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Simulation and Economic Study of the MED-TVC Units at Umm Al Nar Desalination Plant

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Simulation and Economic Study of the MED-TVC Units at Umm Al-Nar Desalination Plant

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SUMMARY

A rigorous mathematical model has been used to develop a steady state simulation program MEDNAR to analyze the multi-effect thermal vapor compression desalination (MED-TVC) plant at Umm Al-Nar power and desalination plant. The effect of thermodynamic losses on the thermal performance ratio, the specific heat transfer area and the specific flow rate of the cooling water are taken into account. The losses contemplated are the boiling point elevation and the temperature depression corresponding to the pressure drop during the vapor condensation process. The MEDNAR also takes into consideration the variation in the physical properties of the seawater with temperature and salt concentration, and the effect of the presence of non-condensable gases on the heat transfer coefficients in the evaporators. Sensitivity analyses, using this software, were conducted to study the effect of a number of process variables on the plant performance.

As a part of this thesis, an economic study has been carried out to determine the total water unit cost. This includes a cost break down in details for capital investment and production costs. Also an economic sensitivity analysis was performed to investigate the relationships between the fuel cost, plant production rate, operation and maintenance cost and plant running factor, and the total water unit cost.

TABLE OF CONTENTS

TABLE OF FIGURES

TABEL OF TABELS

1 INTRODUCTION

Desalination of seawater has been developed considerably over the past four decades. The number of desalination plants in 1960 was 5 units with total production rate of 5000 m^3/d, while in 1999 the number of units was around 12500 with a capacity of $22.8*10^6$ m^3/d. This capacity is expected to double within the next 20 years [1].

Nevertheless, many other countries still suffer from water shortage and cannot afford the high capital required and running costs of desalination plants. Desalination processes have so far mostly been built when there is no practical alternative, cost of other water resources is high, low cost energy is readily available and high living standards override the cost factor. These conditions are met in the Arabian Gulf countries. The first application of seawater desalination in the gulf began in Kuwait in 1958, with a multi-stage flash (MSF) plant of 1.0 MIGD [2].

Presently, the majority of desalination plants in operation are based on the multi-stage flash (MSF) desalination system. However, the Multi-Effect Desalination (MED) process has recently acquired a potential interest as a large-scale desalinations technique. This dose not means that the MED process is new technology. The problems were in low unit size and gain ratio. Up to the 1980s, the MED with its different combinations was well known as a small-scale desalination technique.

In UAE, tens of MED units have been supplied in different places during the last three decades. In general, the unit size varies from 17 to 700 m^3/d. The MED units were installed in arid areas, islands and offshore platforms. In 1973, the first two multiple effect thermal vapor compression units were installed in Das Island, each unit consists of two effects and a capacity of 125 m^3/d. Many other plants had been also installed in Sir Baniyas Island, Abu Mussa Island, Delma Island, Alfutaysi Island and Al-Ruwais. The largest size was in Al-Ruwais with 600 m^3/d in 1978 [3].

From the year 1989, UAE played a major rule in developing the MED's plant size. The French company SIDEM developed a process of MED technique with Thermal Vapor Compressor (MED-TVC). This combination allowed for large-scale plants.

Three plants were erected in Jebel Dhanna, Sila and Ras Al Khiema, each plant capacity was 1.0 MIGD [4]. Another plant had been also erected in Mirfa with 2.0 MIGD capacity. In the year 2000, two units were commissioned in Umm Al-Nar Power Station; each plant has a nominal capacity of 3.5 MIGD. These units were then the largest using this desalination technique. A contract had been signed for 14 units -3.8 MIGD each- in Al-Taweelah power plant. Recently, the largest MED-TVC units commissioned in Al-Layyah power station, in Sharjah have a nominal capacity of 5 MIGD. This process is developing rapidly with regard to unit capacity and is gaining more market shares in the desalination market in UAE.

2 RESEARCH OBJECTIVES

The French company Sidem has designed and installed two MED-TVC units at Umm Al Nar. The nominal capacity and thermal energy requirement of each plant is 3.5 MGD and 85 tons of low-pressure steam per hour, respectively.

The proposed study covers the following tasks:

A- Development of a computer algorithm for modeling and simulation of the MED-TVC process at Umm Al Nar based on a rigorous mathematical model. The software should be capable of computing temperature, concentration and flow rate profiles for the whole plant.

B- Validation of the developed software through comparison between the computed results with both design and operational data from the plant, and with corresponding results using the commercial software "Evapolund".

C- Calculation of energy consumption under different operation conditions.

D- Estimation of unit cost of water at full and partial load operating conditions.

3 LITERATURE SURVEY

There are limited number of publications that handle the low temperature multi effect evaporation with the thermal vapor compression process. Most of the available publications have concentrated on describing the application of the process in desalination plants with relatively low capacity and on comparing its features with other desalination process. The following abstracts represent the most related articles to this thesis:

Jernqvist, et al., [5] developed a computer program called Evapolund. It simulates different flow sheets for thermal desalination processes. The program includes a comprehensible database for the physical properties of seawater as well as other liquids. The program can be used for the design and evaluation of all types of thermal desalination processes. Simulation results can be displayed either graphically, or as a text, or in any form such as temperature, pressure and concentration profiles.

Ettouney and El-Dessouky [6] developed a computer package for the design and simulation of thermal desalination processes. The package includes models for multi effect evaporation (MED) systems; both stand alone and with vapor compression. The vapor compression systems include mechanical, thermal, absorption and adsorption heat pumps. The compute

rating, flow charting and performance calculations.

Al-Shammiri and Safar [7] discussed the general features of existing commercial MED plants and associated technical aspects related to steam, condensers, evaporators, pumps and capacity. They also discussed the gain ratio, operating temperature, materials of construction, operation and maintenance of these plants, as well as associated problems, and other available information.

Al-Najem et al. [8] conducted a parametric analysis for the TVC system using first and second laws of thermodynamics, e.g., steam ejector, evaporator, condenser as well as the system as a whole.

Temstet et al. [9] presented a case study for a plant in Sicily having 12 effects with thermo-compression. The design gain output ratio of the unit is 16.7. Heat and mass balance diagrams, material choice, description of plant cells and overall layout were given. Main construction and erection details are described with particular reference to the environmental aspects of this installation.

3.1 Types of Evaporators

The MED process has many possible configurations. The heat transfer area in evaporators can be classified as vertical climbing film, vertical falling film and horizontal falling film. The choice between these configurations depends mainly on the physical properties of the evaporated liquid and the maximum allowable temperature.

3.1.1 Vertical Climbing Film Evaporators

In this configuration steam is condensing on the outside surface of vertical tubes in which seawater is fed at the lower end of the tube. At a predetermined level, the steam causes the brine to boil, vapor to be released and a thin film of brine to be established on the inside surface of the tube wall. The vertical straight-tube design has good heat transfer characteristics, with good operation at low temperature differentials, allowing a reasonable number of effects to be put in series between the available temperature limits and thus achieving high performance ratios.

Water production costs with this design are favorable particularly when a reasonable performance ratio can be achieved [10].

3.1.2 Vertical Falling Film Evaporators

The difference between this design and the climbing film evaporator lies in the way the seawater is distributed within the tubular heat transfer surface. Here, the seawater is allowed to flow inside of the tube wall as a thin film. The objective is to eliminate the wastage on temperature differential caused by the hydrostatic head, and to create better heat transfer properties, which is less dependent on operating temperature and temperature differential. Figure 3-1 shows schematically the different streams entering and leaving this type of evaporators [10].

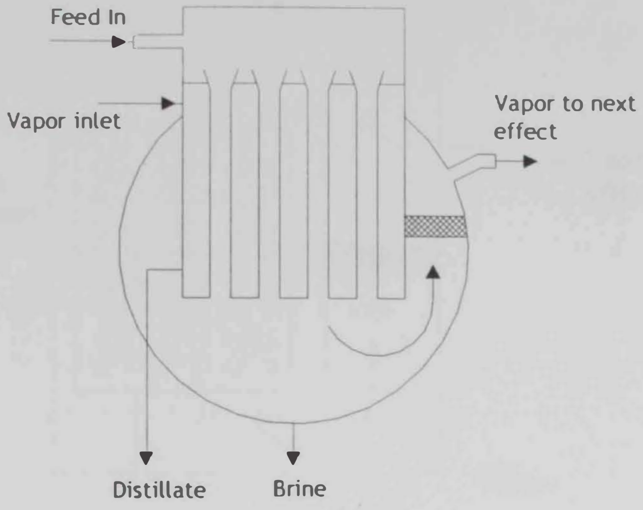


Figure 3-1: Vertical Falling Film Evaporator

3.1.3 Horizontal Falling Film Evaporators

In this configuration, the brine is distributed as a thin film over the outside surface of horizontal tubes, and the heating steam is condensing on the inside surface of the tubes. High heat transfer coefficients are achieved, and this design has a number of distinctive practical *advantages:*

1. Visual monitoring to prevent scale and corrosion.
2. Non-condensable gases are driven positively, and uni-directionally, out of the heat transfer zone.
3. Release of vapor is very gentle, so that droplets carryover is minimal [10].

For potable water, therefore, demisters are not necessary. For process and boiler feed water, the purity obtained with demisters is very high indeed. This design is very stable when operated at reduced loads. The response to changes in terminal conditions is very smooth and rapid owing to the relatively small quantity of brine in the effects. Figure 3-2 shows schematically the different streams entering and leaving this type of evaporators.

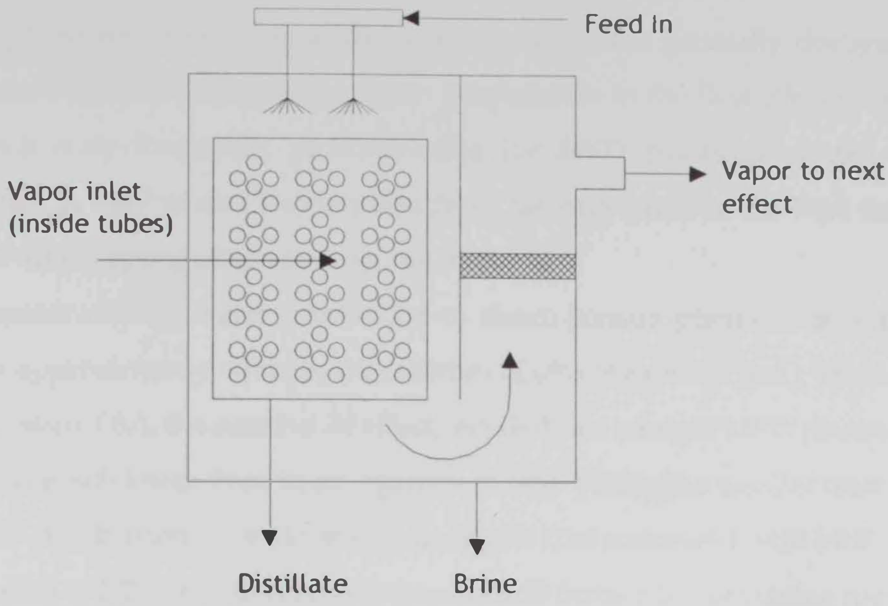


Figure 3-2: Horizontal Falling Film Evaporator

3.2 The Multiple Effect Desalination (MED) Processes

The MED-TVC is the major competitive distillation alternatives to MSF process. The vapor formed in each effect, except the last one, flows to the condensing side of the next lower temperature effect. The latent heat of condensation is transferred through the tube wall to evaporate part of the feed water flowing across the surface. The main difference between MSF and MED is the way evaporation and heat transfer take place. In an MED plant, vapor is generated from seawater film in contact with the heat transfer surface; whereas in the MSF plant seawater is only heated by convection within the tubes and vapor is generated by flashing from a streaming in each stage.

High heat transfer rates are achieved in the MED process due to the thin film boiling and condensing conditions. Also the evaporation process takes place at a uniform temperature within each effect. The required heat transfer area for the MED plant is close to that used by MSF, even though the plant is operated over a smaller temperature range than MSF. Most modern large MED plants have a horizontal tube configuration (HTE), with the feed water film sprayed on the outer side of the tubes and vapor 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. MED plants using polymer additive scale control are generally designed for low temperature operation. The evaporation temperature in the first effect is around 65 °C to limit scale formation. Acid cleaning for MED plants is needed more frequently than in MSF plants Evaporation from the outer side of the heat transfer surface tubes makes sponge ball cleaning not suitable.

The performance ratio of water production to steam consumption of conventional MED plant is approximately equal to the number of effects minus 1 (n-1). To obtain a performance ratio of 8.0, the number of effects needed in a straight MED plant would be 9 [2]. This is much lower than in an equivalent MSF plant. The smaller number of effects in MED plants contributes to saving in capital cost compared with MSF.

The performance of MED plants can be improved still further by combining the MED process with different heat pumps.

The idea from these heat pumps is to improve the thermal performance of the system by integrating the energy of the steam supply to the process.

The modern combinations are:

1) Thermal vapor compressor (steam jet ejector).
2) Mechanical vapor compressor.
3) Absorption column.
4) Adsorption bed.

3.2.1 Process Description

A flow sheet for conventional Multiple Effect Desalination (MED) seawater desalination process is shown in Figure 3-3. The system consists of number of evaporators, final condenser and a feed heater. Cooling seawater is introduced to the inner side of the condenser tubes where it is heated and then divided into two parts. The first and larger part is rejected back to the sea and the second smaller part represents the feed to the evaporator. The rejected seawater removes the excess heat added to the system by the heating steam. The heating of the feed seawater in the condenser increases the thermal performance of the process. The heat source in the condenser is supplied by condensing the vapor formed by boiling in the last evaporator. Accordingly, the condenser has three functions: to remove the extra heat from the system, to improve the process thermal efficiency, and to adjust the boiling temperature inside the evaporator.

Within an evaporator, the feed water is sprayed at the top where it falls in the form of a thin film down the succeeding rows of tubes arranged horizontally. The feed water temperature is raised to the boiling temperature and part of this feed is evaporated. The boiling temperature in the first effect is dictated by the nature of chemicals used to control the scale formation and the state of the heating steam. This temperature is mastered in other effects through adjusting the pressure in the vapor space of the evaporator. The vapor formed by boiling is free of salts and its saturation temperature is less than the boiling temperature by the boiling point elevation. The vapor flows through a knitted wire separator, demister, to remove the entrained brine droplets. The saturation temperature of the vapor departing the demister is reduced to a somewhat lower value because of the frictional pressure loss in the demister.

The vapor leaving the demister of the first effect is transported to the second effect and so on to the last effect. This transport inevitably involves a pressure drop and hence a corresponding decreases in saturation temperature. Another pressure fall and consequent depression in the saturation temperature of the vapor is associated with vapor condensation inside the heat transfer tubes in the evaporators or over the heat transfer area in the pre-heaters. The remaining brine in the first effect flows to the second effect, which operates at a lower pressure.

The vapor is formed in any effect, other than the top effect, by two different mechanisms. The first is by boiling over the heat transfer surfaces and the second by flashing within the liquids (distillate and brine) bulk moving from upstream effect due to pressure drop, and generating a much less amount of vapor. The temperature of the vapor formed by flashing is less than the effect boiling temperature due to non-equilibrium allowance.

It is worth mentioning that the amount of steam generated by evaporation in each effect, other than the first, is less than the amount generated in the previous effect. This is due to the increase in the specific latent heat of vaporization with the decrease in the effect temperature. Moreover, part of the latent heat of an incoming vapor to the effect is used to raise the temperature of the feed to its boiling point. Consequently, the amount of vapor generated in an evaporator by boiling is less than the amount of condensing steam used for heating in the following evaporator. The brine flowing into the last effect reaches its final concentration by evaporating more vapors. The remaining brine is rejected to the sea. The vapor formed by boiling and flashing in the last effect passes to the final condenser.

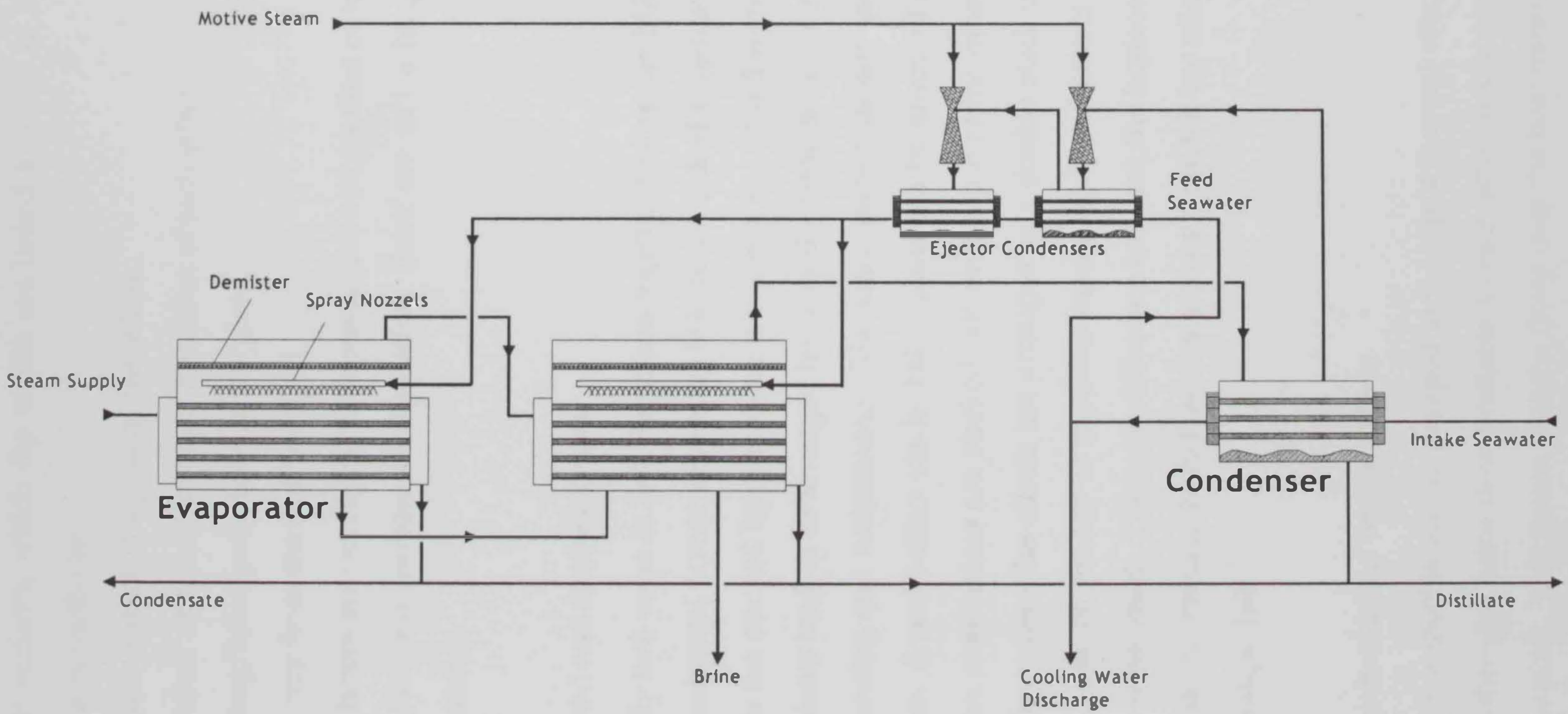


Figure 3-3: Process flow diagram for the multiple effect evaporator (MED) desalination process

3.2.2 MED Flow Sheets

Once a heat transfer surface configuration has been selected for a plant there are still large numbers of variants, which the effects are linked together to form an MED plant. These variants methods:

1. Horizontal or a stacked layout of the effects.
2. 'Re-circulation' or 'Once Through' operation of each effect.
3. Feed heating/brine path through the plant.
4. Heat recovery from the distillate. [10]

Although these choices are not strictly independent of one another or for that matter independent of the heat transfer configuration, there are still a large number of possible combinations.

3.2.2.1 Horizontal and Stacked layout

There are basically two ways to link the effects together to form an MED plant. The first one is the horizontal layout where the effects are linked together horizontally, and the second is the stacked layout where the effects are linked vertically. Almost all larger MED plants tend to be arranged horizontally because of their stability and simplicity in operation and maintenance. The stack layout already exists in many small MED plants. A multi-effect stack (MES) plant can be arranged in two ways: simple MES where evaporators are stacked one on the top of the other or a double stack configuration where the effects are arranged in a double stack configuration, i.e., effects 1,3,5... one on the top of the other in one stack and effects 2,4,6... above each other in the other stack. The main difference between the horizontal and stack arrangement is that the brine in MES flows by gravity from the top effect to the next effects without pumps. [10]

3.2.2.2 Brine Re-circulation within Effects

If the brine from one effect is simply pumped to the top of the next effect and there is no re-circulation then the brine flow available to each effect decreases as the brine flows through the plant. This means that the flows over the heat transfer surfaces in each effect are different and tied to the brine throughput. If, however, the brine is re-circulated over the heat transfer surface by a pump from the last effect sump, then the re-circulation flow rate can be varied independently of the brine feed rate to the effect and can be set at the optimum value regardless of the rest of the plant operation. [10]

3.2.2.3 Brine paths through MED Plants

Another configuration in MED plants is based on brine flow-direction with respect to the vapor direction from one effect to the other. This type of arrangement includes a forward configuration, backward configuration and parallel feed configuration.

The normal arrangement is to have the feed preheated in heat exchangers up to a temperature just below the boiling temperature of the first effect. It is then heated to the evaporation temperature in the first effect. The brine blowdown from the first effect is passed to the second effect as its feed. The brine from the second effect becomes the feed for the third effect and so on. This is the forward configuration.

Separate feed heaters may be eliminated from the flow sheet by feeding the brine in the opposite direction up the temperature gradient of the plant - a little bit of feed heating being done in each effect as the brine passes through. This system is referred to as 'reverse' or 'backward fed'. One of the reasons for using a backward fed system is to avoid separate feed heaters where it is most likely to get trouble with product contamination due to tube leaks.

Parallel feed configuration is also possible in which the brine is fed in parallel to several effects at once. It is not usual for all the effects of a plant to be parallel fed at once. Normally the brine would be fed in series through groups of three or four parallel fed effects. [10]

3.2.2.4 Feed Heating and Heat Recovery Systems

The distillate produced from the high temperature effects contains significant sensible heat. This energy is usually recovered by allowing the distillate to flash off vapor at the various intermediate effects further down the plant. The vapor produced may either be fed to the main evaporator/condensers on the way down or may be used to preheat the feed in the feed heater train. As has been mentioned flashing both of the distillate and of the brine in fact produces a significant amount of vapor in a normal MED plant when the brine is being operated with the feed forward configuration. [10]

3.3 Energy Integrated MED Vapor Compression Processes

There are four types of modern combinations used in MED vapor compression plants, low temperature MED with mechanical vapor compression (MED-MVC); low temperature MED combined with thermal vapor compression (MED-TVC); MED combined with absorption heat pump (MED-ABS); and MED combined with adsorption heat pump (MED-ADS).

Al-Juwayhel et al. [11] conducted a comparative study between these four different types on a single effect evaporator desalination system. The study includes the development of mathematical models for the four systems. The analysis was based on comparison of the performance ratio, specific power consumption, specific heat transfer area, and specific cooling water flow rate. The study shows that the performance ratio for the thermal vapor compression decreases as the boiling temperature and pressure of motive steam increases.

El-Dessouky and Ettouney [7] found through a mathematical simulation model that the MED-ABS system can reach a performance ratio of 25 with only 12 effects at 110 °C, and the MED-ADS system can reach a performance ratio of 20 at the same number of effects and operating temperature. For comparison, the performance ratio of a conventional MED system with the same number of effects cannot exceed 10.

3.3.1 Thermal Vapor Compression (TVC)

The performance of conventional MED plants can be improved further by means of vapor compression whereby part of the vapor formed in a low temperature effect is recompressed and reintroduced into the first effect. In large plants, the method used is thermal vapor compression alternative.

A schematic diagram for the single-effect thermal vapor compression (TVC) seawater desalination process is given in Figure 3-4. The system constitutes of an evaporator, a steam jet ejector, and a final condenser. The vapor leaving the evaporator splits into two parts. The first part flows to the condenser where it condenses and second part is sucked by the steam jet ejector.

In general, the ejector is a pumping device, which uses jet action of a high pressure and temperature primary motive fluid to entrain and accelerate a slower secondary fluid. The steam ejector is used to compress the vapor from pressure P_{ev} (which is the vapor pressure leaving the evaporator) to P_s (which is the vapor pressure entering the first effect) by using an external source of steam at a pressure P_{ms} greater than both P_{ev} and P_s. A typical ejector consists of four parts nozzle, suction chamber, mixing section and diffuser as shown in Figure 3-5. The motive steam is expanded through the nozzle to a low pressure, and high velocity steam jet. This high velocity steam entrains the low vapor pressure leaving the evaporator and entering the ejector from the suction chamber on the low pressure mixing zone. The velocity after mixing is lower than that at the nozzle exit. The loss in the kinetic energy is converted into the pressure discharge head, and the mixture flows through the diffuser, emerging from the discharge at a pressure between the suction and the motive fluid pressures [12].

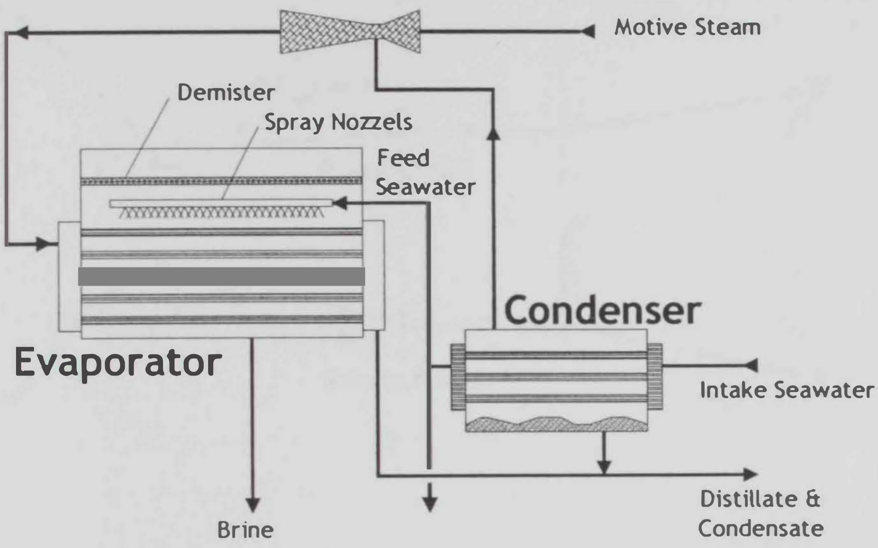


Figure 3-4: Thermal vapor compression evaporator-desalination.

