
Thermal Load	QT	6249.217	kW

Figure C.3 Results of med in Zawia power plant for data in Summer at temperature (25.6°C)



CELL NO 1

Boiling-point elevation	BPE	0.415	°C·kg/mol
Vapor Temperature	T _v	65.410	C°
Latent heat at brine temperature	λ at T _b	2344.452	kJ/kg
Latent heat at vapor temperature	λ at T _v	2345.492	kJ/kg
Latent heat at motive steam	λ_s	2334.445	kJ/kg
Mass of vapor which formed by boiling	M _v	4.007	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.000	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.000	kg/s
Specific heat at constant pressure	C _{pt}	3.895	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	7.839	kg/s
Mass of brine stream leaving effect	M _{bn}	7.839	kg/s
Total distillate flow rate	M _{dn}	0.000	kg/s
Salt concentration inside effect	X _{ben}	52.137	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	52.137	g/kg
Thermal Load	QT	9547.881	kW

CELL NO 2

Boiling-point elevation	BPE	0.405	°C·kg/mol
Vapor Temperature	T _v	61.445	C°
Latent heat at brine temperature	λ at T _b	2354.385	kJ/kg
Latent heat at vapor temperature	λ at T _v	2355.393	kJ/kg
Latent heat at motive steam	λ_s	2334.445	kJ/kg
Mass of vapor which formed by boiling	M _v	3.927	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.000	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.052	kg/s
Specific heat at constant pressure	C _{pt}	3.896	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	7.919	kg/s
Mass of brine stream leaving effect	M _{bn}	15.706	kg/s
Total distillate flow rate	M _{dn}	4.007	kg/s
Salt concentration inside effect	X _{ben}	51.611	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	52.043	g/kg
Thermal Load	QT	9399.329	kW

CELL NO 3

Boiling-point elevation	BPE	0.395	°C·kg/mol
Vapor Temperature	T _v	57.480	C°
Latent heat at brine temperature	λ at T _b	2364.246	kJ/kg
Latent heat at vapor temperature	λ at T _v	2365.222	kJ/kg
Latent heat at motive steam	λ_s	2334.445	kJ/kg
Mass of vapor which formed by boiling	M _v	3.733	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.028	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.103	kg/s
Specific heat at constant pressure	C _{pt}	3.897	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	8.113	kg/s
Mass of brine stream leaving effect	M _{bn}	23.716	kg/s
Total distillate flow rate	M _{dn}	7.791	kg/s
Salt concentration inside effect	X _{ben}	50.375	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	51.698	g/kg
Thermal Load	QT	8978.484	kW

CELL NO 4

Boiling-point elevation	BPE	0.386	°C·kg/mol
Vapor Temperature	T _v	53.515	C°
Latent heat at brine temperature	λ at T _b	2374.034	kJ/kg
Latent heat at vapor temperature	λ at T _v	2374.979	kJ/kg
Latent heat at motive steam	λ_s	2334.445	kJ/kg
Mass of vapor which formed by boiling	M _v	3.537	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.055	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.155	kg/s
Specific heat at constant pressure	C _{pt}	3.900	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	8.309	kg/s
Mass of brine stream leaving effect	M _{bn}	31.871	kg/s
Total distillate flow rate	M _{dn}	11.351	kg/s
Salt concentration inside effect	X _{ben}	49.187	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	51.294	g/kg
Thermal Load	QT	8549.233	kW

CELL NO 5

Boiling-point elevation	BPE	0.376	°C·kg/mol
Vapor Temperature	T _v	49.549	C°
Latent heat at brine temperature	λ at T _b	2383.749	kJ/kg
Latent heat at vapor temperature	λ at T _v	2384.665	kJ/kg
Latent heat at motive steam	λ_s	2334.445	kJ/kg
Mass of vapor which formed by boiling	M _v	3.339	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.079	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.207	kg/s
Specific heat at constant pressure	C _{pt}	3.902	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	8.508	kg/s
Mass of brine stream leaving effect	M _{bn}	40.171	kg/s
Total distillate flow rate	M _{dn}	14.687	kg/s
Salt concentration inside effect	X _{ben}	48.039	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	50.870	g/kg
Thermal Load	QT	8110.231	kW

CELL NO 6

Boiling-point elevation	BPE	0.367	°C·kg/mol
Vapor Temperature	T _v	45.583	C°
Latent heat at brine temperature	λ at T _b	2393.391	kJ/kg
Latent heat at vapor temperature	λ at T _v	2394.278	kJ/kg
Latent heat at motive steam	λ_s	2334.445	kJ/kg
Mass of vapor which formed by boiling	M _v	3.137	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.102	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.260	kg/s
Specific heat at constant pressure	C _{pt}	3.904	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	8.709	kg/s
Mass of brine stream leaving effect	M _{bn}	48.619	kg/s
Total distillate flow rate	M _{dn}	17.797	kg/s
Salt concentration inside effect	X _{ben}	46.928	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	50.436	g/kg
Thermal Load	QT	7660.154	kW

CELL NO 7

Boiling-point elevation	BPE	0.358	°C·kg/mol
Vapor Temperature	T _v	41.617	C°
Latent heat at brine temperature	$\lambda_{at Tb}$	2402.961	kJ/kg
Latent heat at vapor temperature	$\lambda_{at Tv}$	2403.820	kJ/kg
Latent heat at motive steam	λ_s	2334.445	kJ/kg
Mass of vapor which formed by boiling	M _v	2.932	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.123	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.314	kg/s
Specific heat at constant pressure	C _{pt}	3.907	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	8.914	kg/s
Mass of brine stream leaving effect	M _{bn}	57.219	kg/s
Total distillate flow rate	M _{dn}	20.681	kg/s
Salt concentration inside effect	X _{ben}	45.850	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	49.998	g/kg
Thermal Load	QT	7197.698	kW

CELL NO 8

Boiling-point elevation	BPE	0.350	°C·kg/mol
Vapor Temperature	T _v	37.651	C°
Latent heat at brine temperature	$\lambda_{at Tb}$	2412.458	kJ/kg
Latent heat at vapor temperature	$\lambda_{at Tv}$	2413.289	kJ/kg
Latent heat at motive steam	λ_s	2334.445	kJ/kg
Mass of vapor which formed by boiling	M _v	2.724	kg/s
Mass of vapor which formed by distillate flashing	M _{dfn}	0.142	kg/s
Mass of vapor which formed by brine flashing	M _{bfm}	0.368	kg/s
Specific heat at constant pressure	C _{pt}	3.909	kJ/kg.C°
Mass of brine which formed inside effect	M _{ben}	9.123	kg/s
Mass of brine stream leaving effect	M _{bn}	65.973	kg/s
Total distillate flow rate	M _{dn}	23.335	kg/s
Salt concentration inside effect	X _{ben}	44.800	g/kg
Salt concentration at brine stream leaving effect	X _{bn}	49.559	g/kg
Thermal Load	QT	6721.580	kW