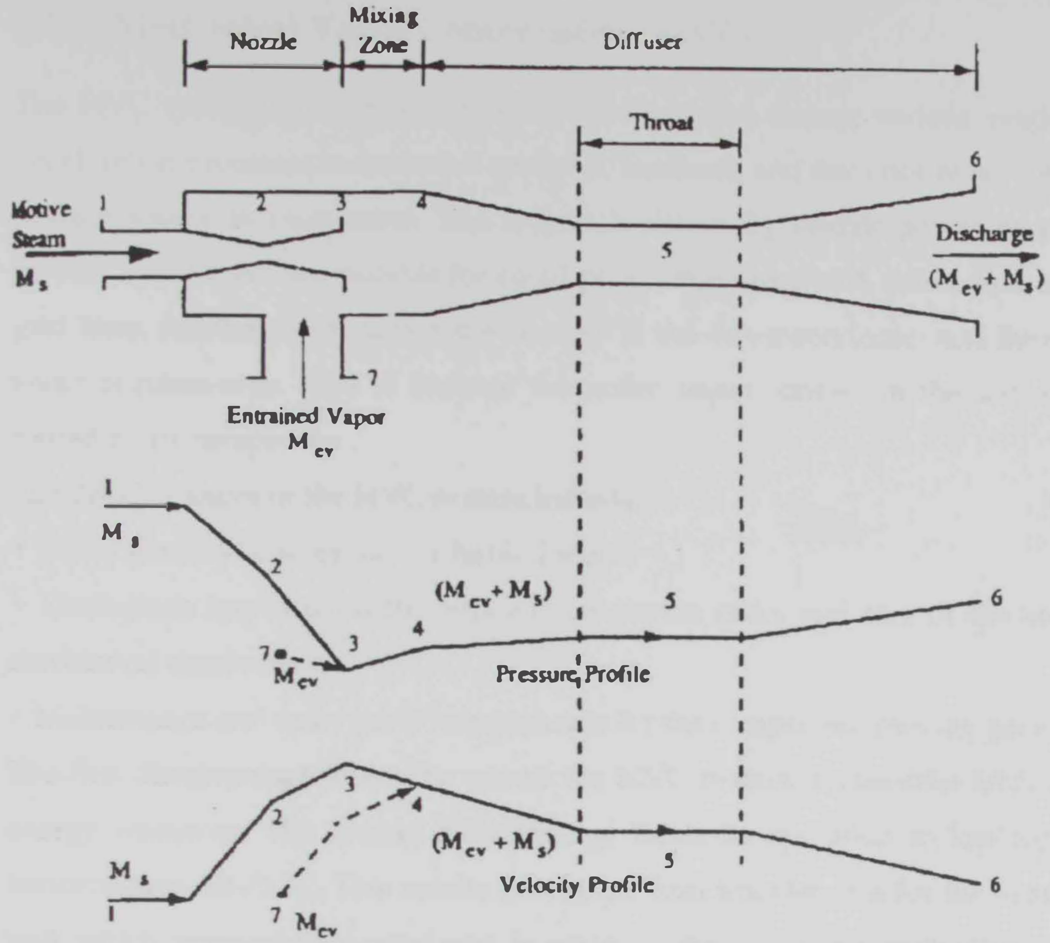


Figure 3-5: Different processes in steam jet ejector [1]

3.3.2 Mechanical Vapor Compression (MVC)

The MVC system is the most attractive configuration among various single stage desalination processes because it is compact, confined, and does not require external heating source in most cases. The system is driven by electric power or a prime mover; and is therefore suitable for small population areas with access to the power grid lines. Another advantage is the absence of the down condenser and the cooling water requirements. This is because the entire vapor formed in the last effect is routed to the compressor.

The disadvantages of the MVC system include:

- Use of electrical energy, or mechanical work.
- Limitations imposed on the vapor compression ratio, and size of the available mechanical compressor.
- Maintenance and spare parts requirements for the compressor moving parts.

The first disadvantage limits the use of the MVC system in countries with limited energy resources. The second disadvantage limits its operation to low top brine temperatures, 60-70 °C. This results in a larger heat transfer area for the evaporator unit, which increases the capital cost. In addition, the single unit production capacity is limited to about 800 m^3/d (0.21 MIGD). This problem is addressed to some extent by operating MVC with multi-effect arrangement with 3 to 4 effects, where the capacity increases to about 3,000 m^3/d (0.8 MIGD) [13]. The last disadvantage increases the operating cost and dictates the use of highly skilled labor.

Energy conservation within the MVC system is maintained by recovery of energy in the rejected brine and distillate product. The specific power consumption of the MVC plant can be below 10 kWh/m^3 [2].

A schematic diagram for the proposed system is shown in Figure 3-6. As shown, the compressor raises the pressure of the vapor formed in the evaporator. The compressed vapor is superheated vapor with saturation temperature higher than the temperature of the boiling brine. This vapor is introduced into the inner side of the evaporator tubes; hence, it is de-superheated and condensed by releasing its latent heat. The condensate and the rejected brine have substantial amount of energy, which is recovered by heating the feed water in multi flow exchanger. Therefore, the feed temperature is increased from a low value of about 25 °C to higher value of 3-6 °C below the condensate and the rejected brine temperature.

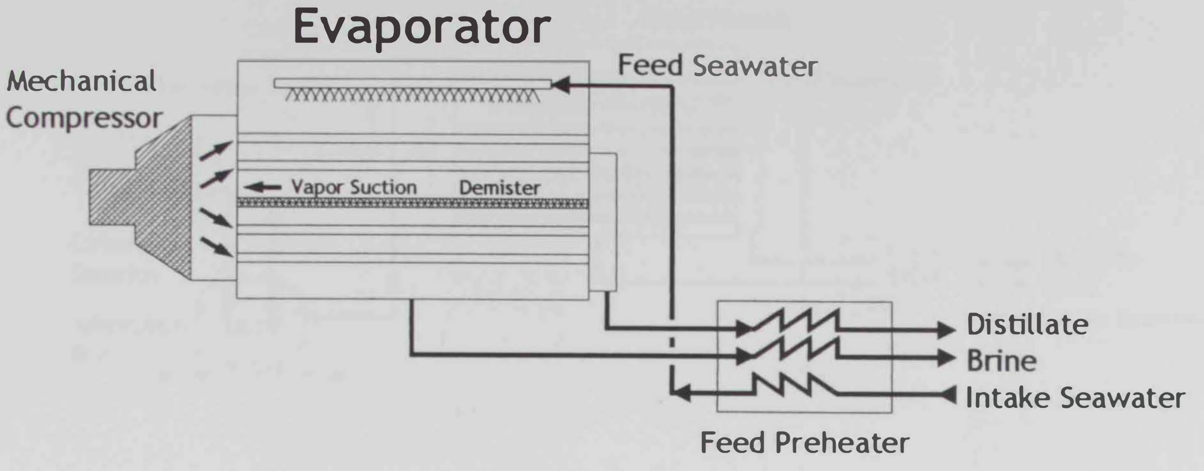


Figure 3-6: Mechanical vapor compression evaporator-desalination process.

3.3.3 Adsorption Vapor Compression (MED-ADS)

The process does not include moving parts, has a long life, and is vibration free. For these reasons, the adsorption-desorption heat pump started to attract attention due to the concern of replacing the traditional compressor-based systems, which utilize ozone harmful fluids. Applications of the adsorption-desorption heat pumps are found in air-conditioning and in ice making. The ADVC system is shown in Figure 3-7. The system includes the evaporator/condenser unit, two adsorption beds, and two heat exchangers. It is interesting to note that the evaporator and condenser form a single unit in this configuration, which replaces the individual condenser and evaporator in a conventional adsorption heat pump. Also, the liquid-to-liquid heat exchanger (HE1) preheats the feed seawater by condensed distillate product and rejected brine. The adsorber (bed II in Figure 3-7) plays the role of the bottom condenser in the TVC system. That is, this adsorber adsorbs or rejects the excess heat added to the system in the second adsorber [11].

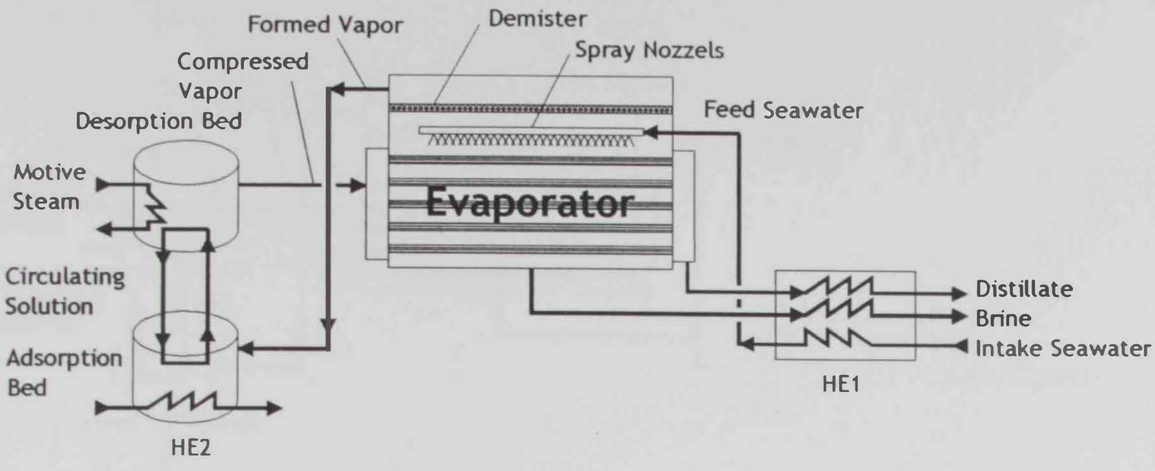


Figure 3-7: Single-effect evaporator driven by an adsorption heat pump.

3.3.4 Absorption Vapor Compression (ABVC)

The absorption heat pumps are used widely for air-conditioning purposes. There are many working fluids that are used in the absorption heat pump such as ammonia-water, LiBr-water [14], KOH and NaOH. The absorption heat pumps include four main components, which include the generator, the absorber, the evaporator, and the condenser. The pump receives heat in the evaporator and the generator and rejects heat in the condenser and the absorber. The pump can be configured for simultaneous heating and cooling purposes.

The ABVC system is shown in Figure 3-8. The system contains six elements, which include the generator, the absorber, the evaporator/condenser, and three heat exchangers. In this configuration, the evaporator desalination unit acts as the individual evaporator and condenser; this is different from conventional absorption heat pumps [11].

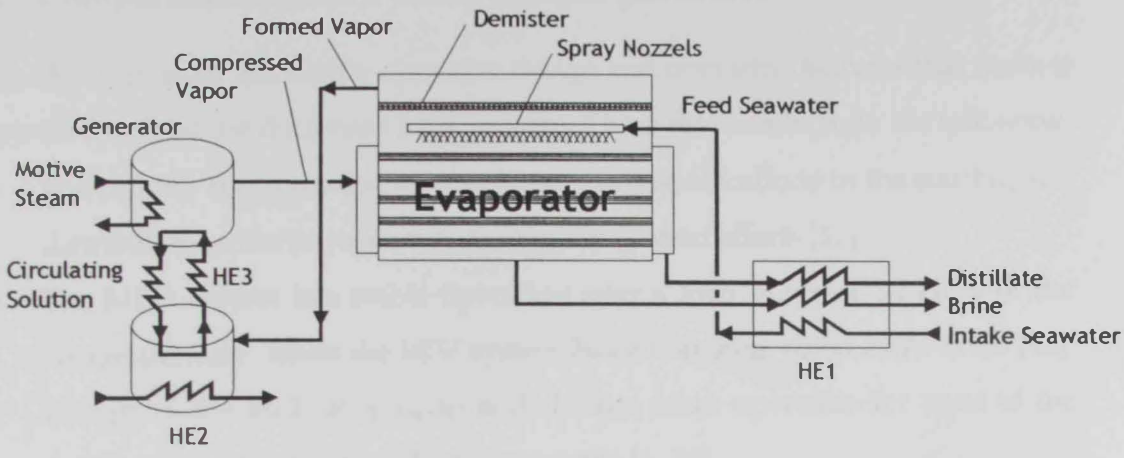


Figure 3-8: Single-effect evaporator driven by an absorption heat pump.

3.4 Advantages of the MED Process

3.4.1 Comparison between MED & other processes

The MED process has highly attractive design and operating features that make it competitive against the dominant MSF process. These features include the following:

- The process configuration allows for simple modifications in the routing and distribution of the brine stream among the system effects [11].
- The MED system has stable operation over a load range of 30-120% of the design capacity, while the MSF system has a narrower range of 70-110% [15]. This gives the MED system an added edge, since operation for most of the desalination units is tied with power plants [1, 16].
- The MED system is more efficient, from a thermodynamic and heat transfer point of view, than the predominant MSF system [17].
- The MED process can be operated at low temperatures and hence with much less scaling than MSF [16].
- The specific power consumption for running the pumps of MED is less than that of MSF. Rautenbach [18] shows 2 kWh/m^3 for MED and 5 kWh/ m^3 for MSF.
- For the same thermal performance ratio, the MED system has fewer effects than the MSF system: typically the conventional MED system has 12 effects, while the MSF system has 24 stages for a typical performance ratio of 9. Assuming a similar specific heat transfer area for both systems, the capital cost of the MED is lower than the MSF system because of the fewer effects, tube connections, and partition walls [1].
- The MED process is not yet fully developed. The unit capital costs ($/IGD) at various performance ratios are [16]:

Process Vs PR	8	10	12
	($/IGD)	($/IGD)	($/IGD)
MSF	7.68	8.7	10.08
MED-TVC	7.11	7.8	8.55

- The total production cost is $1.86/m^3 for the MSF process and $1.49/m^3 for the MED system [16].
- Al-Juwayhel et al. [11] indicated that the water cost in 1997 basis at the distribution point is $1.35/m^3 for MED-TVC process.

3.4.2 Advantages of the Low Temperature MED process

Low temperature distillation is the basis for a series of features, forming the core of the plants' highly economical operation [19]. These features are summarized below:

Low corrosion rates: The reduced corrosiveness of seawater, at the low operating temperature, permits safe and economic use of corrosion proof plastic materials and coatings both for piping and for vessel linings, as well as the use of aluminum for heat transfer tubing and vessel internals. Low maintenance and extended plant life (exceeding twenty-five years) result from the combination of the low corrosion rates and the use of a mild anti-scalant.

Flexibility: MED plants have short startup periods with little time loss for heating up. They also have excellent load following capabilities allowing for plant production that closely match both water demand and energy supply.

Thermodynamic efficiency: The use of generous heat transfer surfaces results in a reduction of heat fluxes and temperature differentials and therefore in an increase of thermal efficiencies. As a result, the evaporators can be operated with overall temperature differentials, including thermal driving forces, boiling point elevations and non-condensable gases and fouling factors, as low as 2 - 2.5 °C.

Minimal scaling rates: The operating temperatures are well below the saturation limits of problematic scalants found in seawater and most ground waters. The reduction of scales is to insignificant level, enables the plants to operate for long periods between cleanings. Low cost polyelectrolyte feed pretreatment is adequate. De-scaling is a simple procedure, consisting of mild acid re-circulation, using the plants own re-circulation pumps.

High purity distillate: An additional advantage is the high purity of the product water (usually less than 20 ppm and as low as 2-5 ppm for special applications). This allows the water to be used directly for industrial processes, such as in refineries, power stations, breweries etc, where boiler water quality is required or in municipal installations to reduce further the production costs by blending the high purity distillate with local brackish or poor quality water and satisfy the potable water standards.

Reliability: Experienced engineering, rugged construction and proven equipment combined with extremely low corrosion and scaling rates result in minimal maintenance, and lead to annual plant availability in excess of 95%.

Low energy costs: The low temperature operation enables the low temperature distillation units to utilize low grade, low cost sources of heat, which would otherwise be lost through being released into the environment in the form of stack gases or low pressure exhaust steam. The motive energy cost component for the desalination process is reduced to a minimum and consequently the water production costs are lower than any other seawater desalination system [10].

Cheap material of construction: The economy of using aluminum tubes for heat transfer as compared with copper alloy tubes which are essential for higher temperature plants enables the doubling the specific heat transfer area in the desalination plant for the same investment costs [1].
El-Dessouky et al. [20] showed that using high quality plastic as heat exchangers in the low temperature process would have a direct cost effect in the desalination industry. Jaakkola et al. [21] developed the AQUAMAX, which is a heat exchanger made of thin plastic film and welded to plastic envelope elements "Plastic bag". These AQUAMAX evaporators yield low potentiality of scale formation, corrosion and the cost of pretreatment.

3.5 Mathematical Modeling

Narmine et al. [22] presented a computer simulation model for steam jet ejectors. The model was developed by the application of the equations of continuity, momentum and energy to individual operation of nozzle, mixing chamber and diffuser, the effect of motive steam pressure, evaporator temperature, and pressure rise across the ejector.

Hamed and Ahmed [23] proposed a mathematical model for a four-effect thermal vapor compression process of 1500 m^3 per day. They predicted the impact of different process variables on the thermal behavior of the system.

El-Dessouky et al. [24] developed a mathematical model for a multiple effect evaporation (MED) desalination process. The influence of the important design and operating variables on the parameters controlling the cost of produced fresh water were analyzed. The model assumes the practical case of constant heat transfer areas for both the evaporators and feed pre-heaters in all effects. In addition, the model considered the impact of the vapor leak in the venting system, the variation in thermodynamic losses from one effect to another, the dependence of the physical properties of water on salinity and temperature, and the influence of non-condensable gases on the heat transfer coefficients in the evaporators and the feed pre-heaters. Results showed that the heat transfer coefficients increase with increasing the boiling temperature. Also, the heat transfer coefficient in the evaporator is always higher than that in the feed pre-heater at the same boiling temperature. The plant thermal performance ratio is nearly independent of the top brine temperature and strongly related to the number of effects.

Darwish and El-Dessouky [15] presented technical factors affecting the choice of a distillation system for desalting water. The thermal vapor compression process is compared with the predominant multi-stage flash (MSF) desalting system. They concluded that mechanical or thermal vapor-compression desalting systems are more cost effective when compared with directly boiler operated MSF systems. Thermal analysis of the multi-effect thermal vapor compression system is presented.

A simple analytical simulation of MED plant is presented by W. T. Hanbury [25] based on the alternative assumptions of a linear decreases in the boiling heat transfer coefficient and equal effect thermal loads from the second effect down. This results in a simple simulation giving a plant temperature profile far closer to reality.

El-Dessouky [26] reported a steady state mathematical model developed to analyze the single effect thermal vapor compression (TVC) desalination process. The model considered the effects of all thermodynamic losses on the thermal performance ratio, the specific heat transfer area and the specific flow rate of the cooling water. The model also takes into consideration the variation in the physical properties of the seawater with temperature and salt concentration, the effect of fouling factors and the presence of non-condensable gases on the heat transfer coefficients in the evaporator and the condenser. The relationships between the parameters that control the product water cost, and important parameters such as the motive steam pressure, the vapor compression ratio and the boiling temperature are presented.

El-Dessouky and Ettouney [1] proposed a thermal analysis for multiple-effect evaporation desalination systems. Several operating configurations were analyzed, including the parallel flow, the parallel/cross flow, and systems combined with thermal or mechanical vapor compression. Results were presented as a function of parameters controlling the unit product cost, which include the specific heat transfer area, the thermal performance ratio, the specific power consumption, the conversion ratio, and the specific flow rate of the cooling water.

4 CASE STUDY (UMM AL-NAR PLANT)

4.1 General Description

The seawater desalination plant of Umm Al-Nar in Abu Dhabi represents one of the latest examples of the application of the multiple effect process at low temperature with thermal vapor compressor for seawater desalination. The nominal capacity and thermal energy requirement of each unit is 3.5 MIGD and 85 ton of low-pressure steam per hour, respectively. The plant includes multi effect evaporation desalting system with steam jet ejector. This multi effect unit is of the horizontal tube falling film type with plain tubes, see Appendix B. The plant has six effects operating at different levels of temperature. The effects are arranged horizontally with satisfactory permanent access to all effects. The thermal vapor compression (TVC) technique is applied to two parallel sections of three effects each. Motive steam is admitted to the unit through two parallel thermo-vapor compressors one for each of the TVC section with three effects. The other part of vapor is allowed to pass through cells 4,5,6 and final condenser. Each unit produces a guaranteed rated output of 3.5 MIGD. Scale control is achieved by controlled dosing of polymer type additives (the anti-scalant (Sokalan) and anti foam (Belite M8)) into the seawater feed. The distiller is capable of operating at reduced output up to 50% of the rated design output. The design temperature of seawater input to the distillers is 33 °C and minimum temperature is 16 °C. the design seawater salinity is 52000 ppm, typical seawater intake analysis of Umm Al-Nar and its expected seasonal temperature variations are given in Appendix B.

The maximum design vapor temperature of the first and highest temperature, effect shall not exceed 65 °C, main operating parameters are given in Appendix B. The corrected performance ratio, with all evaporator tubes fouled to the design fouling factors is not less them 8.0 kg of product per 2300 kJ net heat input. The quality of distillate produced is guaranteed to have a total dissolved solids content not grater than 25 mg/l. Most of material used for the distiller is stainless steel; see Appendix B for distiller materials.

4.2 Process Description

The incoming seawater after being heated up in pre-heaters, is sprayed on the main tube bundles of each effect, where it is partially evaporated as it flows as thin films outside the tubes due to the heat released by condensing vapor flowing inside the tubes. The condensed steam represents the distillate production of each effect. Vapor generated by this evaporation goes to the next adjacent effect where it condenses inside the tubes forming again the distillate production of that effect. The stage process illustrate tube bundle and make up shower as shown in Appendix B.

The process is repeated in each of the six effects. Distillate and brine from each effect are cascaded to the next effect. In the last effect, brine is extracted and distillate is cascaded to the final condenser. Both brine and distillate streams are extracted by horizontal centrifugal pumps. Part of the production of the first effect is returned to the boiler by a condensate pump.

Part of the vapor leaving the third effect of the TVC section is re-compressed to the first effect of this section by the steam ejector; while the other part is used as heating steam for the top effect of 3 effects conventional MED system. Vapor generated in the 4th effect is introduced as heating vapor in the fifth effect and so on to the last effect.

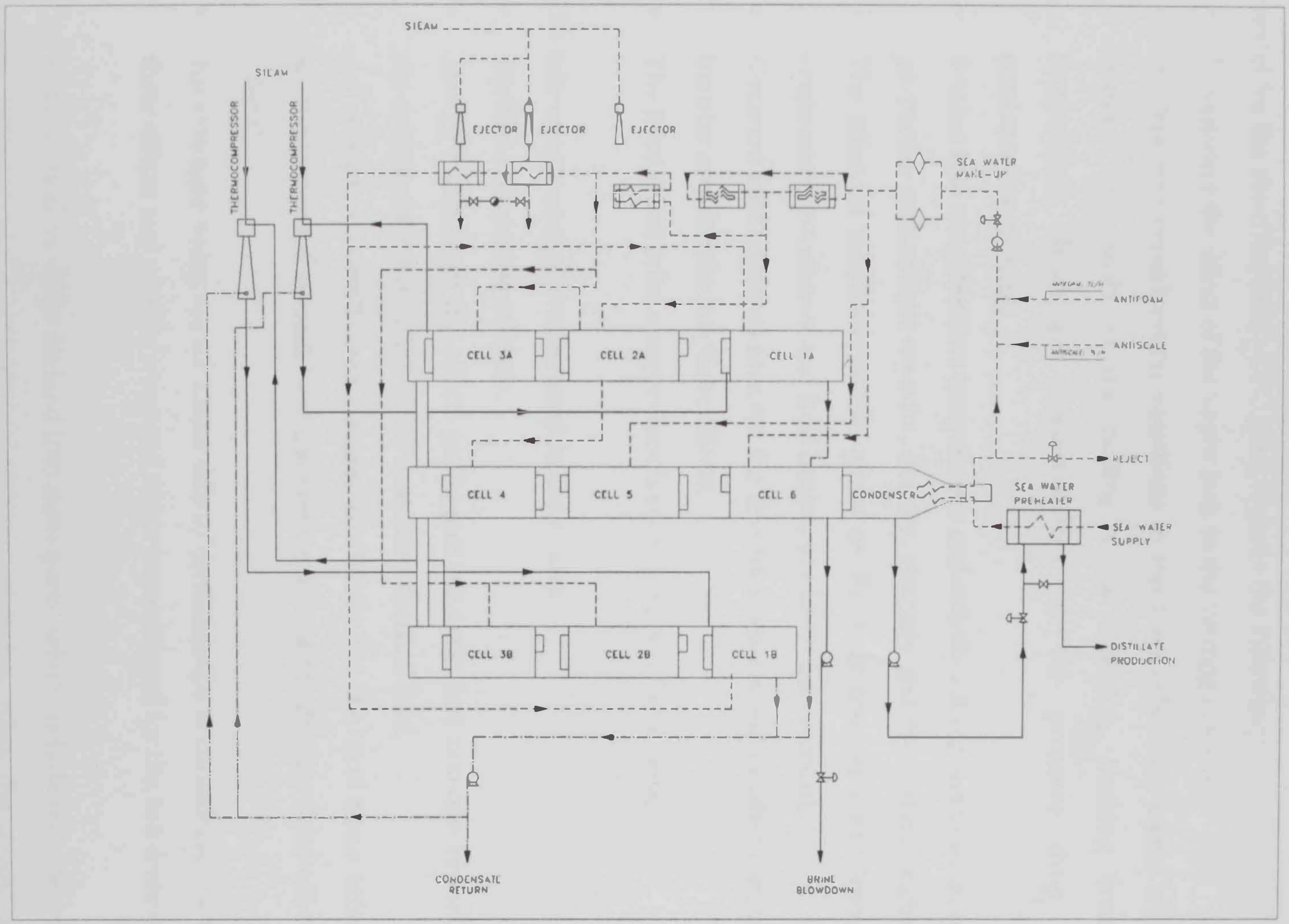


Figure 4-1: Process flow diagram for MED-TVC plant at Umm Al-Nar

5 MATHEMATICAL MODELING OF THE MED-TVC PROCESS

The steady state MED-TVC model includes a set of material and energy balances, heat transfer equations, and thermodynamic correlations. The **main features** of the model for the six-effect MED-TVC plant include the following:

- It considers the effect of the vapor leak to the venting system.
- It takes into consideration variations in the thermodynamic losses within the system. This includes the boiling point elevation, flashing boxes and temperature depression corresponding to the pressure drop during condensation process.
- It takes into consideration temperature and salinity effects on the water physical properties such as heat capacity, density, viscosity and thermal conductivity.
- The effect of non-condensable gases on the heat transfer coefficients in the evaporators, condenser and feed heaters are taken into account.
- Constant heat transfers area for the first three effects and another constant heat transfer area for the last three effects.
- The third effect brine temperature is controlled as a fixed value.

The following **assumptions** are used in the model:

- Distillate product is salt free.
- Energy losses from the effects to the surroundings and non-condensable gases are considered equal to 2% of the first effect thermal load.
- An averaged overall heat transfer coefficient for the first three effects and another averaged overall heat transfer coefficient for the last three effects (cold effects).
- An averaged value for ΔT losses due to condensation is considered for the first three effects and another averaged value is considered for the last three effects.

The mathematical model is divided into three parts, which include material balances, energy balances, and the heat transfer rate equations. Also, the model includes equations for the heat transfer coefficients, thermodynamic losses, and the physical properties of seawater.

The following section gives balance equations for the evaporators, down condenser, flash boxes, temperature profiles of effects, and steam jet ejector (thermo-compressor).

5.1 Material Balance Equations

The total mass and the salt balance in the system are given by

$$M_f = M_d + M_b \tag{5-1}$$

$$M_b = \frac{M_f * X_f}{X_b} \tag{5-2}$$

Where M is the mass flow rate, X is the salt concentration, and the subscripts d, f and b denote the product water, feed seawater and brine respectively.

5.1.1 Material balance for the first effect

The mass and salt balances for the first effect are

$$M_{f01} = M_{b1} + M_{v1} \tag{5-3}$$

$$M_{b1} = \frac{M_{f01} * X_{f01}}{X_{b1}} \tag{5-4}$$

$$M_s = M_{d1} + M_{cond} \tag{5-5}$$

Where cond is the condensate leavening the first effect to be pumped back to the boiler; v is the vapor formed in the evaporator, b is brine concentrated in the effect bottom, s is the discharged vapor from the thermal vapor compressor.

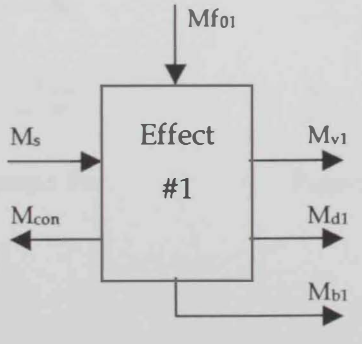


Figure 5-1: Material balance for the first effect

5.1.2 Material balance for the other effects

The mass and salt balances for other effects are

$$M_{f0n} + M_{d(n-1)} + M_{b(n-1)} = M_{bn} + M_{vn} \tag{5-6}$$

$$M_{bn} = \frac{M_{f01} * X_{f01} + M_{b(n-1)} * X_{b(n-1)}}{X_{bn}} \tag{5-7}$$

$$M_{vn} = M_{d(n+1)}' \tag{5-8}$$

Where n is the effect number.

The unit cell process equations for brine, distillate and vapor are

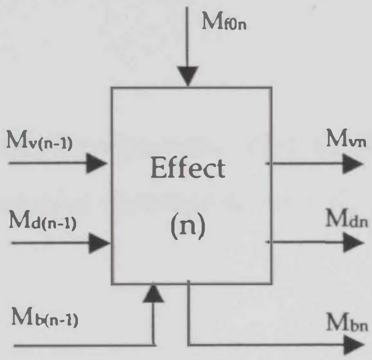


Figure 5-2: Material balance for effect (n = 2,3 and 6)

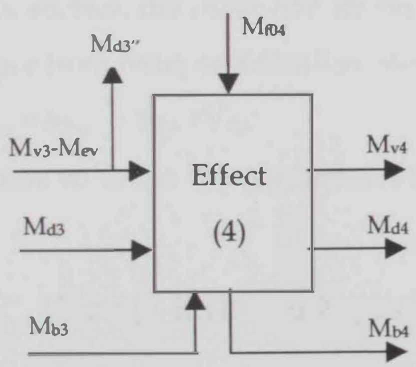


Figure 5-3: Material balance for effect (4)

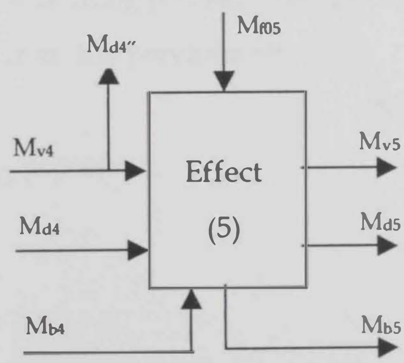


Figure 5-4: Material balance for effects (5)

Brine:

$$M_{b(n-1)} = L_{bn} + V_{bn} \tag{5-9}$$

$$L_{bn} = \frac{M_{b(n-1)} * X_{(n-1)}}{X_{Lbn}} \tag{5-10}$$

$M_{bn} = M_{bn'} + L_{bn}$ (5-11)

Where L_b is the remaining brine from the inlet brine stream due to flashing; V_b is the flashed vapor from the inlet brine stream, b_n' is the brine created from the boiling process inside the effect.

Distillate:

The distillate leaving each effect starting from effect two to effect n is a combination of two streams. One is the vapor condensate inside the tubes and the other one is the remaining distillate from the flashing process of the distillate coming from the previous effect.

$M_{d(n-1)} = L_{dn} + V_{dn}$ (5-12)

$M_{dn} = M_{dn}' + L_{dn}$ (5-13)

Where L_d is the remaining distillate from the inlet distillate stream due to flashing; V_d is the flashed vapor from the inlet distillate stream, d_n' is the condensed vapor inside the tubes.

Vapor:

From effect two to effect n, the total vapor generated in each effect is a combination of three streams. The first one is the vapor generated from the boiling process outside tube surface, the other two streams come from the flashing process, they are flashed vapor from brine and distillate streams cascaded from the previous effect.

$M_{vn} = M_{vn}' + V_{bn} + V_{dn}$ (5-14)

Where vn' is the vapor generated from the boiling process.

5.2 Energy Balance Equations

5.2.1 Energy balance for the final Condenser

Heat transfer between the condensing vapor and the feed seawater in the condenser can be written in terms of an overall heat transfer coefficient (U_{FC}), condenser heat transfer area (A_{FC}), and the logarithmic mean temperature difference $(LMTD)_{FC}$; thus:

$Q_{FC} = U_{FC} * A_{FC} * (LMTD)_{FC} = M_{vn} * \lambda_v$ (5-15)

Where $(LMTD)_{FC}$ is defined by

$$(\mathrm{LMTD})_{FC} = \frac{(T_f - T_{sw})}{\ln\left[\dfrac{(T_v - T_{sw})}{(T_v - T_{f1})}\right]} \tag{5-16}$$

The overall heat transfer coefficient based on the outside surface area U_o is related to the individual thermal resistance by the following expression.

$$\frac{1}{U_0} = \frac{1}{h_i}\frac{r_o}{ri} + R_{f_i}\frac{r_o}{r_i} + \frac{r_o \ln(r_o / r_i)}{k_w} + R_{f_o} + \frac{1}{h_o} \tag{5-17}$$

Where h is the heat transfer coefficient, R_f is the fouling resistance, k_w is the thermal conductivity of tube material, and r is the radius. The subscripts i and o refer to the inner and outer tube surface, respectively.

The inside tube heat transfer coefficient hi is calculated from the empirical formula developed by Wangnick [26] especially for desalination plants.

$$h_i = 3293.5 + T_f(84.24 - 0.1714T_f) - X_f\left(\frac{(8.471 + 0.1161X_f + 0.2716T_f)}{\left(\dfrac{\delta_i}{0.17272}\right)^{0.2}}\right)(0.656V)^{0.8}\left(\frac{\delta_i}{\delta_0}\right) \tag{5-18}$$

Where X_f is the salt concentration in ppm, T_f is the feed temperature, and δ_i and δ_o are the inside and outside the tube diameters, respectively.

Henning, et al. [26] developed the following equation to calculate the heat transfer coefficient during vapor condensation outside the tubes.

$$h_o = 0.725\left(\frac{k_l^{\,3}\rho_l(\rho_l - \rho_v)g\lambda_v}{(\delta_0 \mu \Delta T)^{0.25}}\right)C_1 C_2 \tag{5-19}$$

Where k_l is the thermal conductivity at bulk temperature (W/m°C), ρ is the density (kg/m³), g is the gravity (m/s²), μ is the dynamic viscosity at bulk temperature (Pa.s), subscripts l and v represent liquid and vapor phases respectively. The correction factors C_1 and C_2 consider the influence of the condensate dripping down and the presence of non-condensable gases. The size of the coefficients C_1 and C2 are given by the following equations.

$$C_1 = 1.23795 + 0.353808\, N - 0.0017035\, N^2 \tag{5-20}$$

$$C_2 = 1 - 34.313\, X_{nc} + 1226.8\, X_{nc}^2 - 14923\, X_{nc}^3 \tag{5-21}$$

Where X_{nc} is the weight percentage of the non-condensable gases and N is the number of tube rows inside the condenser. The total number of tubes is calculated by

$$N_t = \frac{4M_f}{\pi \delta_i \rho_l V} \qquad (5\text{-}22)$$

Where M_f is the feed flow rate. The following relation determines the number of tubes N

$$N = 0.564\sqrt{N_t} \qquad (5\text{-}23)$$

For the first loop in the calculation procedure the overall heat transfer coefficient in the final condenser, U_{FC}, is obtained from the correlation developed by El-Dessouky et al. [11]

$$U_{condenser} = 1.7194 + 3.2063E\text{-}2 * T_v - 1.5971E\text{-}5 * T_v^2 + 1.9918E\text{-}7 * T_v^3 \qquad (5\text{-}24)$$

Where U is in kW/m² °C and T_v in °C.

The thermal load of the condenser, Q_{FC} is given by

$$Q_{FC} = (M_{cw} + M_f) * Cp * (T_f - T_{sw}) \qquad (5\text{-}25)$$

In the above equation T_{sw} is the feed seawater temperature

Then,

$$M_{cw} = \frac{U_{FC} * A_{FC}}{\overline{Cp} * \left[\ln \frac{T_{cnd} - T_{sw}}{T_{cnd} - T_{f1}}\right]} - M_f \qquad (5\text{-}26)$$

Where subscript f1 denotes the outlet feed water from the final condenser, T_{cnd} is the condenser vapor temperature. T_{cnd} is less than the temperature of vapor released from the last effect by pressure drop during vapor condensation outside the condenser tubes expressed by the temperature drop ΔT_{cnd}. Thus,

$$T_{cnd} = T_{vn} - \Delta T_{cnd} \qquad (5\text{-}27)$$

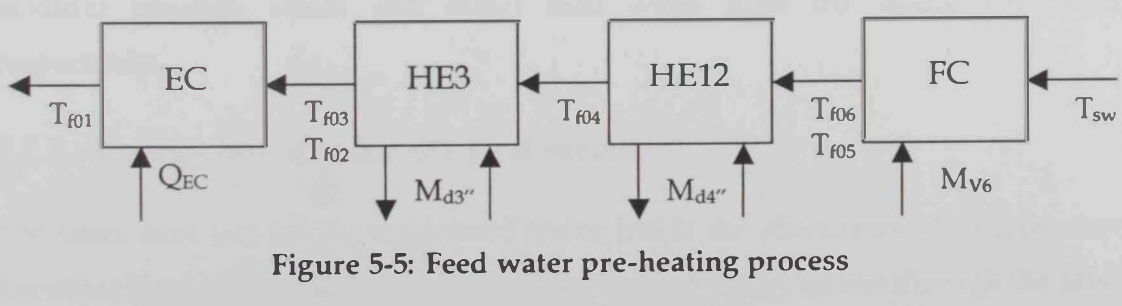


Figure 5-5: Feed water pre-heating process

5.2.2 Energy balance for the feed pre-heaters

Heat transfer between the condensing vapor from the fourth effect and feed seawater coming from the final condenser can be written by the following three equations:

$$Q_{HE12} = U_{HE12} * A_{HE12} * LMTD_{HE12}$$

$$= M_{d4}'' * \lambda_{v4}$$

$$= M_{f2} * Cp * (T_{f04} - T_{f06}) \qquad (5\text{-}28)$$

Where M_{d4}'' is the vapor condensed in pre-heater 12, U_{HE12} is the overall heat transfer coefficient and A_{HE12} is the heat transfer area for pre-heater 12, the subscript f2 denotes outlet feed water from pre-heater 12.

Similar equations can be applied for pre-heater 3:

$$Q_{HE3} = U_{HE3} * A_{HE3} * LMTD_{HE3}$$

$$= M_{d3}'' * \lambda_{v3}$$

$$= M_{f3} * Cp * (T_{f02} - T_{f04}) \qquad (5\text{-}29)$$

Where M_{d3}'' is the vapor condensed in pre-heater 3, U_{HE3} is the overall heat transfer coefficient and A_{HE3} is the heat transfer area for pre-heater 3, the subscript f3 denotes outlet feed water from pre-heater 3.

The last step for feed preheating is the ejector-driven condensers. Medium pressure steam is used as motive steam. Usually its flow and temperature are constant, so the thermal load of these ejector condensers could be considered constant

$$Q_{EC} = M_{MPS}'' * \lambda_{MPS}$$

$$= M_{f4} * Cp * (T_{f01} - T_{f02}) \qquad (5\text{-}30)$$

Where Q_{EC} is the thermal load for ejector condenser, MPS and f4 subscripts are medium pressure steam and outlet feed water from the ejector condenser, respectively.

5.2.3 Energy balance for the first effect

The latent heat (λs) for the condensed vapor inside the effect tubes (Ms) represents the effect thermal load and it is equal to the thermal energy access through the tubes area (A_{e1}). The driving force for the heat transfer process is the apparent temperature difference between the condensed vapor temperature (T_s) and brine temperature (T_1).

The overall heat transfer coefficient (U_{e1}) includes the convective local film coefficients and the conductive resistances of the tube wall and any scales.

$$Q_{e1} = M_s * (H_{mix} - h_{c1}) = U_{e1} * A_{e1} * (T_s - T_1) \quad (5\text{-}31)$$

The brine salinity can be calculated from the above equation resulting in

$$X_1 = \frac{M_{f01} * X_{f01} * (H_{v1} - Cp_{@T1} * T_1)}{M_{f01} * (H_v - Cp_{@Tf01} * T_{f01}) - M_s * \lambda_s} \quad (5\text{-}32)$$

Where Cp is the specific heat at constant pressure; T_1 is the brine temperature; H is the enthalpy for saturated vapor; λ_s is the latent heat for vapor-steam mixture from the thermal vapor compressor; and the subscript s denotes the compressed vapor.

5.2.4 Energy balance for the other effects

In effects 2 to n, the condensed vapor temperature is decreased due to the decrease in pressure as we move downwards in the plant. The condensate temperature was obtained previously. The thermal load of effects 2-n can be written as:

$$Q_{en} = M_{v(n-1)} * \lambda_c = U_{en} * A_{en} * (T_{cn} - T_n) \quad (5\text{-}33)$$

The boiling process in all effects is similar. The salinity of the brine due to boiling is given by

$$X_n' = \frac{M_{f0n} * X_{f0n} * (H_{vn} - Cp_{@Tn} * T_n)}{M_{f0n} * (H_v - Cp_{@Tf0n} * T_{f0n}) - M_{d(n-1)} * \lambda v_{(n-1)}} \quad (5\text{-}34)$$

Flashing process

The remaining brine flow rate L_{bn} and salinity X_{Lbn} are given by

$$L_{bn} = \frac{M_{b(n-1)} * (Cp_{@T(n-1)} * T_{(n-1)} - H_{vn})}{Cp_{@TLbn} * T_{Lbn} - H_{vn}} \quad (5\text{-}35)$$

and

$$X_{Lbn} = \frac{X_{(n-1)} * (Cp_{@TLbn} * T_{Lbn} - H_{vn})}{[Cp_{@T(n-1)} * T_{(n-1)} - H_{vn}]} \quad (5\text{-}36)$$

The overall brine salinity of each effect is expressed by

$$X_n = \frac{M_{bn}' * X_n' + L_{bn} * X_{Lbn}}{M_{bn}} \quad (5\text{-}37)$$

The remaining distillate L_{d2} is given by

$$L_{dn} = \frac{M_{d(n-1)} * (h_{Ld(n-1)} - H_{vn})}{h_{Ldn} - H_{vn}} \tag{5-38}$$

5.3 Equations for Temperature Profiles

5.3.1 Temperature profile of the feed seawater

Seawater feed is preheated in four steps as was shown in Figure 5.3. The first step is in the final condenser. The outlet temperature is expressed by

$$T_{f1} = T_{v6} - \left[\frac{(T_{v6} - T_{sw})}{\exp\left[\frac{U_{FC} * A_{FC}}{(M_{cw} + M_f) * \overline{CP}} \right]} \right] \tag{5-39}$$

Where T_{f1} is the feed water outlet temperature from the final condenser. The cooling water temperature T_{cw}, sixth effect feed water temperature T_{f06} and fifth effect feed water temperature T_{f05} are equal to this temperature.

The second step in feed water heating take place in pre-heaters 1 and 2

$$T_{f2} = T_{v4} - \left[\frac{(T_{v4} - T_{f1})}{\exp\left[\frac{U_{HE12} * A_{HE12}}{M_{f2} * CP} \right]} \right] \tag{5-40}$$

Where T_{f2} is the feed water outlet temperature from pre-heaters 1 and 2. The fourth effect feed water temperature T_{f04} is equal to this temperature. U_{HE12} is the overall heat transfer coefficient for pre-heaters 1 and 2.

The third step in feed water heating take place in pre-heater 3

$$T_{f3} = T_{v3} - \left[\frac{(T_{v3} - T_{f2})}{\exp\left[\frac{U_{HE3} * A_{HE3}}{M_{f3} * CP} \right]} \right] \tag{5-41}$$

Where T_{f3} is the feed water outlet temperature from pre-heater 3. The second and third effects feed water temperature, T_{f02}, T_{f03} respectively, are equal to this temperature. U_{HE3} is the overall heat transfer coefficient in the pre-heater 3.

The fourth step takes place in the ejector condensers. The outlet temperature feed to first effect T_{f01} is:

$$T_{f01} = T_{f3} + \frac{Q_{EC}}{M_{f4} * CP} \tag{5-42}$$

5.3.2 Temperature profile for the first effect

The brine temperature in the first effect is:

$$T_1 = T_s - \frac{M_s * (H_{mix} - h_{c1})}{U_e * A_e} \tag{5-43}$$

The saturation temperature of the formed vapor is given by

$$T_{v1} = T_1 - (BPE_{mean} + \Delta T_m + \Delta T_h) \tag{5-44}$$

The boiling point elevation $(BPE)_{mean}$, at a given pressure, results in an increase in the boiling temperature due to the effect of dissolved salts. In horizontal tube falling film evaporators, the mean BPE is used to determine the vapor temperature. This is due to the extreme change in salinity for falling film evaporation. The BPE is calculated from the following empirical formula [12]:

$$BPE = \frac{x * T^2}{13832}\left[\begin{array}{l} 1 + 1.373 * 10^{-3} * T - 2.72 * 10^{-3} * \sqrt{x} * T + 17.86 * x - \\ 1.52 * 10^{-2} * x * T * \left(\dfrac{T - 225.9}{T - 236}\right) - \dfrac{2583 * x * (1 - x)}{T} \end{array}\right] \tag{5-45}$$

Where

x = salt concentration, wt. fraction

T = Temperature, K

The above equation is valid over the following ranges: 20000 < X < 160000 ppm, 20 < T < 180 °C.

ΔT_m expresses temperature decrease due to vapor pressure drop through demister. In general, the pressure loss during the vapor flow through a wire mesh pad, which is widely used as the mist eliminator in water desalination industry, is relatively small because of the high void fraction of these pads.

The boiling point rise caused by the hydrostatic head ΔT_h has a negligible effect in horizontal falling film evaporators.

5.3.3 Temperature profile for the other effects

The brine temperature from the boiling process is given by

$$T_n' = T_{c(n-1)} - \frac{M_{dn} * \lambda_{cn}}{U_e * A_e} \quad (5\text{-}46)$$

Where T_{bn}' is the brine temperature generated from the boiling process on the tube outside surface, T_c is the condensation temperature of vapor inside the evaporator tubes and ΔT_c compensates for the pressure drop due to condensation.

The condensation temperature of vapor inside the tube bundle of the second effect T_{c1} is less than the boiling temperature in the first effect T_1 by the mean boiling point elevation $(BPE_{mean})_1$ and the saturation temperature depressions associated with the pressure loss during the vapor flow in the demister $(\Delta T_m)_1$, vapor transmission lines $(\Delta T_t)_1$ and vapor condensation inside the horizontal tubes $(\Delta T_c)_1$. Thus,

$$T_{cn} = T_n - (BPE_{mean} + \Delta T_m + \Delta T_t + \Delta T_c)_1 \quad (5\text{-}47)$$

The pressure drop during the vapor condensation inside the evaporator tubes is the sum of the frictional, gravitational and acceleration components.

The final brine temperature in each effect is expressed by the brine temperature, which is generated from the boiling process and the remaining brine temperature from the flash process, can be calculated by the following equation:

$$T_n = \frac{M_{bn} * Cp * T_{bn} * + L_{bn} * Cp * T_{Lbn}}{(M_{bn} + L_{bn}) * Cp} \quad (5\text{-}48)$$

5.4 Equations for Steam Jet Ejector

The general material and energy balance equations for the steam jet ejector are:

$$M_{ms} + M_{ev} = M_s \quad (5\text{-}49)$$

$$M_{ms} * H_{ms} + M_{ev} * H_{ev} = M_s * H_s \quad (5\text{-}50)$$

The most important and critical step in modeling the thermal vapor compressor in this type of desalination process is the evaluation of the performance of the steam jet ejector. The main data required for analyzing the steam jet ejector is the determination of the mass of motive steam required per unit mass of the entrained vapor (Ra), given the pressure of the motive steam (P_{ms}), discharge pressure (P_s) and the suction pressure (P_{ev}). There is very limited number of methods available in the literature analyzing the steam jet ejector.

The method reported by El-Dessouky et al. [15] was adopted in this work for the design of the steam ejector. The first equation in the model gives the compression ratio, Cr, of the compressed and the entrained vapors,

$$C_r = \frac{P_s}{P_v} \tag{5-51}$$

Where P_s and P_v are the pressures of compressed and entrained vapor. The compression ratio can be varied over a range from 1.8 to 5. Specification of its value together with the pressure of the entrained vapor is used to determine the pressure of the compressed vapor, which is assumed to be at saturation conditions. The entrainment ratio (Ra) is defined as the mass of motive steam (M_{ms}) per unit mass of sucked vapor (M_{ev}). The ratio is obtained from the following relation:

$$Ra = 0.296 \frac{P_s^{1.19}}{P_v^{1.04}} \left(\frac{P_{ms}}{P_v} \right)^{0.015} \frac{(PCF)}{(TCF)} \tag{5-52}$$

Where PCF and TCF are pressure and temperature correction factors expressed as

$$PCF = 3 \times 10^{-7} (P_{ms})^2 - 0.0009 (P_{ms}) + 1.6101 \tag{5-53}$$

$$TCF = 2 \times 10^{-8} (T_v)^2 - 0.0006 (T_v) + 1.0047 \tag{5-54}$$

$$M_{ev} = \frac{M_{ms}}{Ra} \tag{5-55}$$

Where P_m, P_s and P_v are the pressure of the motive steam, compressed vapor, and entrained vapor, respectively. PCF is the motive steam pressure correction factor and TCF is the entrained vapor temperature correction factor. The pressures, temperature and flow rates in the above equations are in kPa, °C, and kg/s, respectively. The above correlation is valid in the following ranges:

$Ra \leq 5$

$500 \geq T_m > 10\ °C$

$3500 \geq P_{ms} \geq 100\ \ kPa$

$\dfrac{P_s}{P_v} \geq 1.81$

5.5 Model Solution

The mathematical model was coded using Excel and the developed computer program was called the MEDNAR simulator. It is designed for the MED-TVC plant at Umm-Al-Nar.

This plant is split in two parts, hot effects (effect 1,effect2 and effect3) using MED-TVC process and cold effects (effect 4, effect 5, effect6) using MED process.

The following set of specifications is used for the solution procedure:

- Seawater temperature, T_{cw} = 33 °C.
- Seawater salinity, X_f = 52000 ppm.
- Number of effects, n = 6.
- Evaporator heat transfer area for hot effects, A_{HOT} = 13099 m².
- Evaporator heat transfer area for cold effects, A_{COLD}= 4453 m².
- Condensate vapor temperature from third effect, T_{c3} = 54 °C.
- Last stage vapor temperature, T_{v6} = 43 °C.
- Condensation temperature losses are:

    ΔT_c for hot effects = 0.415 °C.

    ΔT_c for cold effects = 0.68 °C.
- Feed pre-heaters (1 and 2) area is 578 m² each and pre-heater 3 area is 693 m².
- Down condenser Area, A_{cnd} = 2874 m².
- Motive steam pressure, P_{ms} = 239 kPa.
- Motive steam flow, M_{ms} = 10.6 kg/sec.
- Ejector Condenser thermal load, Q_{EC} = 2475 kW.

The solution procedure is sequential and proceeds as listed below:

5.5.1 Feed water pre-heating

1. The seawater temperature is given.
2. Guess the cooling water flow rate, Mcw.

3. Guess the vapor temperature in effect 6.
4. Calculate U_{FC} using equations (5-17 to 5-23).
5. Calculate the feed temperature T_{f1} using equation (5-39).
6. Guess vapor temperature in effect 4, Tv4.
7. Calculate U_{HE12} using equation (5-17 to 5-23).
8. Calculate the outlet feed temperature from pre-heater 12, T_{f2}, using equation (5-40).
9. Obtain U_{HE3} from equation (5-17 to 5-23), where T_{c3} is input data.
10. Calculate outlet feed temperature from pre-heater 3, T_{f3}, using equation (5-41).
11. Calculate the outlet feed temperature from Ejector Condenser, T_{f01}, using equation (5-42).

5.5.2 Hot effects

12. Guess the overall heat transfer coefficient for hot effects, U_{HOT}.
13. Calculate the vapor condensation temperature, $T_{c(n-1)}$, using equation (5-27).
14. Calculate the effect thermal load, Q_{in}, knowing the condensed vapor flow and temperature using equation (5-33).
15. Use NCG as reduction percentages from the thermal load Q_{in}.
16. Get the boiling temperature T_n from equation (5-46).
17. Guess the brine salinity generated from boiling process, X_n'.
18. Obtain mean salinity X_{mean}, knowing the seawater salinity.
19. Calculate BPE_{mean} from equation (5-45).
20. Calculate the vapor temperature using equations (5-47 and 5-27).
21. Obtain the specific heat, Cp for brine and feed using the correlation from Appendix B.
22. Calculate the brine salinity from the boiling process, X_n', using equation (5-34).
23. Go back to step 17 and re-guess X_n' using the calculated X_n' from the previous step. Do the loop till reaching acceptable accuracy.
24. Calculate the brine and vapor flow rates generated from the boiling process, M_{bn}' and M_{vn}' respectively, using general equations (5-2 and 5-3) respectively.
25. Calculate the salinity and flow rate for the remaining brine generated from the flashing process. X_{Lbn} and L_{bn} respectively, using equations (5-36 and 5-35).
26. Calculate the brine flow rate, M_{bn}, and the salinity, X_n, for overall unit operation in effect using equations (5-11 and 5-37) respectively.
27. Calculate the remaining distillate L_{dn} from equation (5-38).

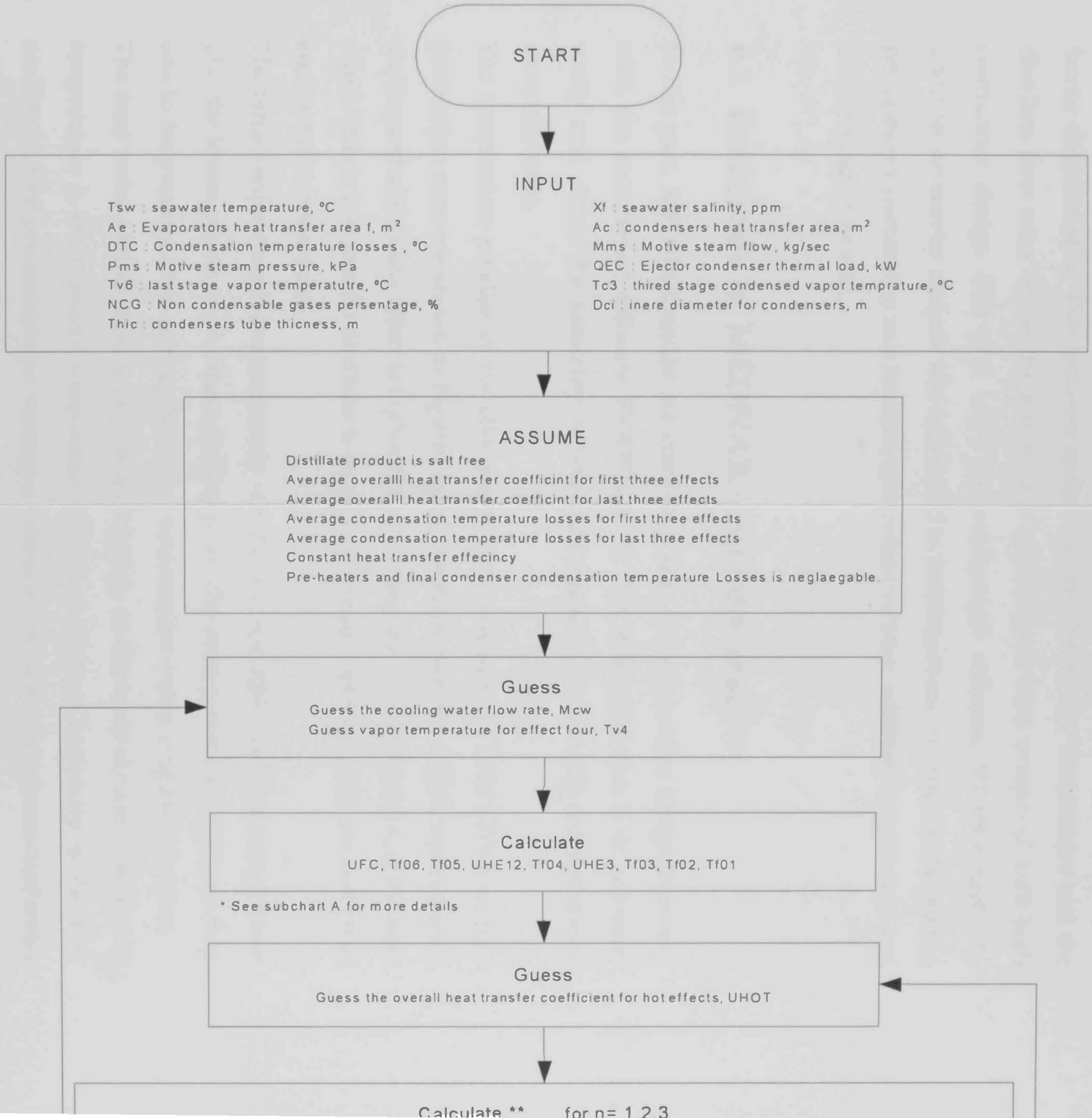
28. Obtain the effect unit operation distillate, M_{dn}, and vapor, M_{vn}, using equations (5-13 and 5-14) respectively.
29. Repeat steps 13 to 28 for second and third effects.
30. Go back to step 12 and re-guess U_{HOT}. Do the loop till calculated condenses vapor temperature, T_{c3}, equals T_{c3} set value.

5.5.3 Cold Effects

31. Guess the overall heat transfer coefficient for cold effects, U_{COLD}.
32. Repeat steps 13 to 28 for effects 4, 5 and 6.
33. Calculate the final condenser thermal load knowing the condensed vapor temperature and flow rate in condenser (shell side), $Q_{FC(out)}$, and compare it with calculated thermal load ,Q_{FC}, knowing overall heat transfer coefficient ,U_{FC}, and heat transfer area ,A_{FC}, using equations (5-15 and 5-25) respectively.
34. Go back to step 31 and re-guess U_{cold}. Stop when $Q_{FC(out)} = Q_{FC}$.
35. Calculate M_{cw} using equation (5-26) and re-use it in step 2. Use vapor temperature of effect 4, T_{v4}, and effect 6, T_{v6}, in steps 6 and 3 respectively. Do the loop till M_{cw} calculated equal M_{cw} guessed.

Notes:

A. There is no flashing in first effect, skip steps 25 to 28.
B. There is no ΔT_c in the first effect, skip step 13.
C. NCG percentage in step 15 is only for the first effect.
D. In the first loop for feed water heating, use the U correlation to find a good guess for U_{FC}, U_{HE12} and U_{HE3}.
E. In effect 3, the vapor temperature, T_{v3} = 54 °C, is controlled by the final condenser cooling water.
F. For the steam jet ejector, get the pressure correction factor from equation (5-51) and the temperature correction factor from equation (5-52). The compressed vapor pressure, P_s, is obtained from the specification of the compression ratio, Cr, and the entrained vapor pressure, P_{ev}. Calculating the saturation temperature, Ts, at the corresponding vapor pressure, Ps, from the steam tables, follows this. Finally, Get the entertained ratio, Ra, from equation (5-50) and the sucked vapor from equation (5-53).



START
INPUT
Tsw : seawater temperature, °C
Ae : Evaporators heat transfer area f, m²
DTC : Condensation temperature losses , °C
Pms : Motive steam pressure, kPa
Tv6 : last stage vapor temperatutre, °C
NCG : Non condensable gases persentage, %
Thic : condensers tube thicness, m
Xf : seawater salinity, ppm
Ac : condensers heat transfer area, m²
Mms : Motive steam flow, kg/sec
QEC : Ejector condenser thermal load, kW
Tc3 : thired stage condensed vapor temprature, °C
Dci : inere diameter for condensers, m
ASSUME
Distillate product is salt free
Average overalll heat transfer coefficint for first three effects
Average overalll heat transfer coefficint for last three effects
Average condensation temperature losses for first three effects
Average condensation temperature losses for last three effects
Constant heat transfer effecincy
Pre-heaters and final condenser condensation temperature Losses is neglaegable.
Guess
Guess the cooling water flow rate, Mcw
Guess vapor temperature for effect four, Tv4
Calculate
UFC, Tf06, Tf05, UHE12, Tf04, UHE3, Tf03, Tf02, Tf01
* See subchart A for more details
Guess
Guess the overall heat transfer coefficient for hot effects, UHOT
Calculate ** for n= 1,2,3

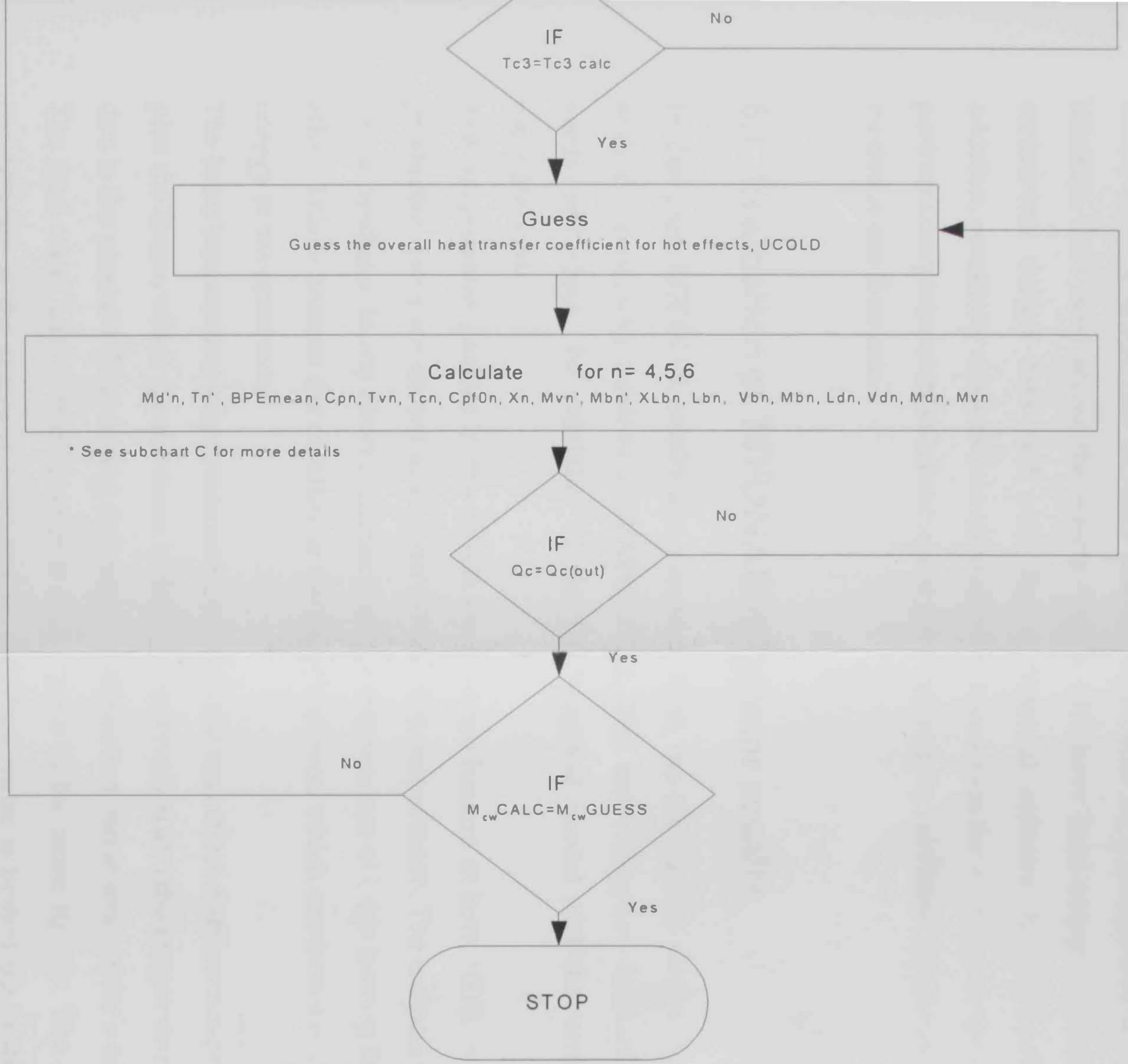


Figure 5-6: Overview of the flow chart for the computer program MEDNAR

6 RESULTS AND DISCUSSIONS

The simulation results of the developed model MEDNAR for the MED-TVC plants at Umm Al Nar power station are presented and discussed in this chapter. The MEDNAR computer program is validated through analysis of the effect and pre-heater characteristics. This includes analysis of profiles for the temperature and the distillate flow rates across the effects. The results have been compared with both contractors design data and with the commercial software EVAPOLAND. In addition, sensitivity analysis was conducted to investigate the variation in the system performance parameters as a function of seawater salinity, seawater temperature and motive steam flow rate.

6.1 Evaluation of MEDNAR simulator results

In this part, MEDNAR results are compared with the contractors design data and with the commercial software EVAPOLAND. The main objective is to calibrate, verify and check the reliability of the mathematical model and the developed computer code.

The temperature profiles in the effects and the pre-heaters at both 100% and 50% production rates are shown in Figures 6-3 and 6-4, respectively. The temperature of vapor produced in any effect is higher than the temperature of brine leaving the next effect. This is because the distillate is the energy stream, which condenses and gives energy to the brine side.

The brine temperature is approximately equal to the next effect distillate temperature plus the losses, which are the mean boiling point elevation and the temperature drop due to the pressure drop during the vapor condensation inside evaporator's tubes.

The feed water temperature profile is also shown in the same figures. The lowest temperature is the seawater temperature. The feed water is heated up in the final condenser, feed pre-heaters and venting condensers. The low temperature feed water streams are feed to last stages. This is due to lower temperature needed for boiling process.

Figures 6-5 and 6-6 display the feed water temperature profiles at 100% and 50% production rates, respectively. In these figures, the design data are compared with the MEDNAR runs. The maximum deviation (1.2%) at 100% production rate was found in the final condenser. At 50% production rate, the largest net deviation (1.7%) was found in pre-heater 12.

Figures 6-7 and 6-8 illustrate the vapor condensation temperature profiles at 100% and 50% production rates, respectively. In these figures, MEDNAR runs are compared with the design data and EVAPOLAND runs. The maximum deviation (1.6%) is observed at the last effect condensation temperature T_{c6} between MEDNAR and EVAOLAND runs.

The brine temperature profiles at 100% and 50% production rates are shown in Figures 6-9 and 6-10, respectively. The maximum deviation (1.6%) is found between MEDNAR run and design data at 100% production rate in the last stage brine temperature T_6. This difference may be attributed to error accumulation based on different assumptions.

In general, the results for the different models runs are very close; see tables 1 and 2 for 100% and 50% production rates results respectively. The tables display the final brine and final distillate flow rates. The maximum brine flow rate deviation is 0.2% compared with design data at 100% production rate run. At 50% production rate run, the distillate flow rate deviation is 3.2% compared with EVAPOLAND run.

The largest deviation was observed in the cooling water flow rates. Compared with design data, the results show 18.9% deviation at 100% production rate and 13% deviation at 50% production rate.

This difference is may be attributed to many reasons. The first reason is the MEDNAR simulator methodology. The program is based on a number of loops. The main loop checks the cooling water flow rate. The deviation is due to the accumulation of all the errors resulted from the different assumptions.

The other reason is the condensers overall heat transfer coefficient, U_{cond}. Henning and Wangnick [26] show different equations calculating the local heat transfer coefficients for inside and outside condensers tube. The results showed different overall heat transfer coefficient valves with different methods. The U value is affects the final condenser thermal load directly and indirectly through the deviation in the feed temperature profile which finally affects the cooling water flow rate. Another reason could be the designer is using different fouling factors and empirical values gained by experience.

Practically speaking, this difference could be accepted if we are considering the feed water flow rate in our calculations in addition to the cooling water flow rate. In this case, the total seawater intake flow rate deviation will reduce from 18.9% to 11% at full production rate and from 13% to 2% at half production rate.

Table 6-1: Comparison of Design data, EVAPOLUND and MEDNAR at 100% production rate

Variable	Design Data	Δ1	EVAPOLUND	Δ2	MEDNAR Simulator
A- Feed water temperature profile °C (100%)					
Tsw	33.0	0.0			33.0
Tf06	40.5	1.2			41.0
Tf05	40.5	1.2			41.0
Tf04	48.3	1.9			49.2
Tf03	52.0	2.3			53.2
Tf02	52.0	2.3			53.2
Tf01	55.9	2.2			57.1
B- Vapor condensation temp. profile °C (100%)					
Tc1	65.0	0.0	65.0	0.0	65.0
Tc2	61.3	0.1	61.8	0.7	61.4
Tc3	57.7	0.2	58.2	0.7	57.8
Tc4	54.0	0.0	54.7	1.3	54.0
Tc5	50.3	1.1	50.5	1.5	49.7
Tc6	46.7	1.0	47.0	1.6	46.2
Tdistillate	43.0	0.0	43.0	0.0	43.0
C- Brine temperature profile °C (100%)					
T1	62.5	0.1	62.7	0.2	62.6
T2	58.9	0.2	59.1	0.1	59.0
T3	55.2	0.5	55.6	0.3	55.5
T4	51.7	1.0	51.4	0.5	51.1
T5	48.1	1.0	47.8	0.5	47.5
T6	44.4	1.6	43.8	0.4	43.6
C- Other variables (100%)					
Mcw, kg/s	929.4	18.9			753.7
Mdistillate, kg/s	184.2	0.1	184.4	0.0	184.3
Mbrine, kg/s	456.7	0.2	455.6	0.0	455.7
Tbrine, °C	44.4	1.7	43.8	0.4	43.6
Xbrine, 1000*ppm	73.0	0.0	73.0	0.0	73.0

Where:

Δ1 = 100 * (design data – MEDNAR result)/ design data

Δ2 = 100 * (EVAPOLUND result – MEDNAR result)/ EVAPOLUND result

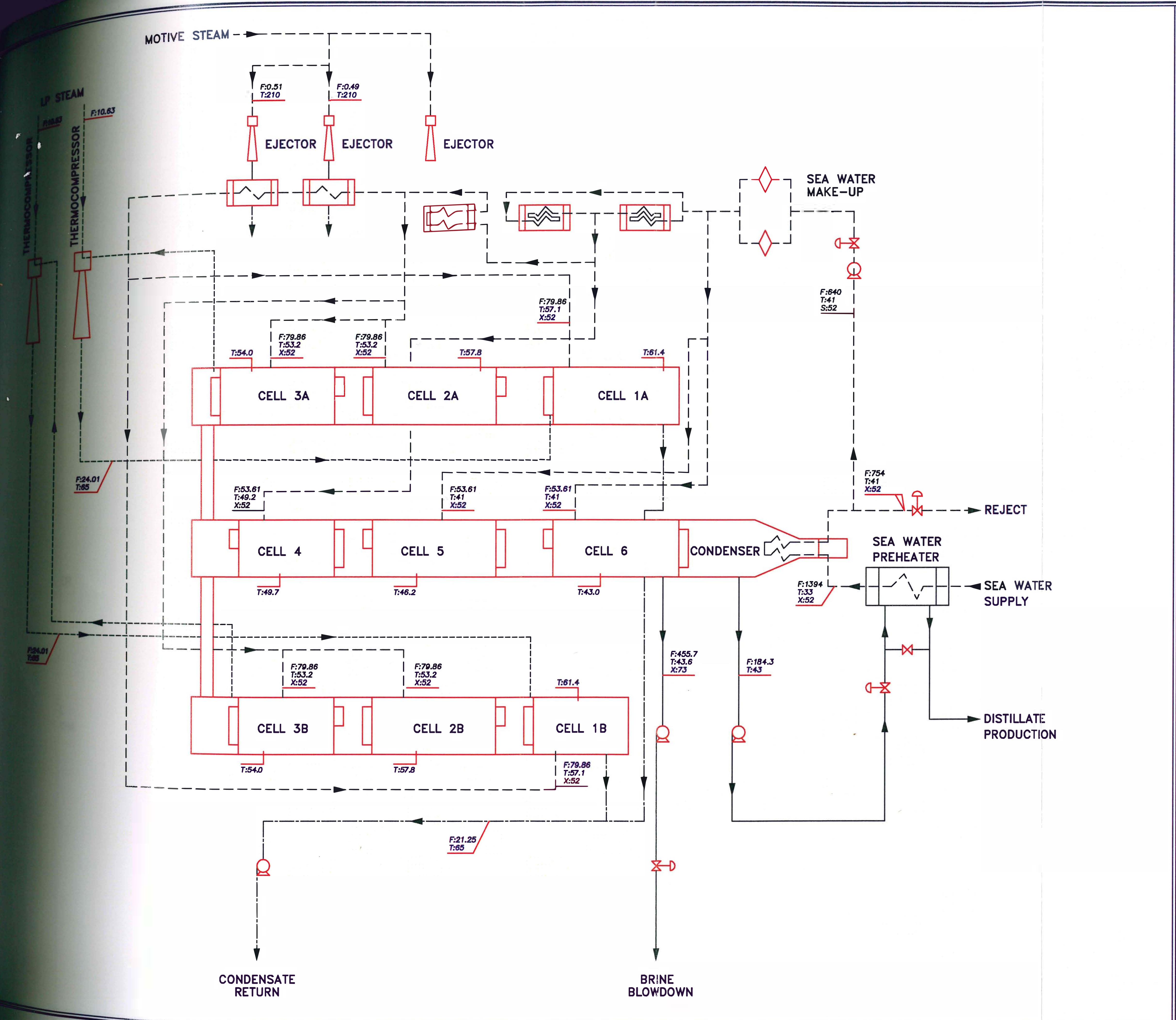
Table 6-2: Comparison of Design data, EVAPOLUND and MEDNAR at 50% production rate

Variable	Design Data	Δ1	EVAPOLUND Run	Δ2	MEDNAR Simulator Run
A- Feed water temperature profile °C (50%)					
Tsw	33.7	0.0			33.7
Tf06	43.4	0.8			43.8
Tf05	43.4	0.8			43.8
Tf04	49.0	2.5			50.2
Tf03	52.0	2.8			53.5
Tf02	52.0	2.8			53.5
Tf01	55.9	2.7			57.4
B- Vapor condensation tempe. profile °C (50%)					
Tc1	62.2	0.0	62.2	0.0	62.2
Tc2	59.5	0.0	59.9	0.7	59.5
Tc3	56.7	0.3	57.3	0.8	56.9
Tc4	54.0	0.0	54.7	1.3	54.0
Tc5	51.0	1.0	51.3	1.6	50.5
Tc6	48.0	0.8	48.4	1.6	47.6
Tdistillate	45.0	0.0	45.0	0.0	45.0
C- Brine temperature profile °C (50%)					
T1	60.7	0.2	60.6	0.0	60.6
T2	57.9	0.1	58.0	0.0	58.0
T3	55.4	0.0	55.4	0.1	55.4
T4	52.4	1.0	52.0	0.3	51.8
T5	49.4	0.9	49.1	0.4	48.9
T6	46.4	1.6	45.7	0.2	45.6
C- Other variables (50%)					
Mcw, kg/s	111.4	13.0			125.9
Mdistillate, kg/s	92.1	2.3	91.2	3.2	94.2
Mbrine, kg/s	548.8	0.5	548.8	0.5	545.8
Tbrine, °C	46.4	1.7	45.7	0.2	45.6
Xbrine, 1000*ppm	60.6	0.6	60.6	0.6	61.0

Where:

Δ1 = 100 * (design data - MEDNAR result)/ design data

Δ2 = 100 * (EVAPOLUND result - MEDNAR result)/ EVAPOLUND result

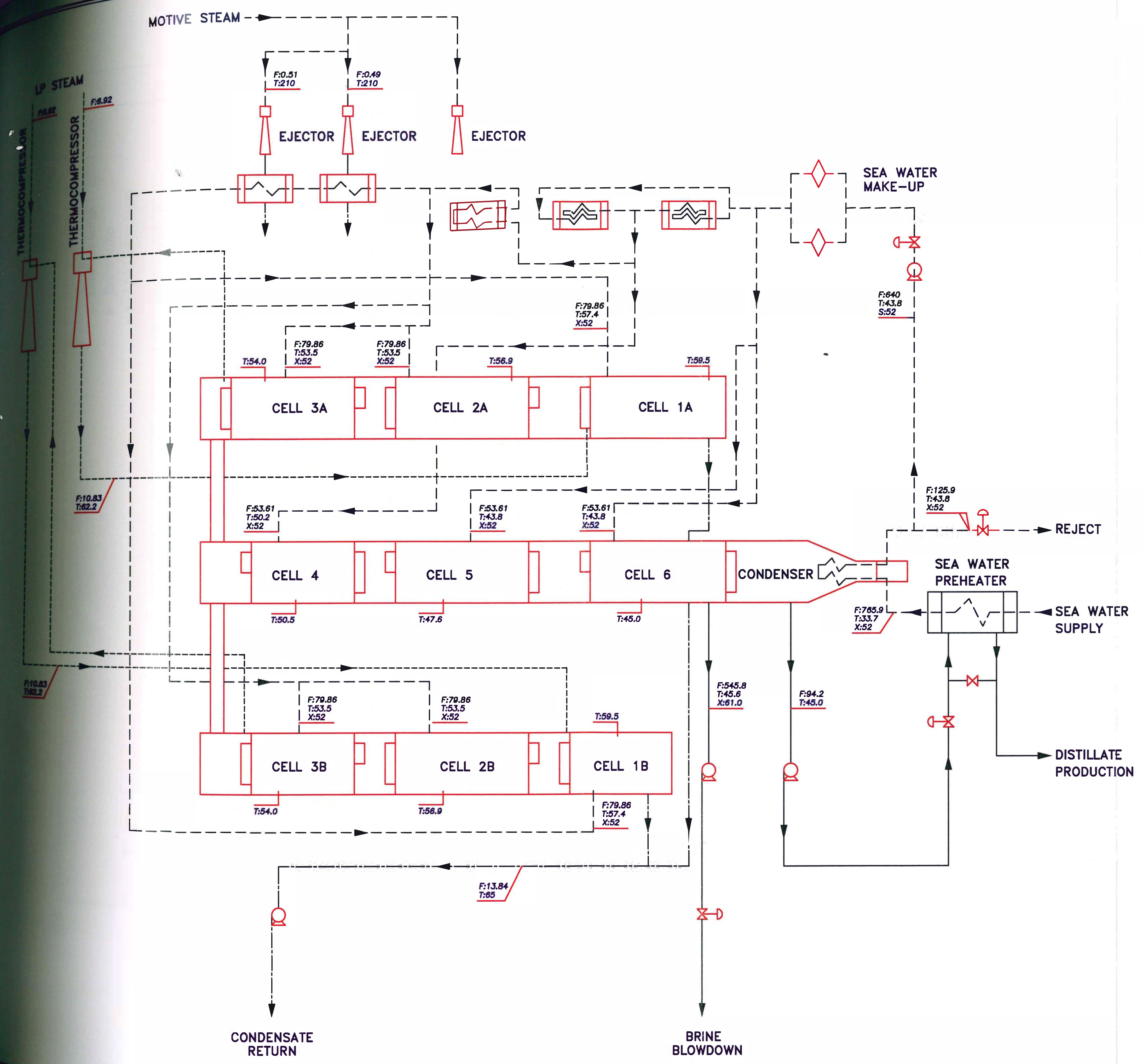


MEDNAR SIMULATOR

UANPC MED-TVC UNITS

U.A.E University

Title: Figure 6–1:
Mass & Energy Balance
for 100% Production Rate



MEDNAR SIMULATOR

UANPC MED-TVC UNITS

Key :

F: Flow Rate (kg/s)
T: Temperature (°C)
X: Salinity (1000 x ppm)

U.A.E University

Title: Figure 6–2: Mass & Energy Balance for 50% Production Rate

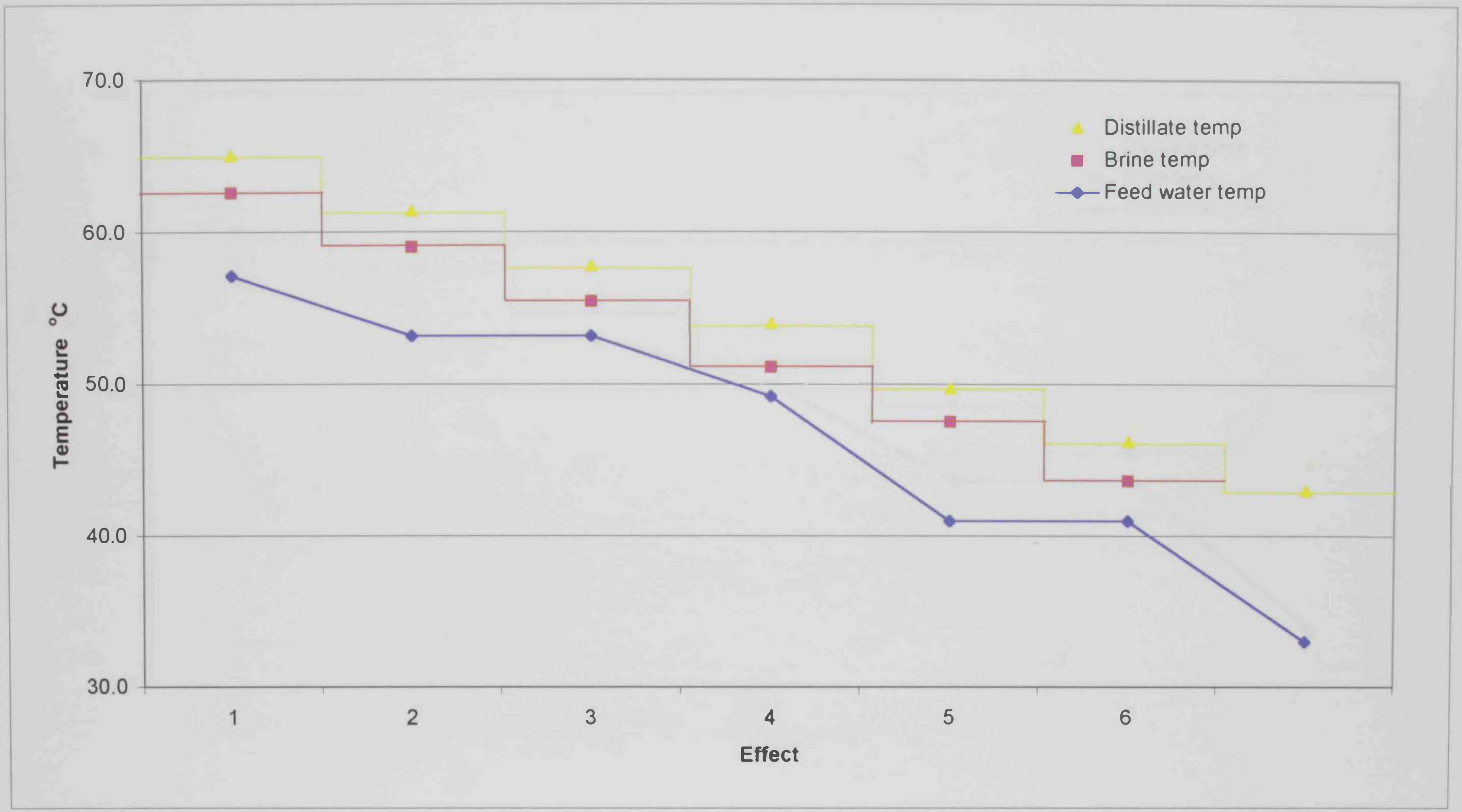


Figure 6-3: Temperature profiles for MEDNAR simulator run at 100% production rate

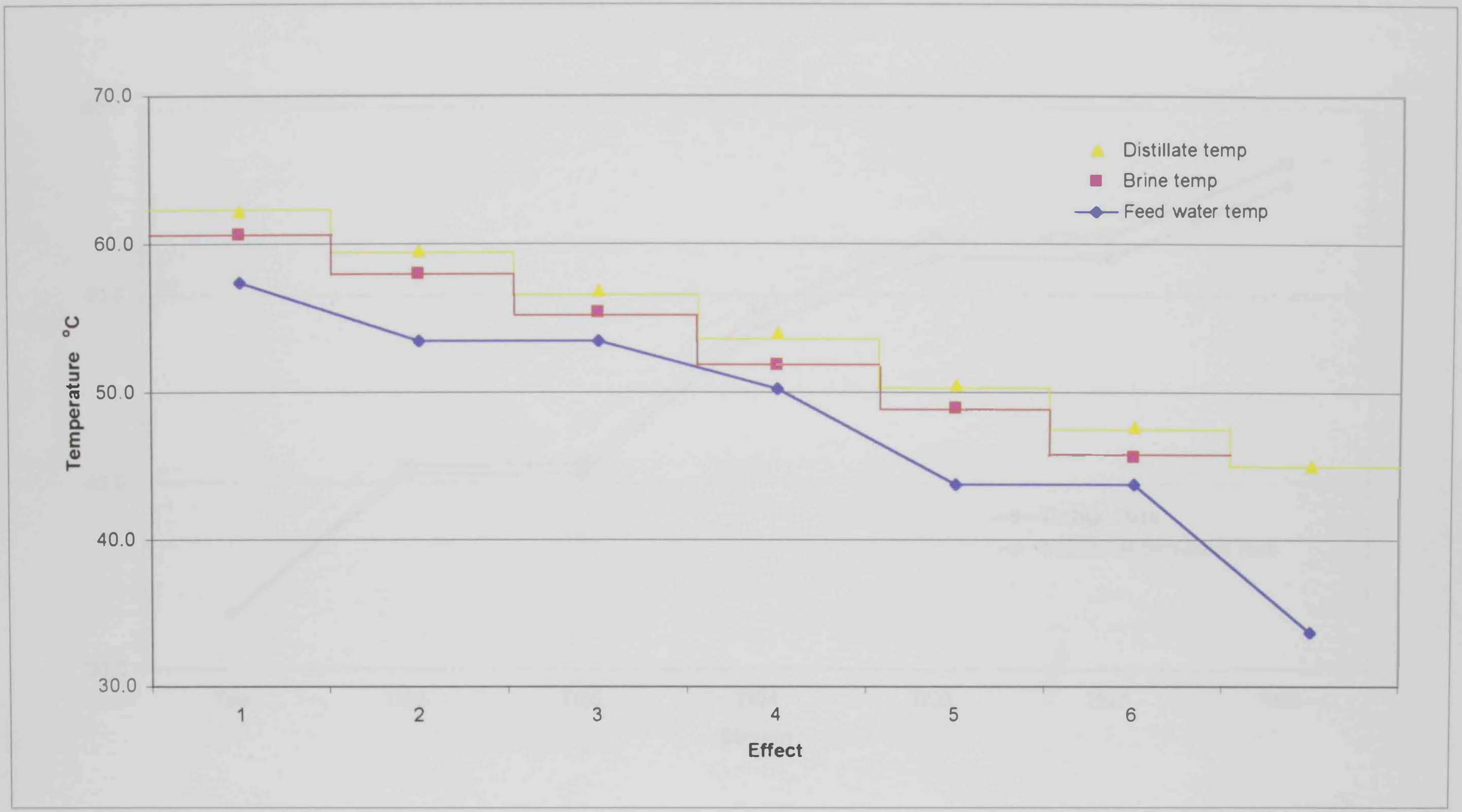


Figure 6-4: Temperature profiles for MEDNAR simulator run at 50% production rate

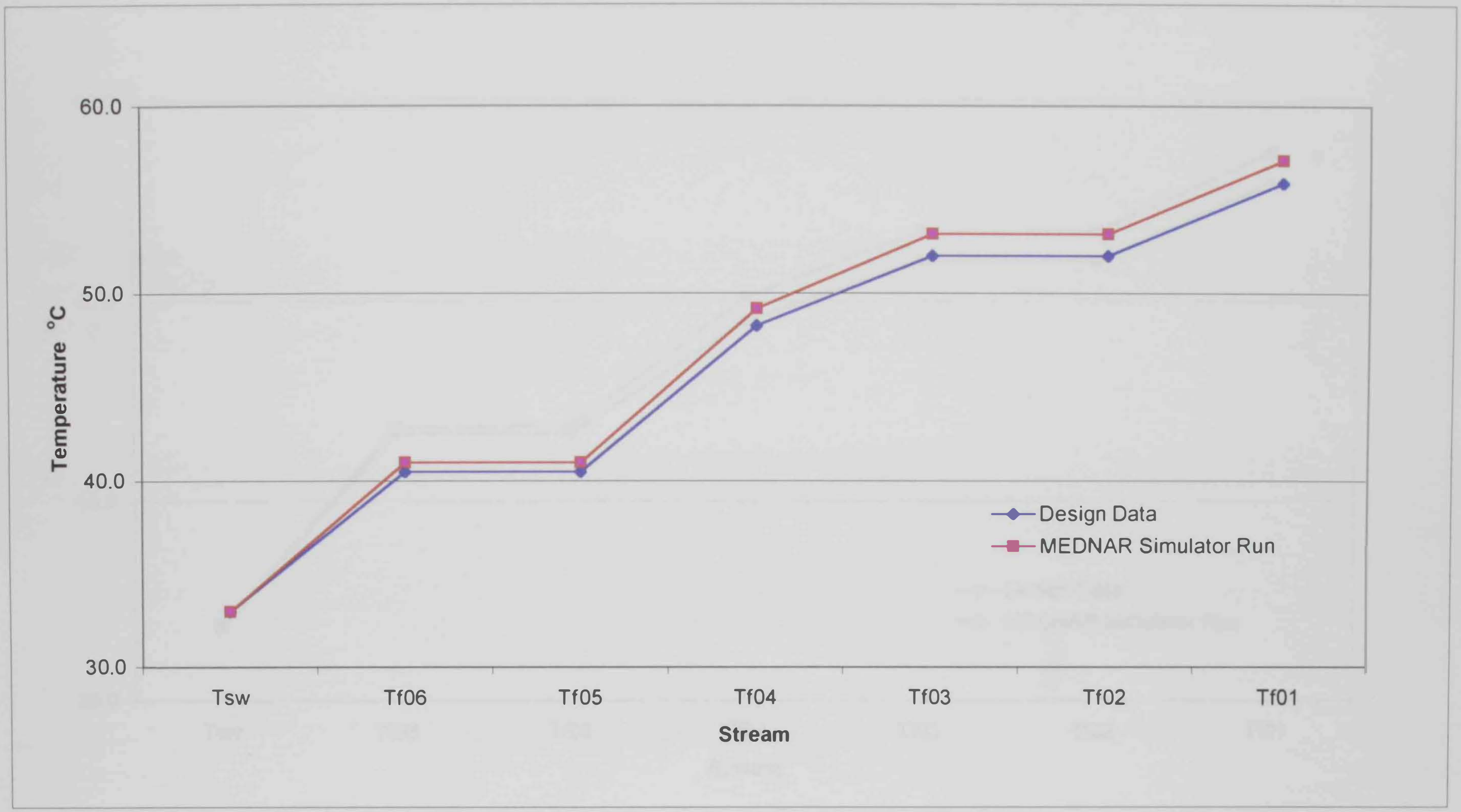


Figure 6-5: Feed water temperature profile for 100% production rate

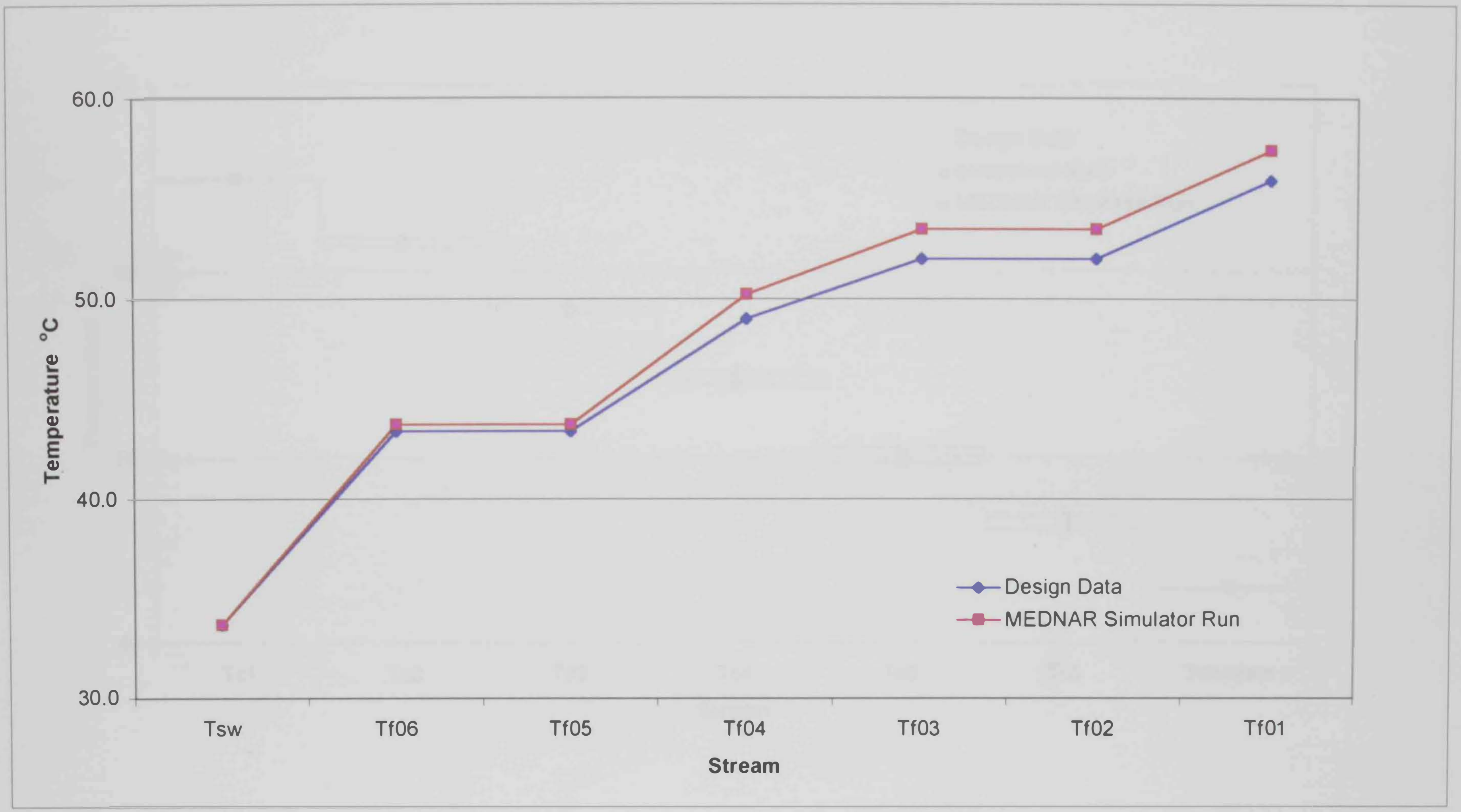


Figure 6-6: Feed water temperature profile for 50% production rate

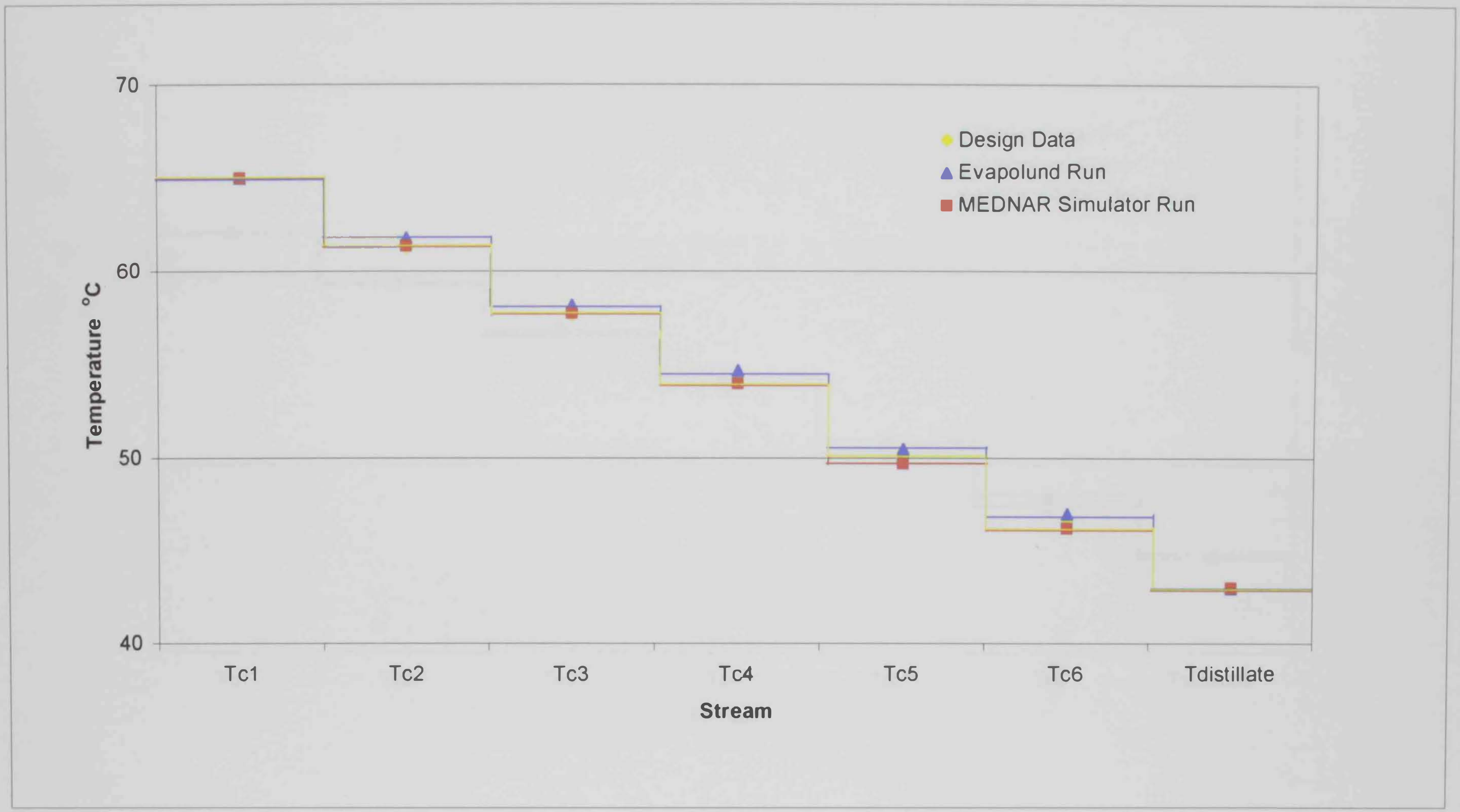


Figure 6-7: Vapor condensation temperature profile at 100% production rate

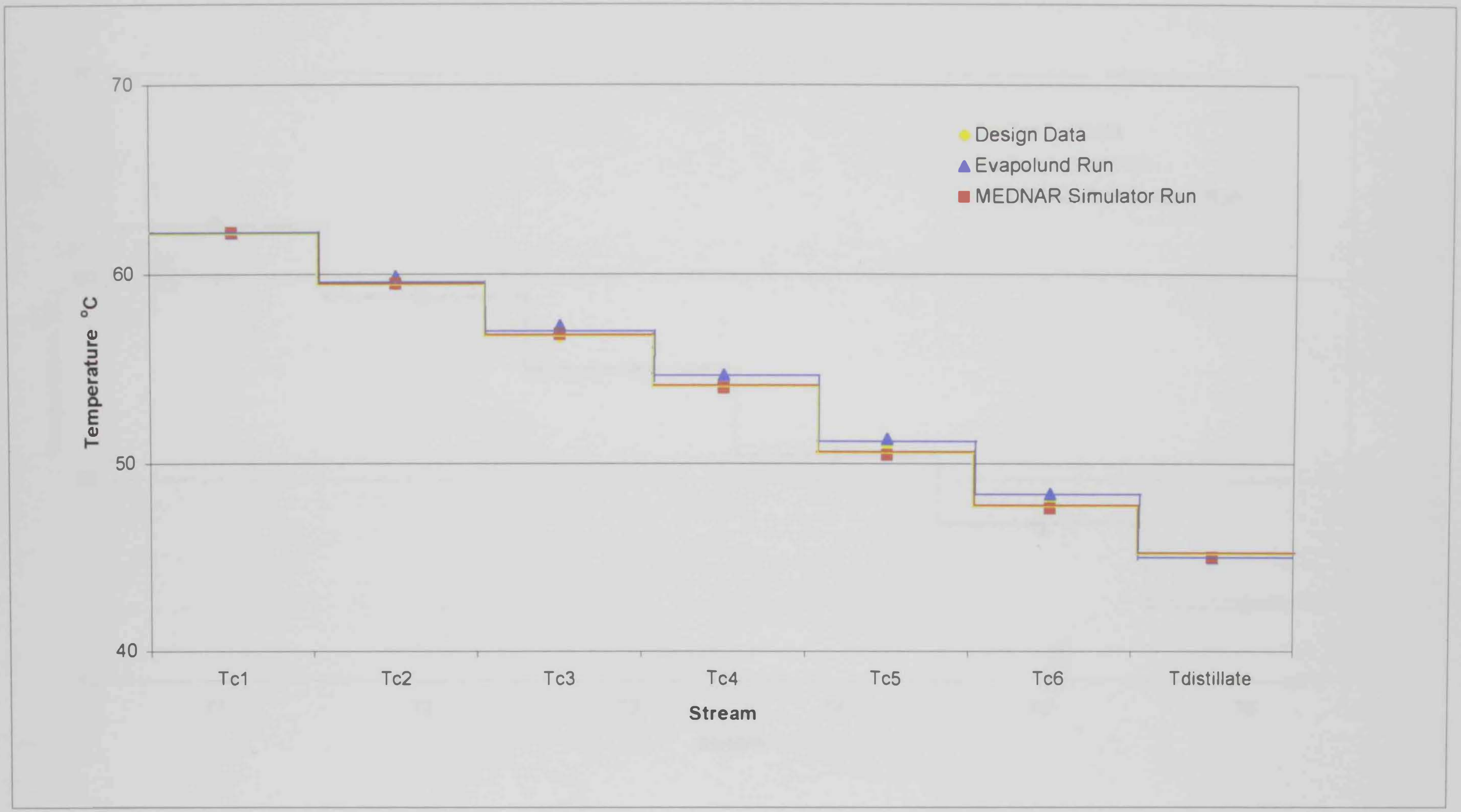


Figure 6-8: Vapor condensation temperature profile at 50% production rate

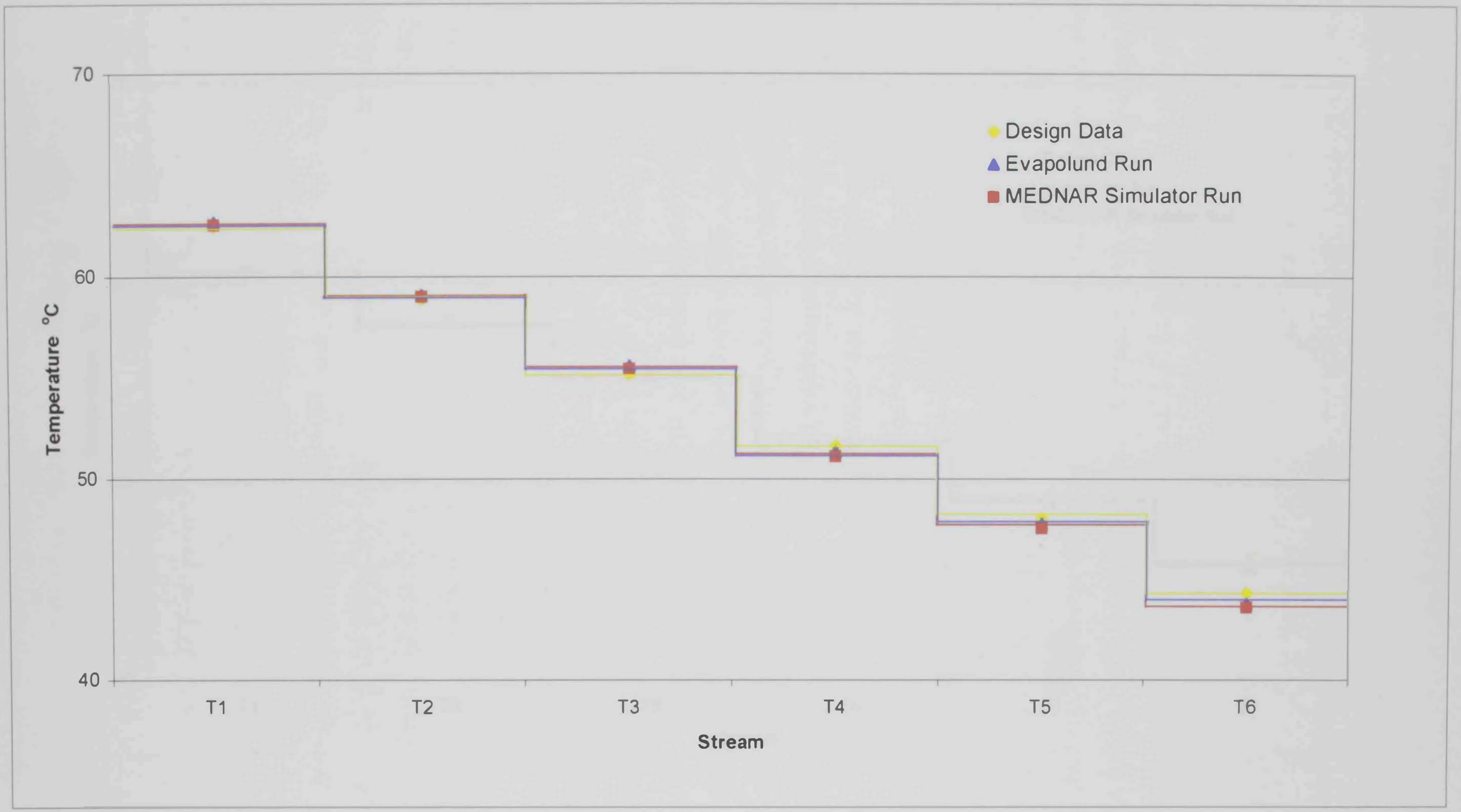


Figure 6-9: Brine temperature profiles at 100% production rate

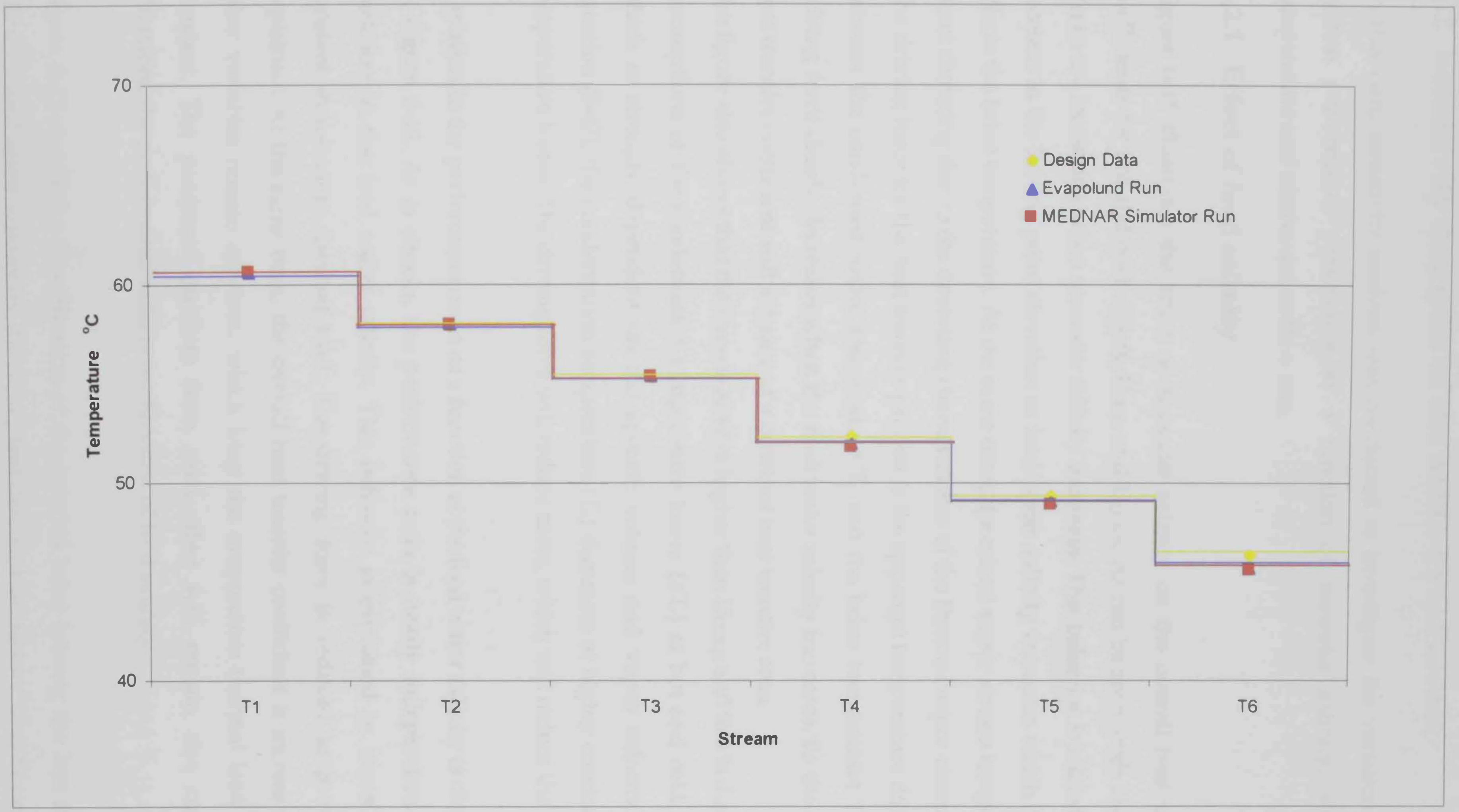


Figure 6-10: Brine temperature profiles at 50% production rate

6.2 Sensitivity Analysis of the MED-TVC Process

In this part, sensitivity analysis was conducted to investigate the variation in the system performance parameters as a function of seawater salinity, seawater temperature and motive steam flow rate.

6.2.1 Effect of feed salinity

Figure 6-11 illustrates the effect of seawater salinity on the overall heat transfer coefficients for hot and cold effects, U_{HOT} and U_{COLD}. As can be seen, both U_{HOT} and U_{COLD} are increasing when seawater salinity increases. This behavior is related to the increase in the boiling point elevation as feed water salinity increases which directly affects the brine temperature. At the same time, the mixed vapor stream temperature is not changing due to the operating characteristics of the thermal vapor compressor. The driving force for the heat transfer process is the apparent temperature difference between the condensed vapor temperature T_s and the brine temperature T_1. This driving force clearly decreases when the feed water salinity increases. So the overall heat transfer coefficient will increase at a constant heat transfer area.

The figure also shows that the U_{HOT} is always higher than U_{COLD} and this is due to the assumptions of the condensation temperature losses (ΔT_c) at hot and cold effects, which is strongly dependent on the specific volume and vapor velocity. From equation (5-47), the condensation temperature (T_c) decreases at higher condensation temperature losses. The driving force will reduce more, which will reduce the U_{COLD}.

Variation in the performance ratio as a function of the feed water salinity is displayed in Figure 6-12. As is shown, the performance ratio is nearly independent of the variation in the feed water salinity. This behavior is explained in terms of the constant evaporators thermal load. The driving force is reduced as previously explained. At the same time, the overall heat transfer coefficient is increased. The other variables remain constant, which keep the evaporators thermal load nearly constant. The produced distillate from each effect will remain also constant. Therefore, the performance ratio is not function of feed salinity.

Figure 6-13 shows the concentration of the rejected brine leaving the last stage at different feed water salinity for different plant distiller load factors. This figure could be very helpful to determine the maximum allowable load at different feed concentration based on the maximum allowable blow down brine salinity.

6.2.2 Effect of the flow rate of motive steam

A number of simulation runs were executed to investigate the effect of the flow rate of motive steam on several process variables.

Figure 6-14 shows the performance ratio (PR) as a function of the flow rate of motive steam. It is clear that PR is strongly dependent on the motive steam. In the MED theory, each 1 kg of steam supplied to an effect will produce 1 kg of distillate. The more steam supplied the higher the evaporator thermal load, see equation (5-31). This will lead to higher heat transfer rates and more vapors will be generated. This extra vapor is used as energy stream in the next evaporator and which condenses at the same time and cascaded as a distillate. In this plant, the first two effects produce most of the vapor, see Figure 6-15. The first three effects have much larger area compared to the next three effects (around three times). Also it has the thermo vapor compressor, which is a recycling part of the vapor from the third effect to the first one. The curves in Figure 6-15 have small drop in the fifth effect, because of the pre-heater 12, which condenses part of the vapor to heat up the feed water.

The increase in the PR as a function of motive steam has a lot of restrictions. Two of these restrictions are the blow down brine salinity and the last effect vapor velocity, see Figure 6-17. As the motive steam increases the brine salinity increases in each effect, see equations (5-32) and (5-34). This rejected salinity has environmental restrictions that do not allow it to be exceeded. The maximum design value in this plant is 73000 ppm.

The last effect vapor velocity is also an important restriction. As the temperature is reduced from effect to effect due to heat transfer losses the pressure will also drop. The last effect vapor velocity is the lowest because of its high specific volume confined in a limited space. As the pressure is reduced the vapor velocity is increased and could reach a limit that disturbs the thin seawater-brine film around the evaporator tubes, which reduce the evaporator efficiency and may create dry spots, which lead to fatal scaling. On the other hand, a tube and baffle erosion could also be created.

Figure 6-16 shows the thermal vapor compressor behavior. The difference between the mixed vapor flow curve and the sucked vapor flow curve is the motive steam flow rate.

The mechanical size for the thermal vapor compressor could be one of the important capacity restrictions. The more vapor sucked, the larger size needed.

The relation between the specific heat transfer area and the specific cooling water flow rate as a function of the motive steam is illustrated in Figure 6-18. The specific heat transfer area strongly depends on the motive steam flow rate. It decreases linearly as the motive steam flow increases. This drop is due to constant area and strong relation between the motive steam and distillate flow rate as was discussed previously. The specific cooling water flow rate is increasing as the load increases. The curve reaches a maximum value at full load and start reducing again. The explanation for this behavior is clear in Figure 6-19. The cooling water flow rate curve increases approximately linear. On the other hand, the distillate flow rate shows a sharp increase at high load due to high performance ratio as explained before, go back to Figure 6-14 for more details. The differences between the two curves increases again at high load, which results in a maximum point.

Figure 6-20 shows the evaporators overall heat transfer coefficients as function of the flow rate of the motive steam. When more motive steam is used, the evaporators thermal load will increase, with constant heat transfer area and temperature driving force, the U value will increase. The difference between the U_{HOT} and the U_{COLD} is due to the assumptions of the condensation temperature losses (ΔT_c) at hot and cold effects, which have been discussed before. The U_{COLD} dotted line expresses the design data, which gives at low load higher final condenser condensation temperature (T_{c6}) than high load. This is to increase the U_{COLD} at low load to an acceptable average as indicated by equations (5-15) and (5-16).

6.2.3 Effect of seawater temperature

The next two Figures (6-21, 6-22) depict the specific cooling water flow rate and final condenser overall heat transfer coefficient as functions of seawater temperature.

In general, when the seawater temperature increases, the U_{cond} will increase as indicated by equations (5-15) and (5-16). At the same time, more cooling water is needed according to equation (5-26).

The interesting point that these two figures are observed a minimum seawater temperature at low load. Seawater pre-heater is therefore a must with low seawater temperature.

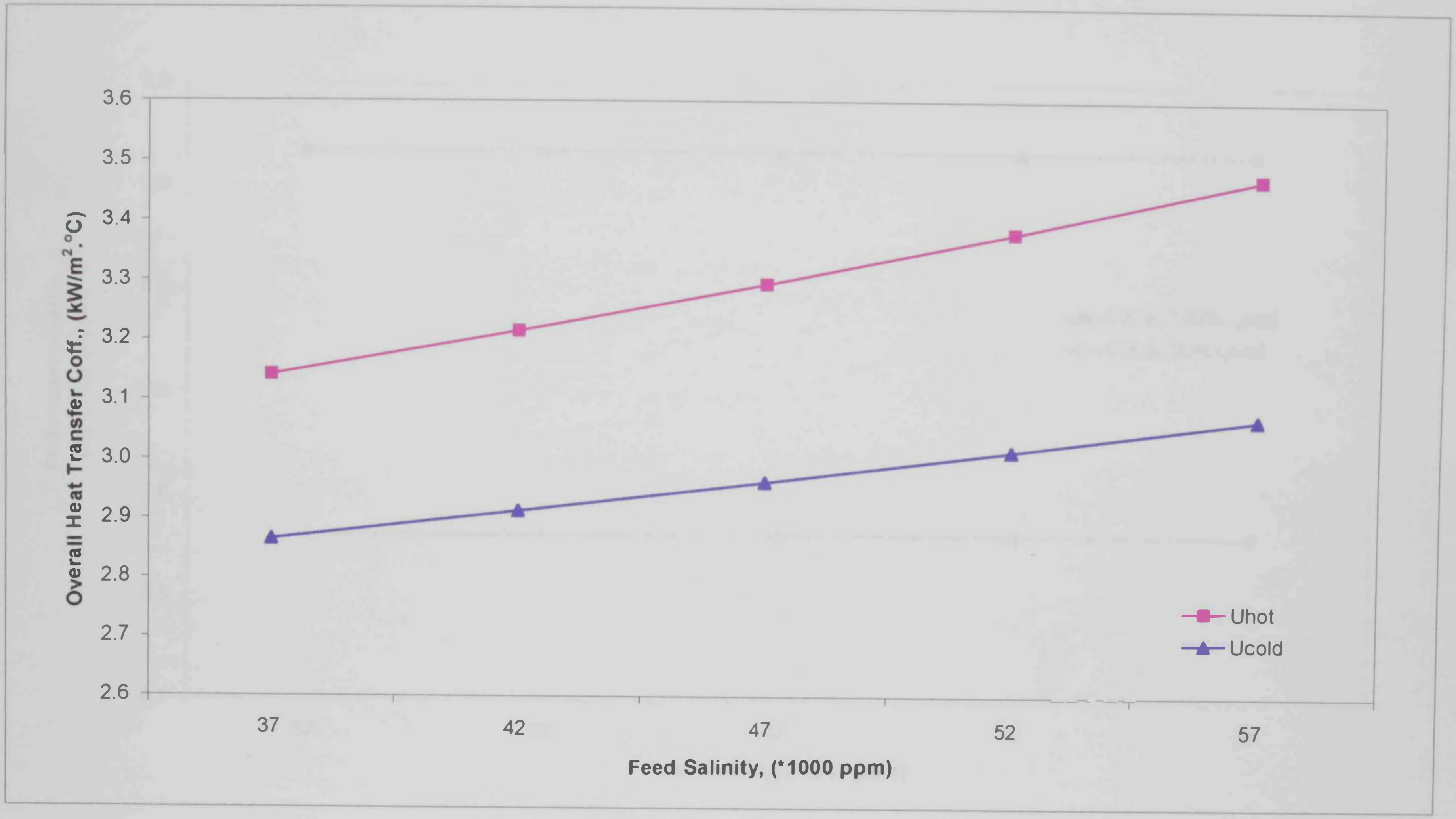


Figure 6-11: MEDNAR runs at 100% production rate for overall heat transfer coefficient as a function of feed salinity

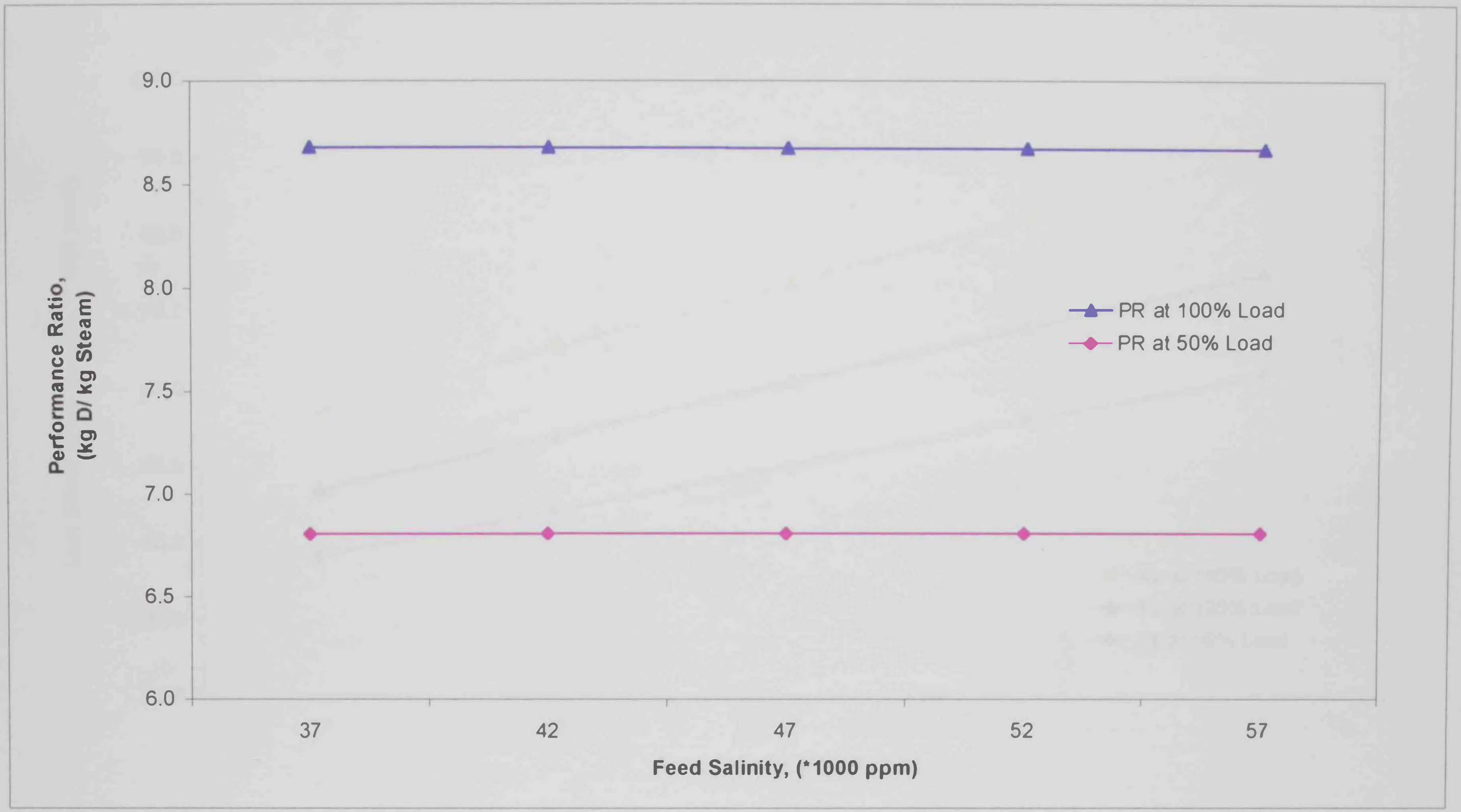


Figure 6-12: MEDNAR runs at 100% and 50% production rates for performance ratio as a function of feed salinity

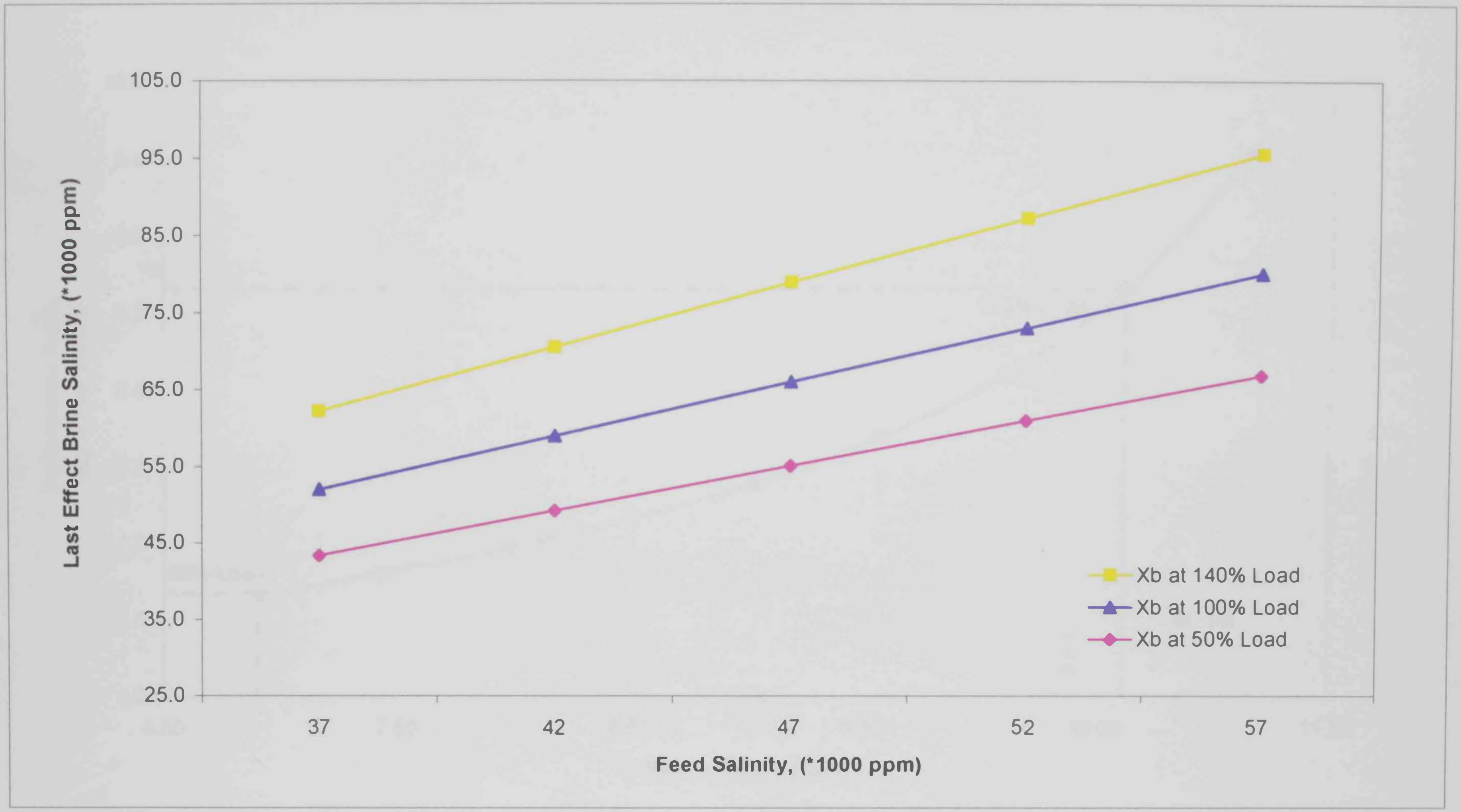


Figure 6-13: MEDNAR runs at different production rates for brine blow down salinity as a function of feed salinity

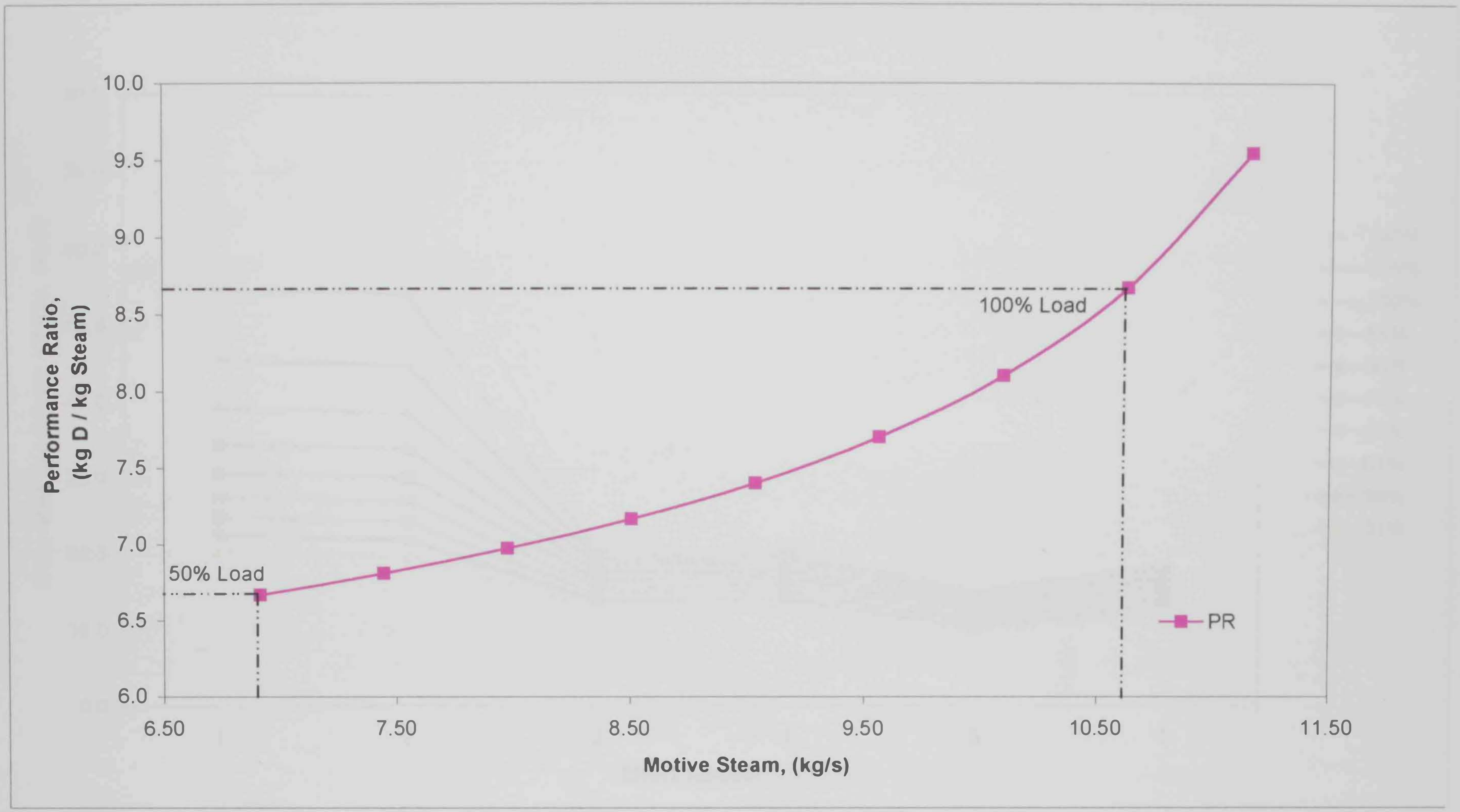


Figure 6-14: MEDNAR runs at different production rates for performance ratio as a function of motive steam

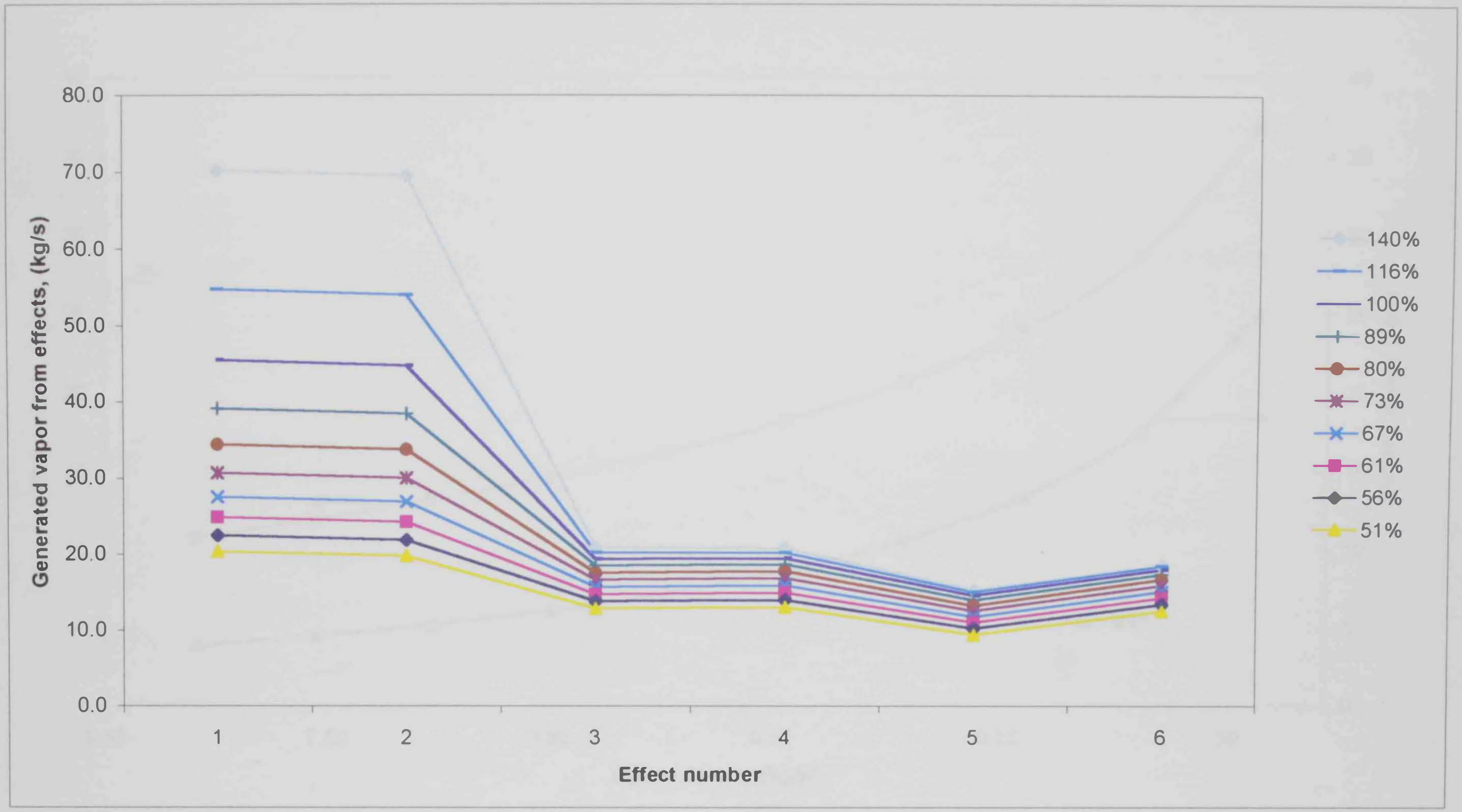


Figure 6-15: Generated vapor flow rates at different production rates

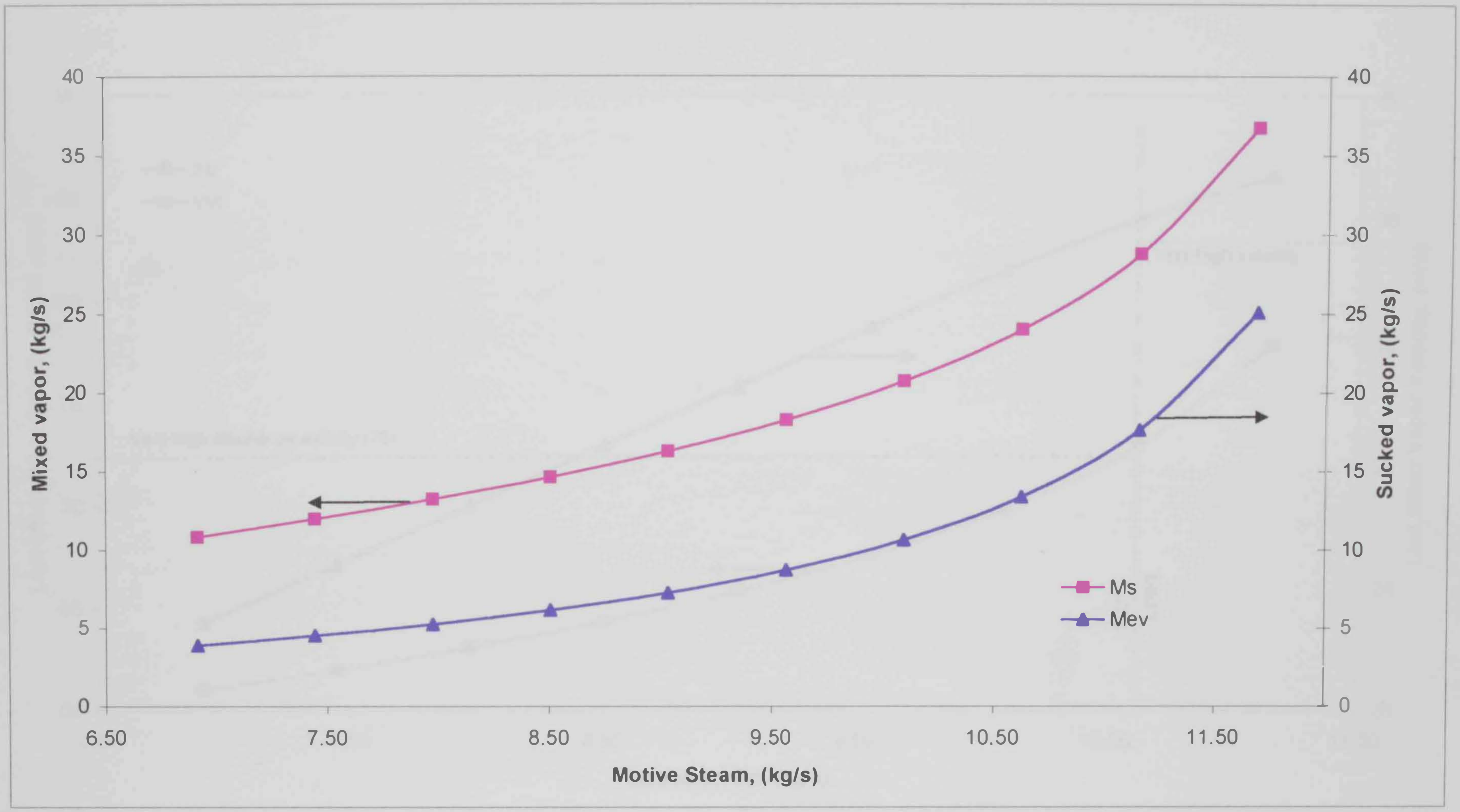


Figure 6-16: Flow rate of sucked vapor and mixed stream from the steam ejector at different motive steam flow rates

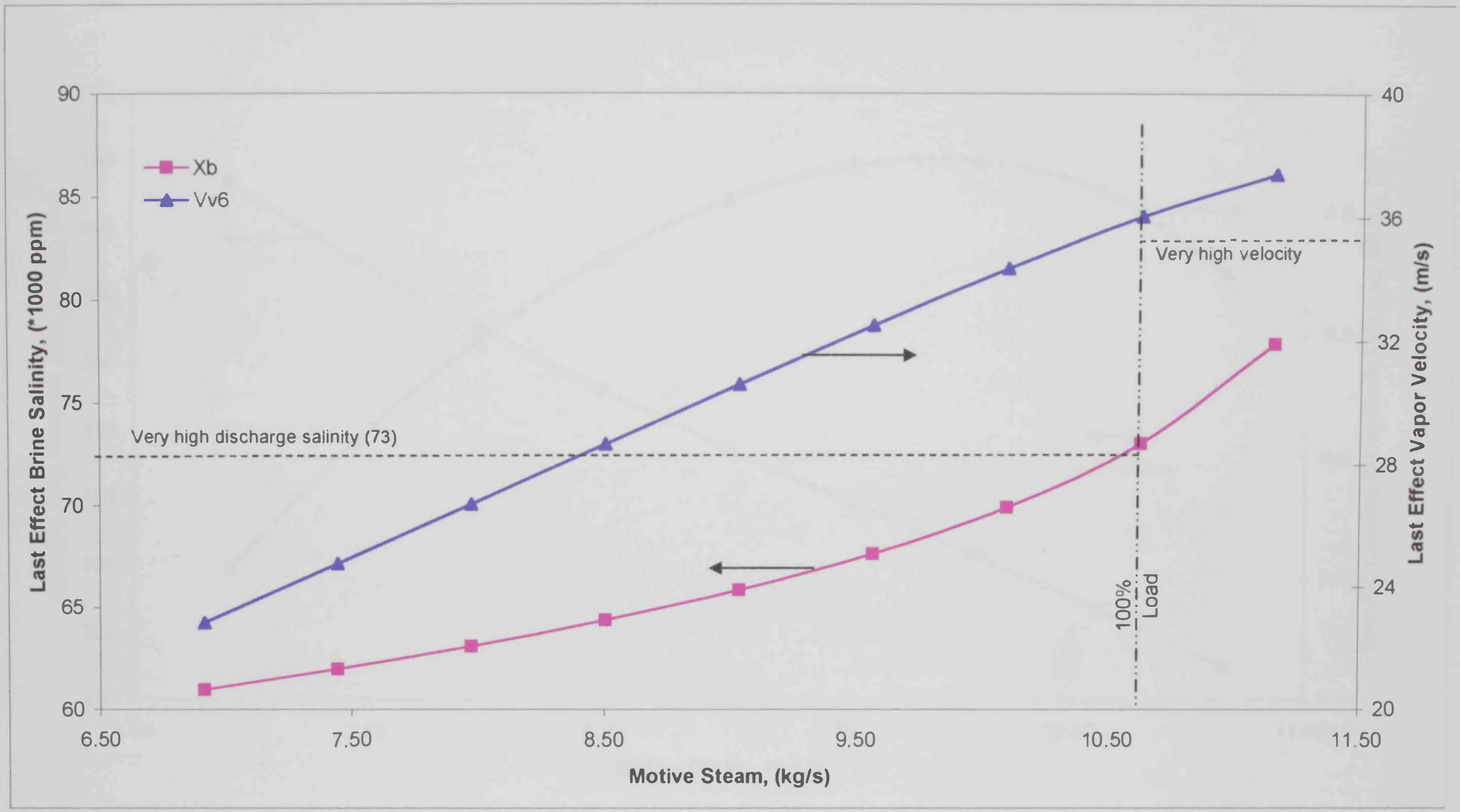


Figure 6-17: MEDNAR runs at different production rates illustrating the parameters that may limit the production rate

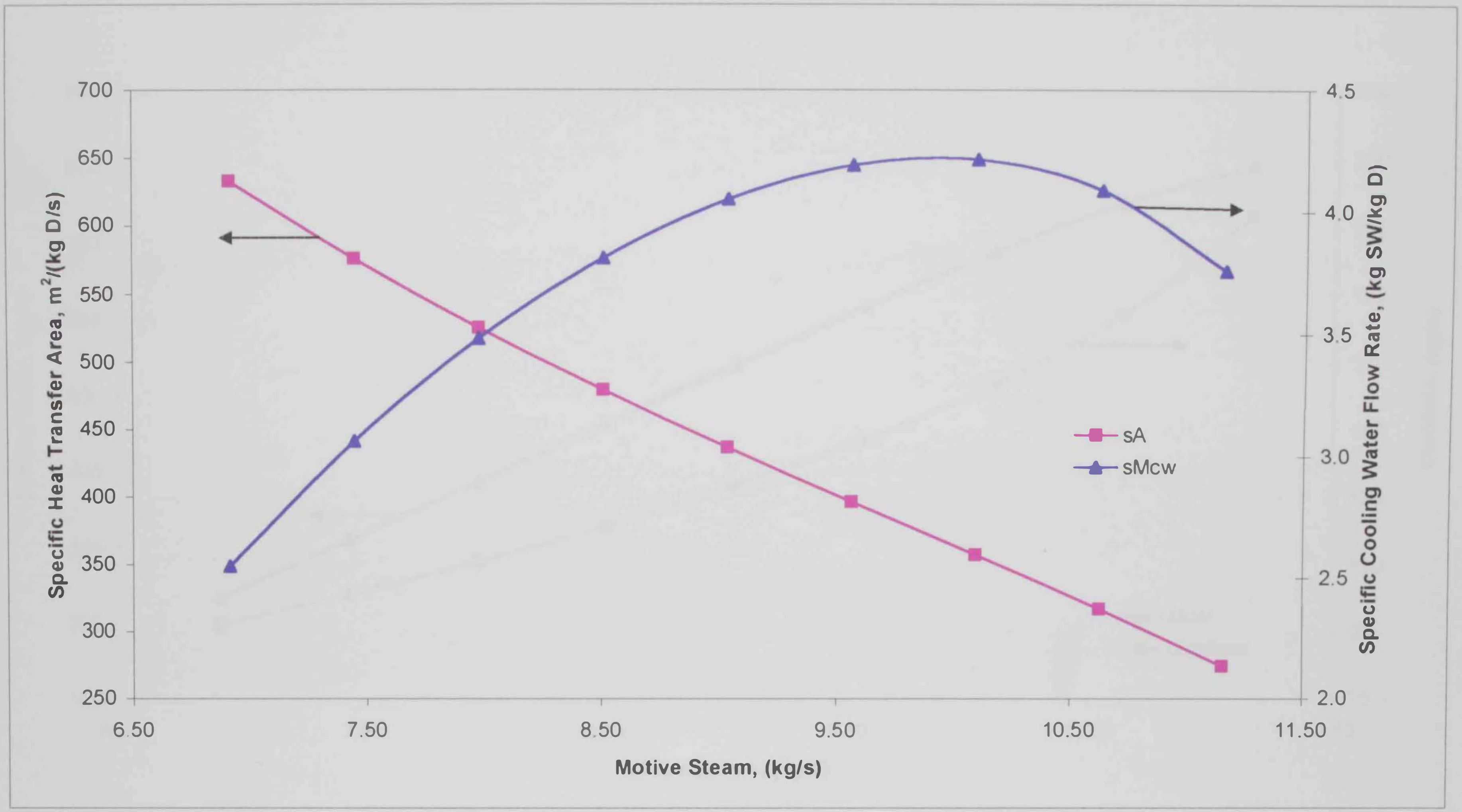


Figure 6-18: MEDNAR runs for specific heat transfer area and specific cooling water flow rates as a function of motive steam

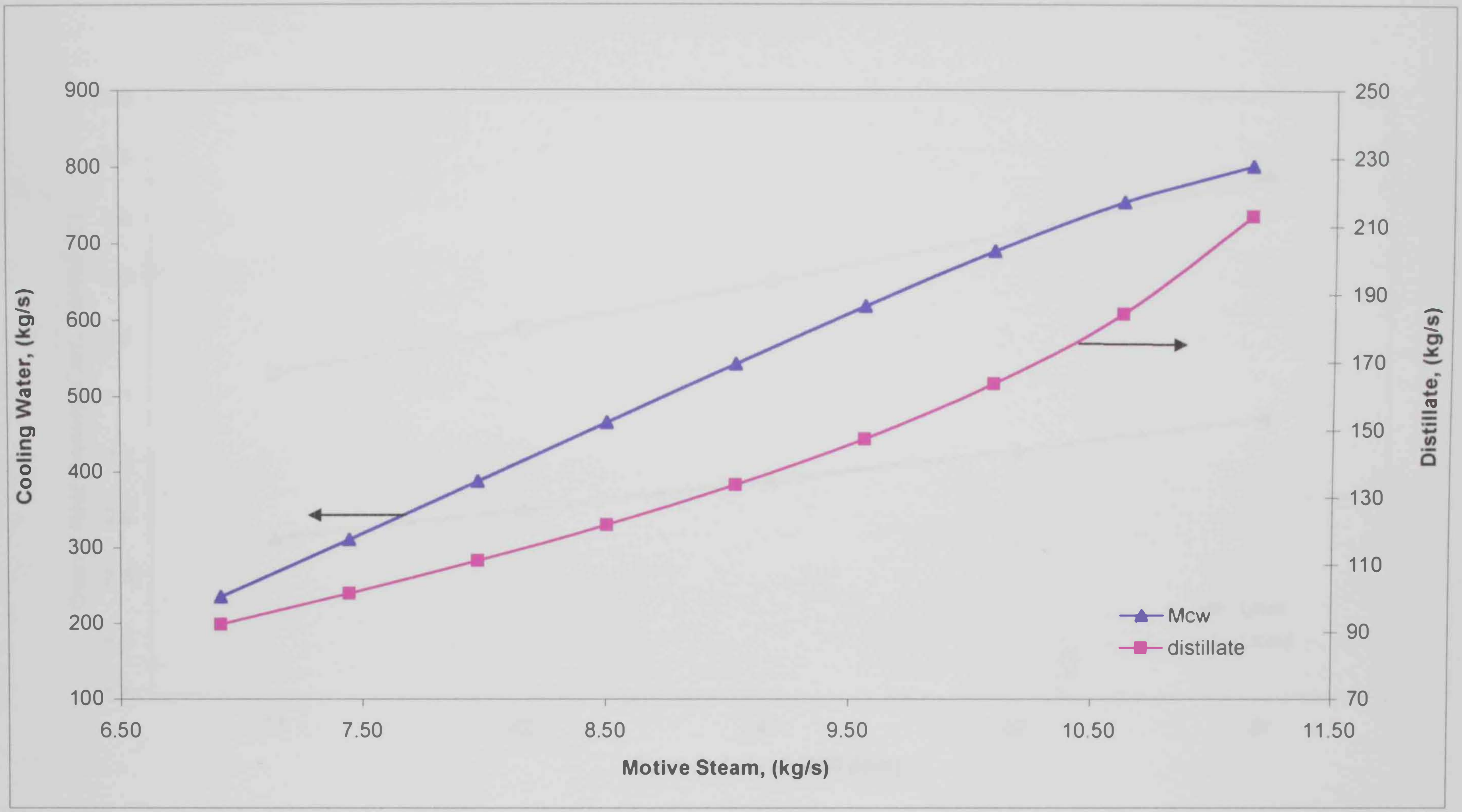


Figure 6-19: MEDNAR runs at different production rates for cooling water and distillate flow rates as a function of motive steam

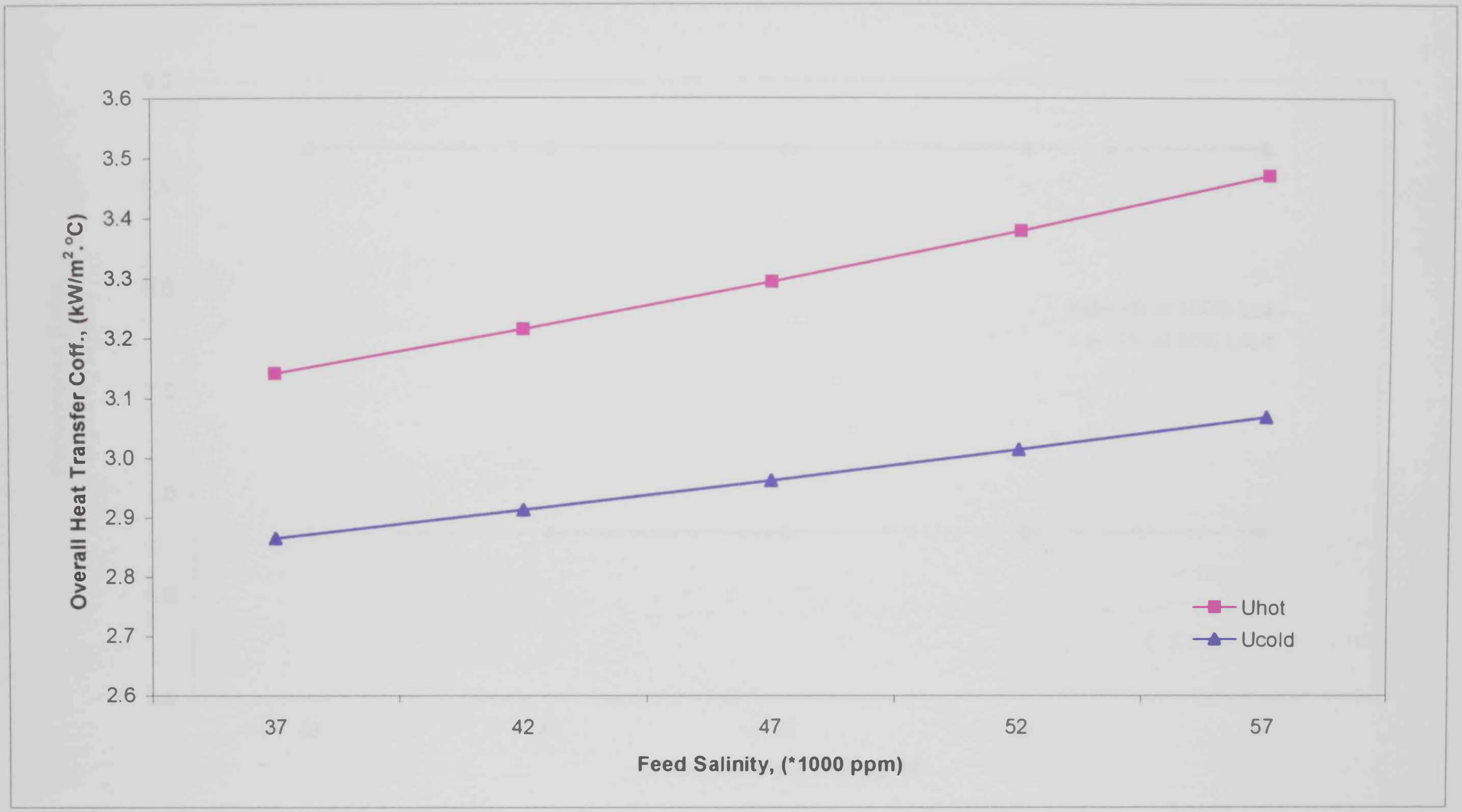


Figure 6-11: MEDNAR runs at 100% production rate for overall heat transfer coefficient as a function of feed salinity

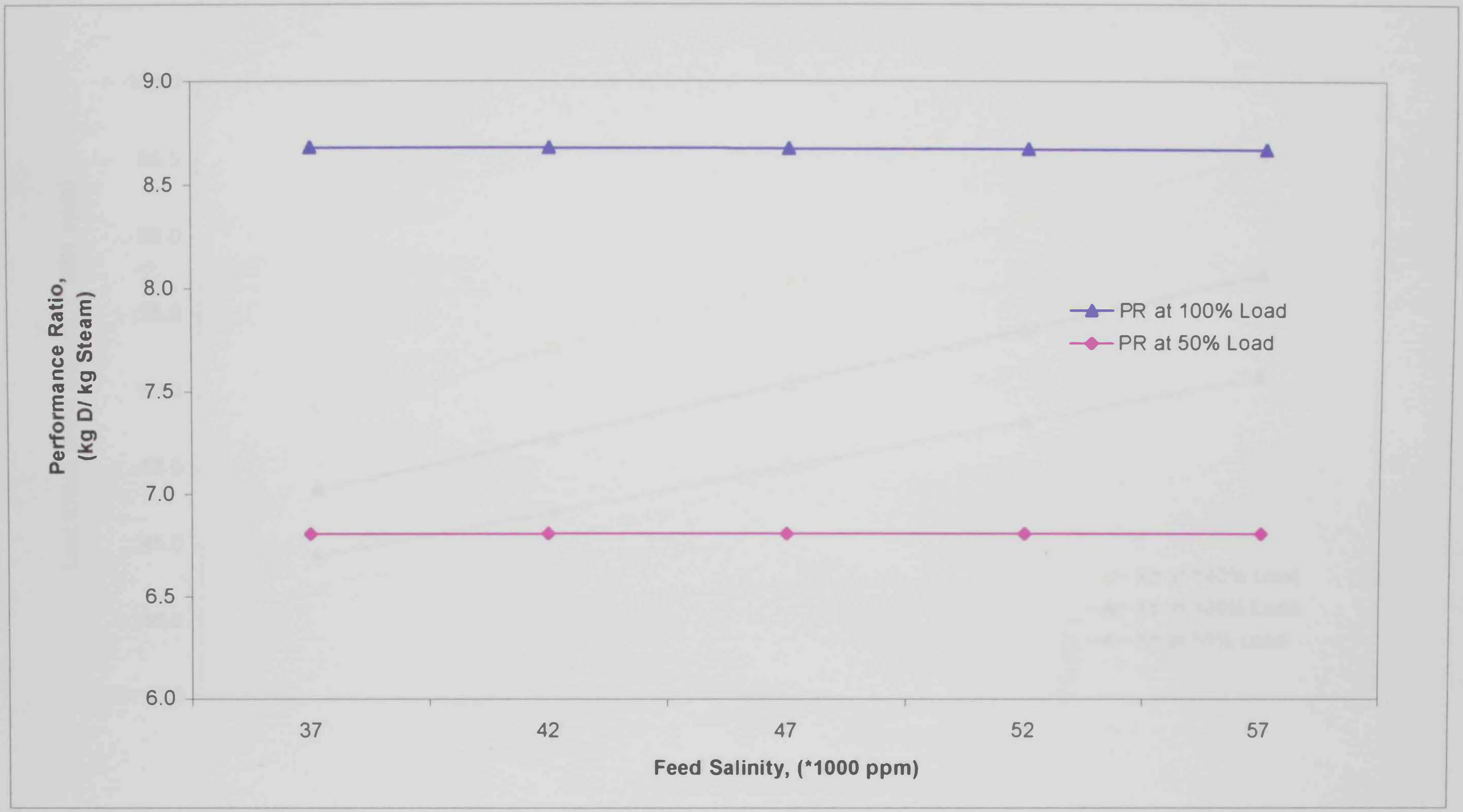


Figure 6-12: MEDNAR runs at 100% and 50% production rates for performance ratio as a function of feed salinity

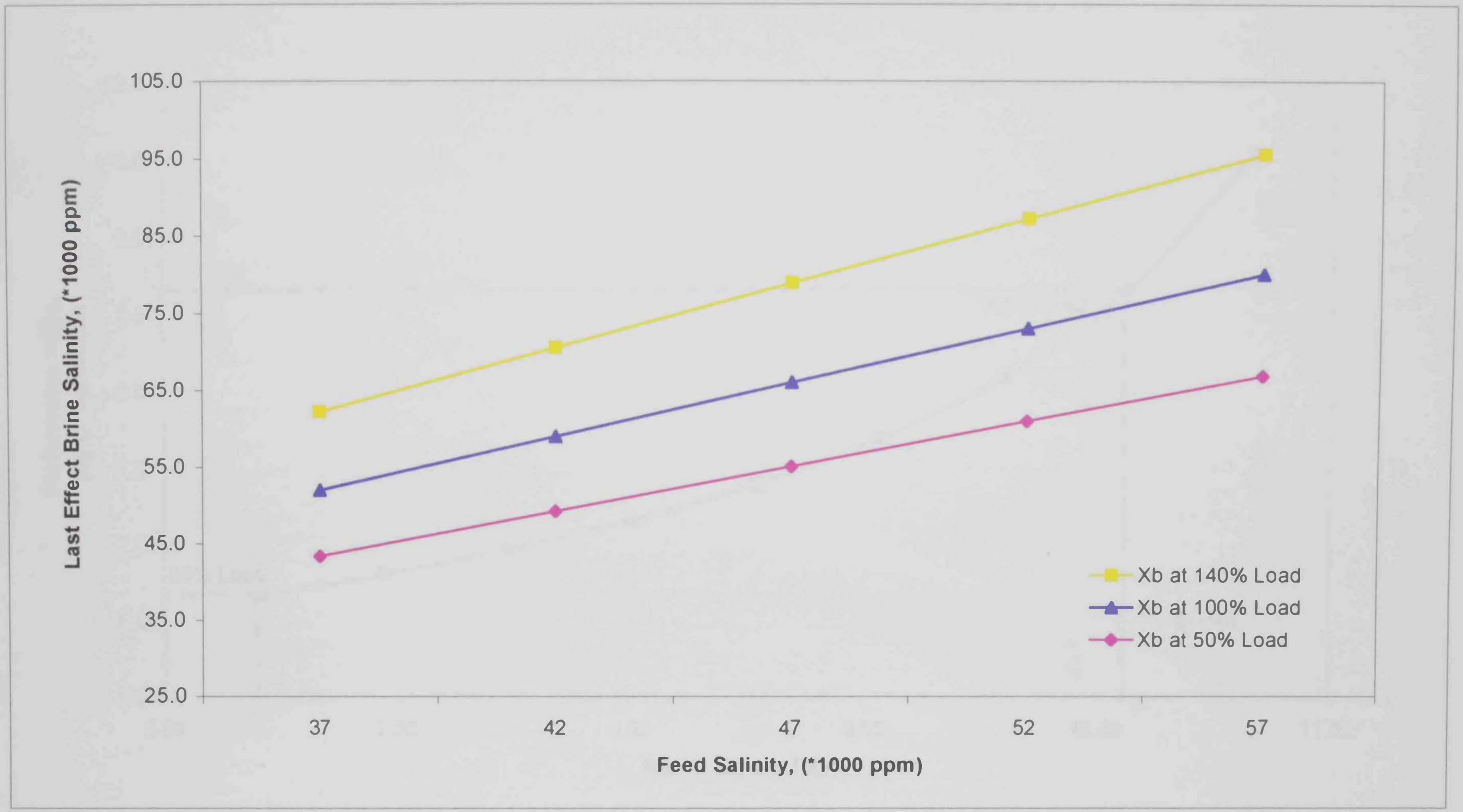


Figure 6-13: MEDNAR runs at different production rates for brine blow down salinity as a function of feed salinity

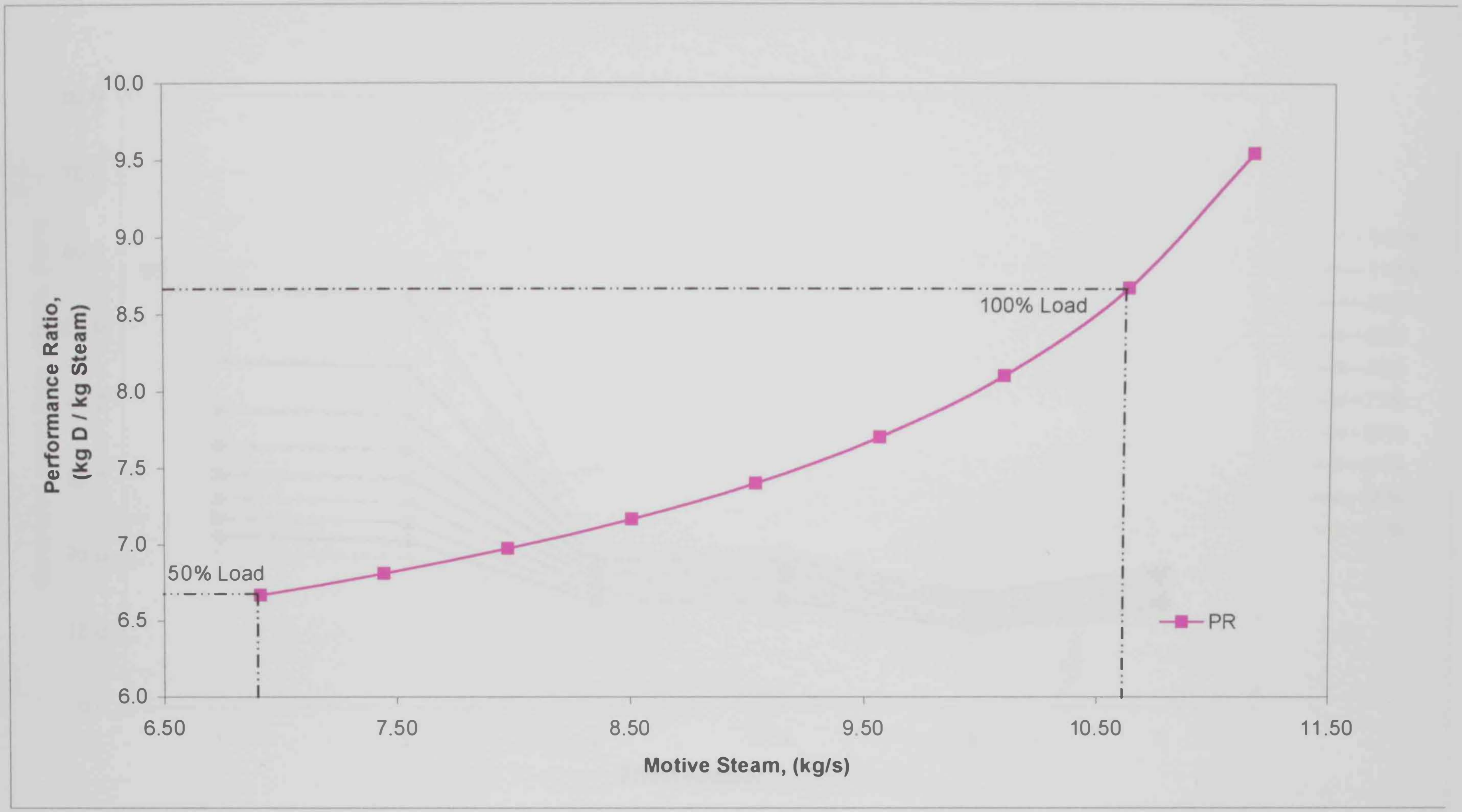


Figure 6-14: MEDNAR runs at different production rates for performance ratio as a function of motive steam

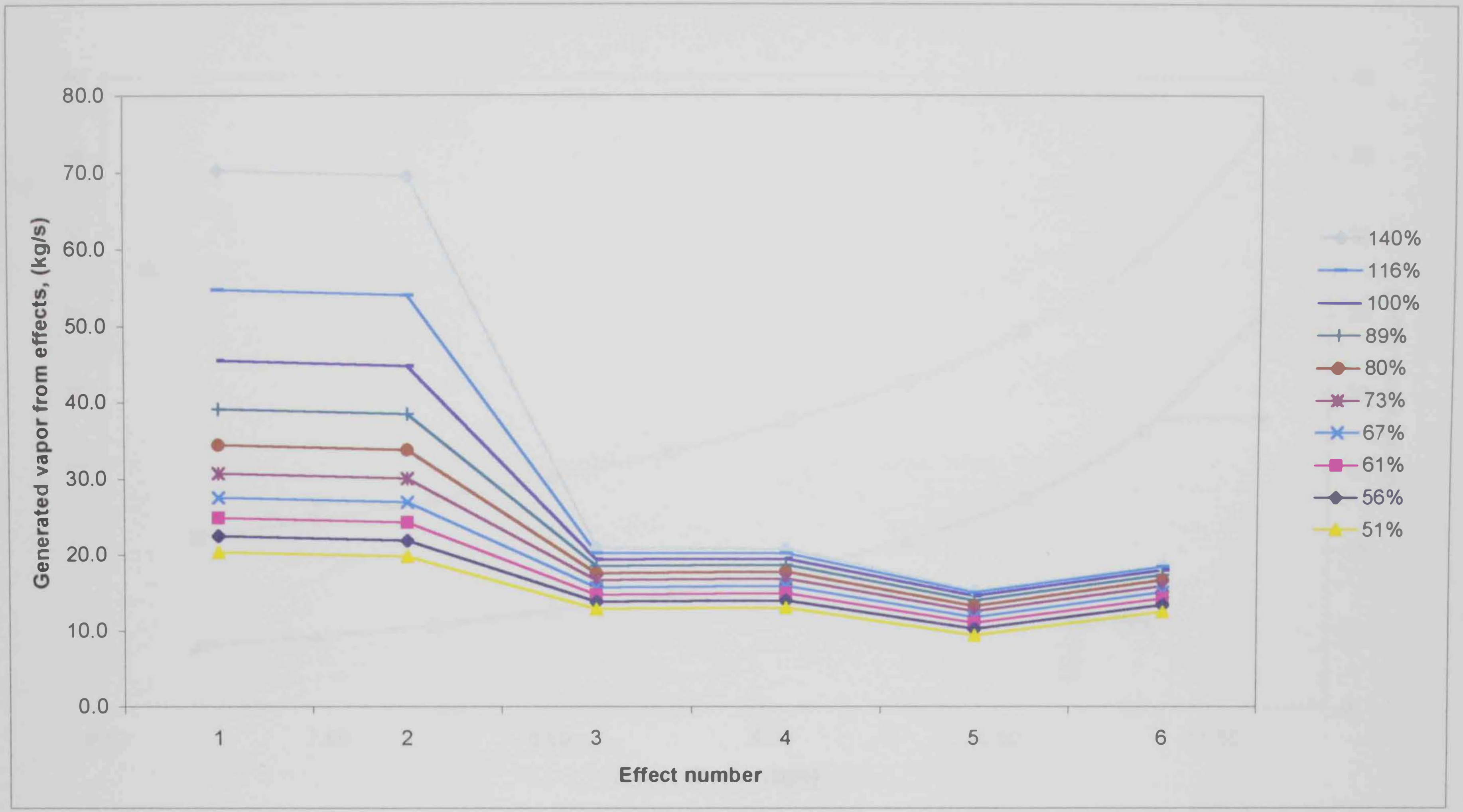


Figure 6-15: Generated vapor flow rates at different production rates

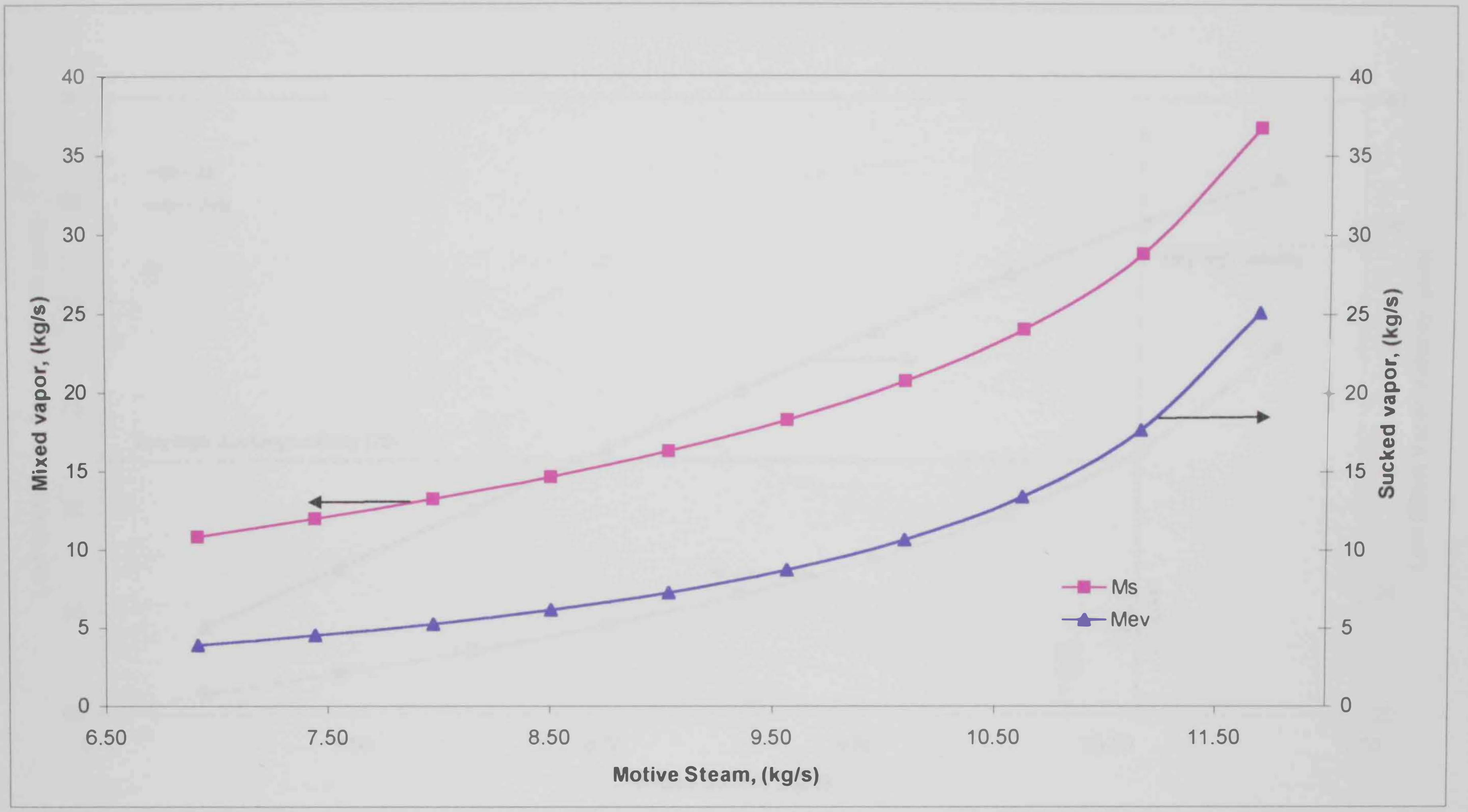


Figure 6-16: Flow rate of sucked vapor and mixed stream from the steam ejector at different motive steam flow rates

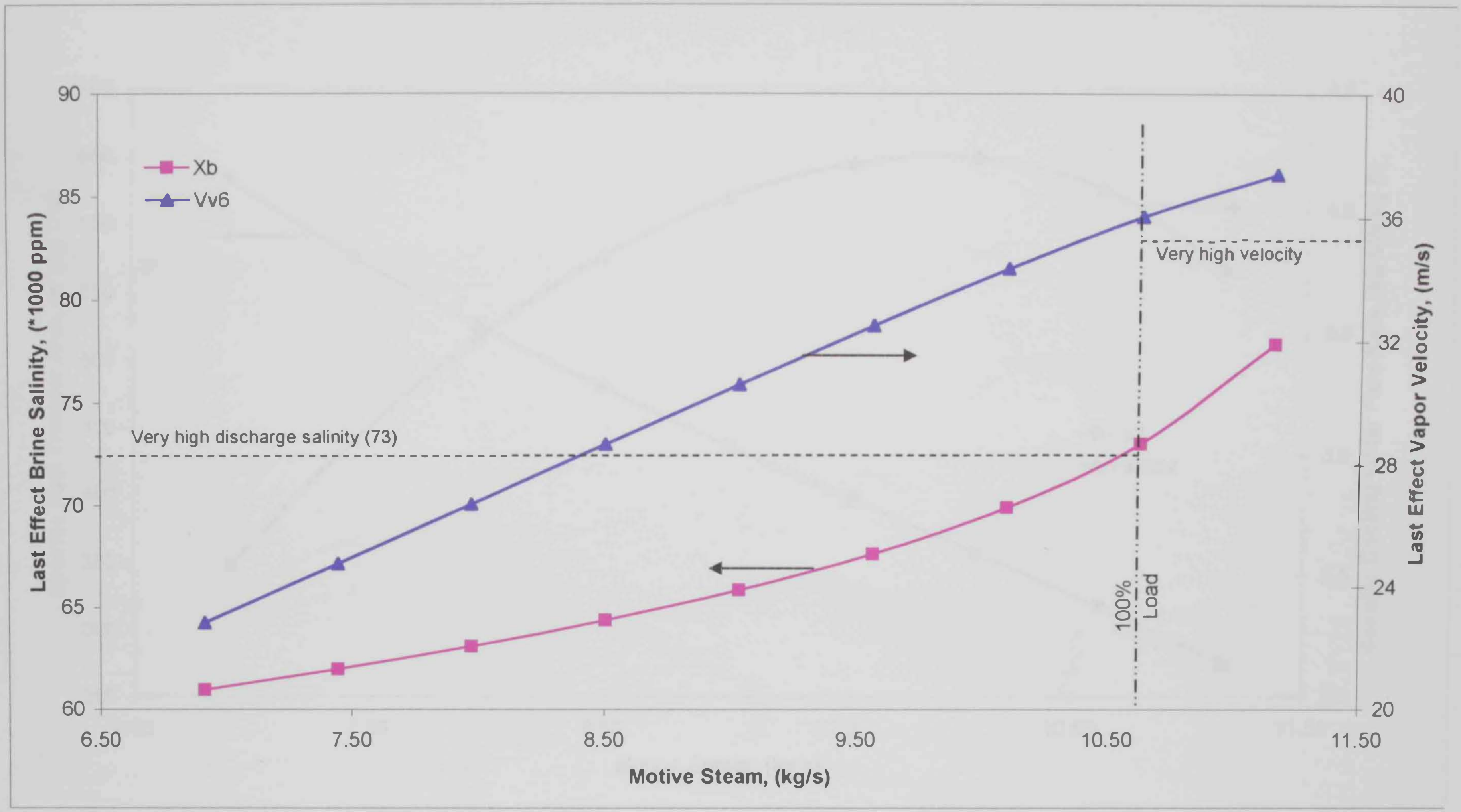


Figure 6-17: MEDNAR runs at different production rates illustrating the parameters that may limit the production rate

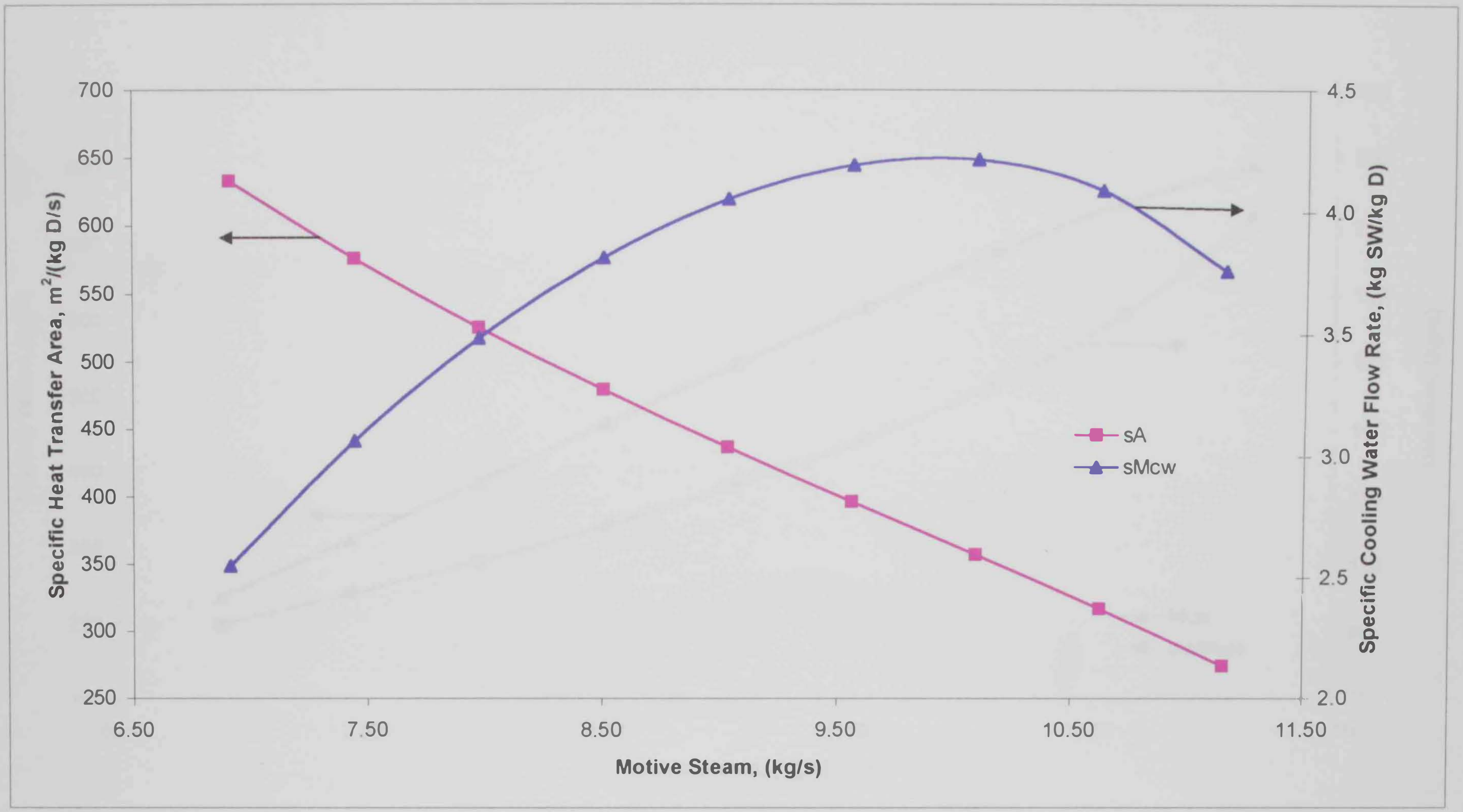


Figure 6-18: MEDNAR runs for specific heat transfer area and specific cooling water flow rates as a function of motive steam

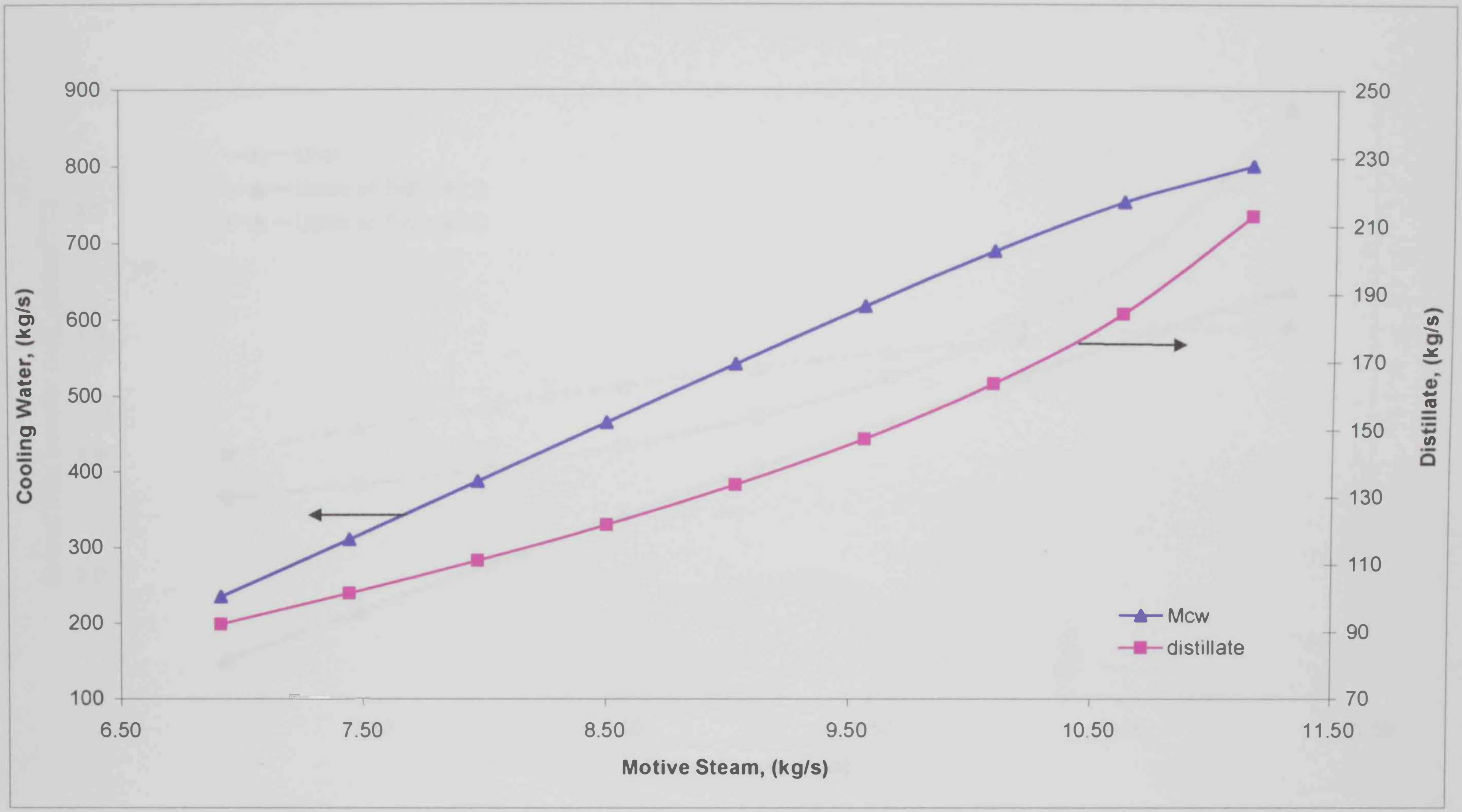


Figure 6-19: MEDNAR runs at different production rates for cooling water and distillate flow rates as a function of motive steam

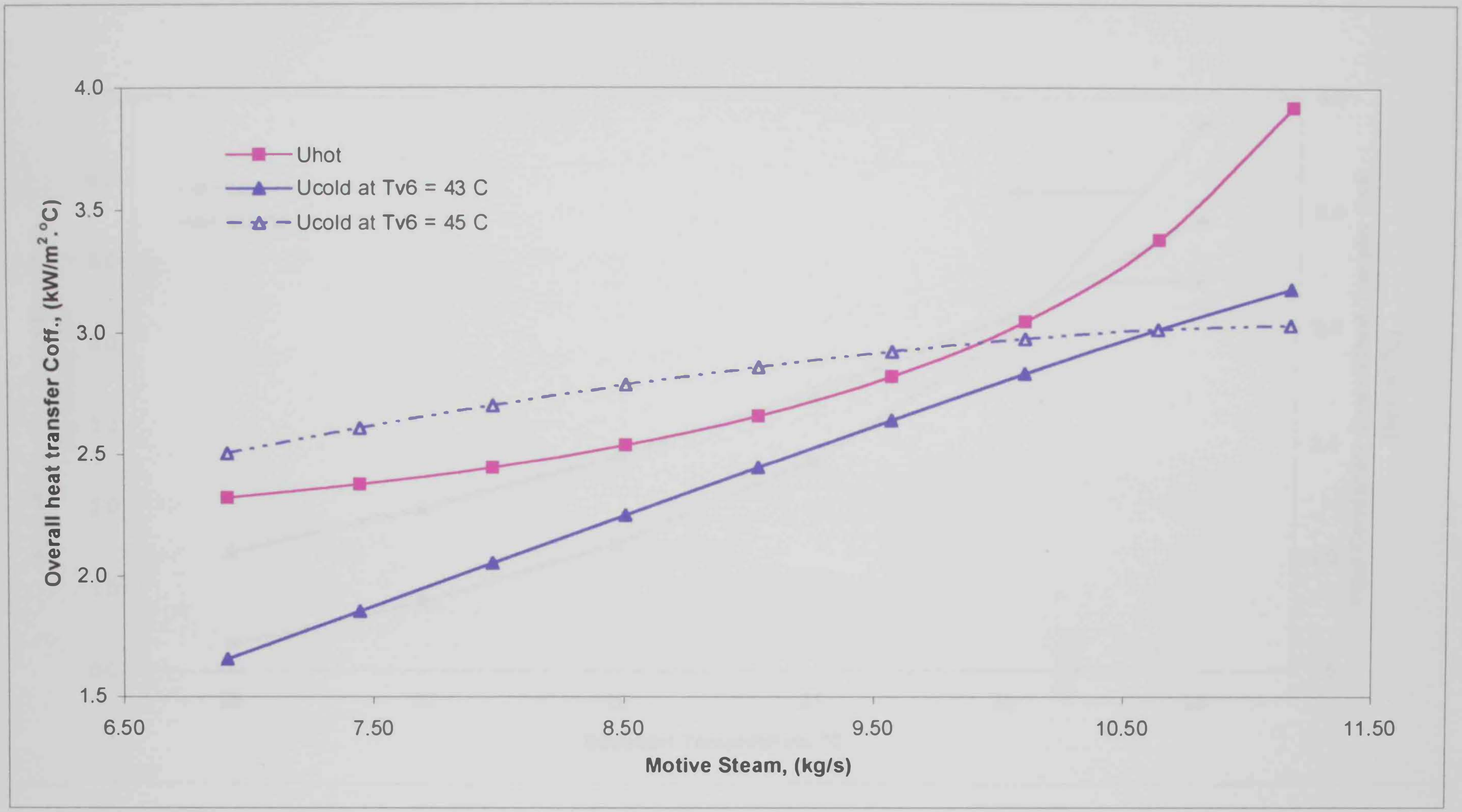


Figure 6-20: MEDNAR runs at different production rates for overall heat transfer coefficients as a function of motive steam

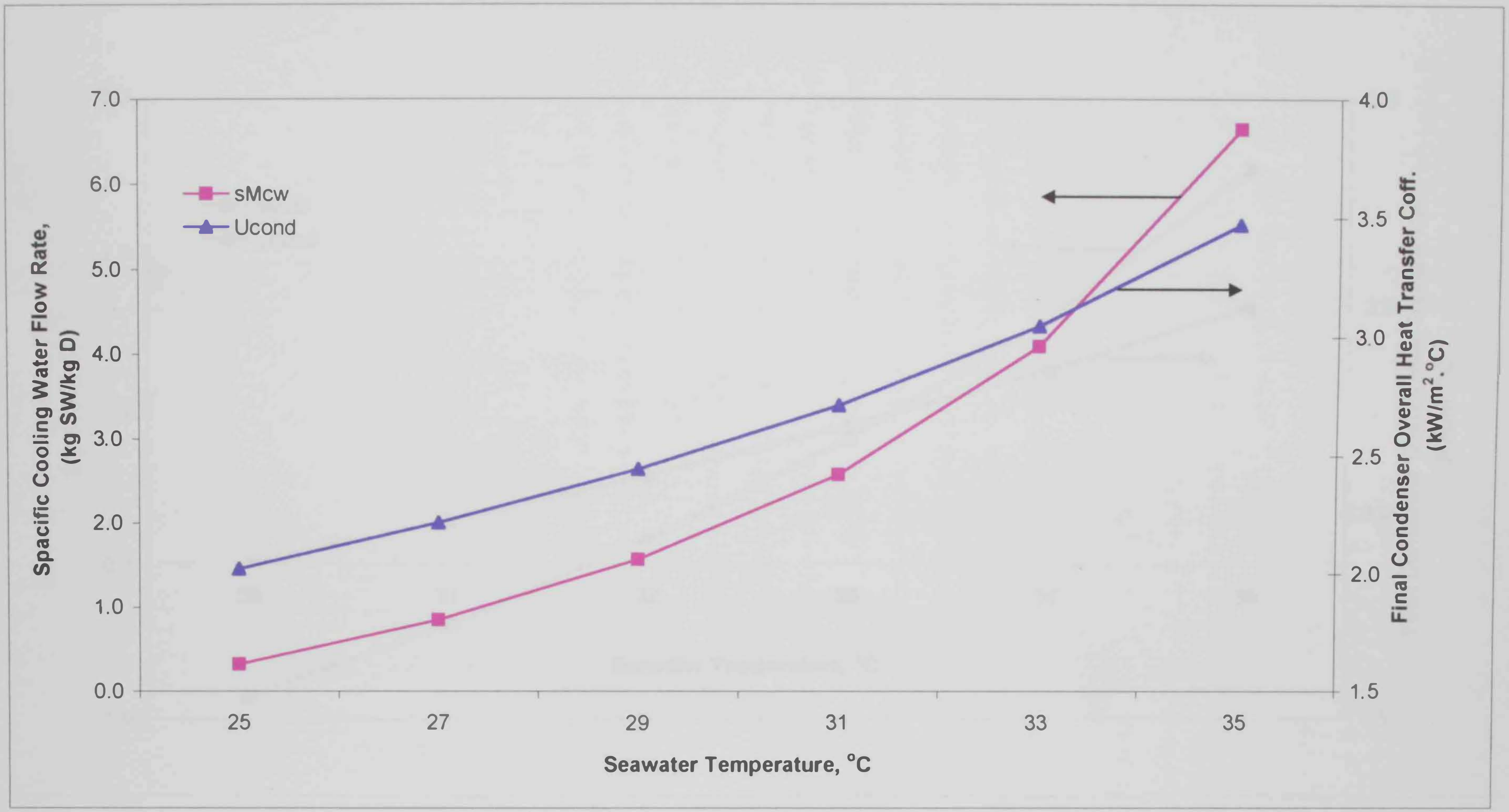


Figure 6-21: MEDNAR runs at 100% production rate for specific cooling water and condenser overall heat transfer coefficient as a function of seawater temperatures

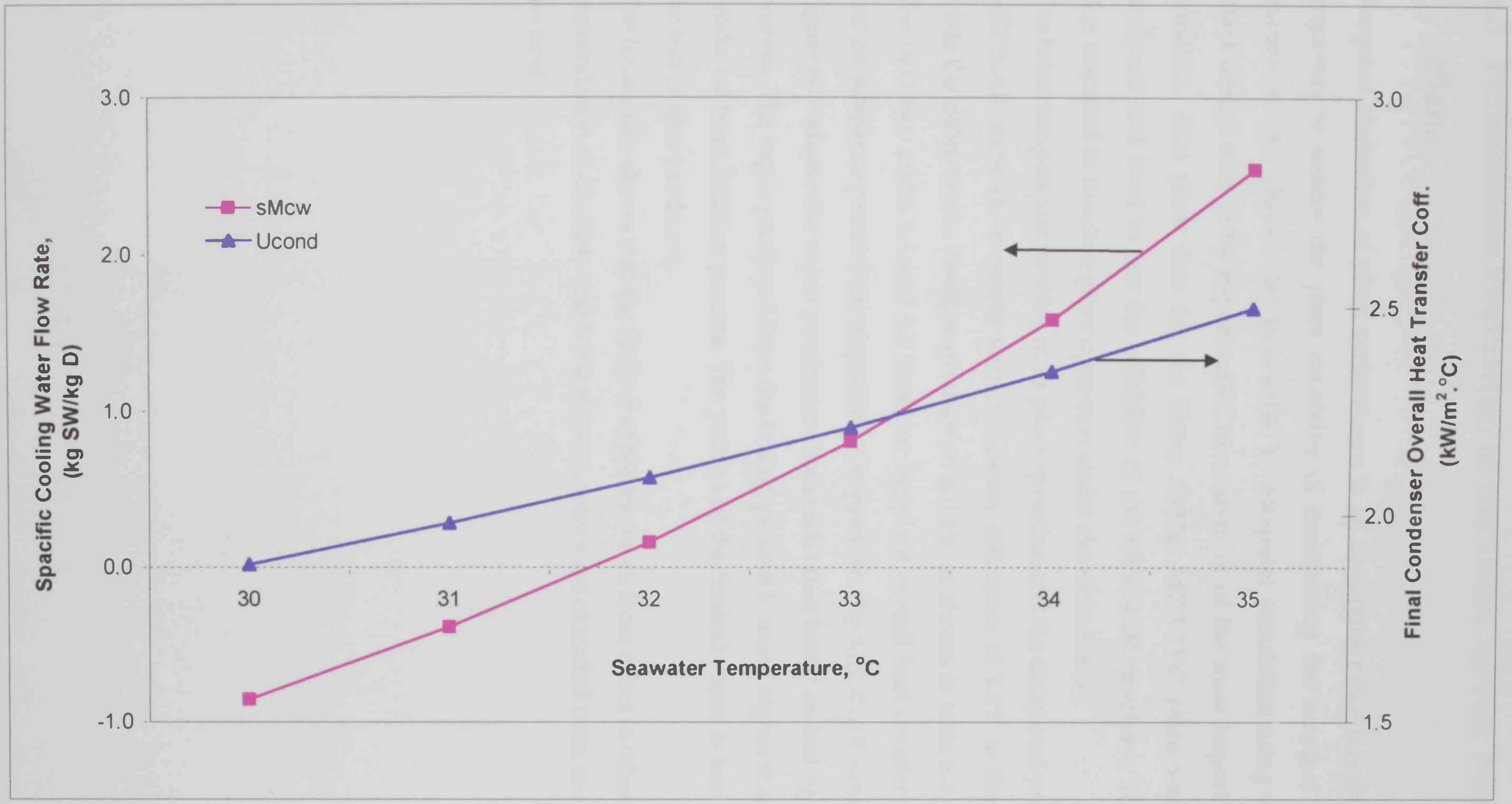


Figure 6-22: MEDNAR runs at 50% production rate for specific cooling water and condenser overall heat transfer coefficient as a function of seawater temperatures

6.3 Performance Evaluation of the Umm Al-Nar MED-TVC Plant

Frequent evaluation of plant performance is an important procedure for operation engineers to ensure the plant capability of maintaining the nominal capacity at minimum cost. This can be done either by computer simulation using well-known plant design data or by experimental measurement of the most important process variables. Real plant data for the Umm Al-Nar MED-TVC plant was collected, analyzed and used to test the reliability of the MEDNAR simulator. The software was executed to simulate plant operation under clean conditions.

The brine temperature profiles from plant operation and the computed profiles using MEDNAR show close results with maximum difference of 1.6% in the first effect while the difference in distillate production is 0.5%, as shown in table 6-3.

The next two tables 6-4 and 6-5 illustrate the plant overall heat transfer coefficients and performance parameters respectively, as predicted by MEDNAR simulator.

Figure 6-23 shows the vapor production from each effect based on boiling or flashing process. The vapor produced from the boiling process is much higher than the vapor produced from flashing process. The portion of the flashed vapor is less than 7% of the total vapor produced.

The figure also shows that the flashed vapor increases from effect to other, due to the accumulation of distillate and brine streams, which are cascaded from one effect to the next.

Table 6-3: Comparison of plant data and MEDNAR run

Variable	Plant Data	MEDNAR Simulator	Δ3
A- Brine temperature profile			
T1	62.5	61.5	1.6
T2	58.9	58.3	1.1
T3	55.2	55.0	0.4
T4	51.7	51.1	1.1
T5	48.1	47.7	0.7
T6	44.4	44.1	0.6
B- Other variables			
Mdistillate, kg/s	187.2	186.4	0.5
Mbrine, kg/s	468.1	469.0	0.2
Tbrine, °C	45.2	44.1	2.5

Where:

Δ3 = 100 * (MEDNAR run - Plant data) / plant data

Table 6-4: The plant overall heat transfer coefficients computed by MEDNAR

Equipment	Value	Unit
Hot effects, U_{HOT}	3.69	kW/m².°C
Cold effects, U_{COLD}	3.52	kW/m².°C
Final condenser, U_{FC}	3.78	kW/m².°C

Table 6-5: Plant performance parameters computed by MEDNAR

Performance parameters	Value	Unit
Performance ratio, PR	8.97	kg D/kg motive steam
Specific cooling water, sMcw	2.88	kg SW/kg D
Specific area, sA	310.7	m²/ kg D

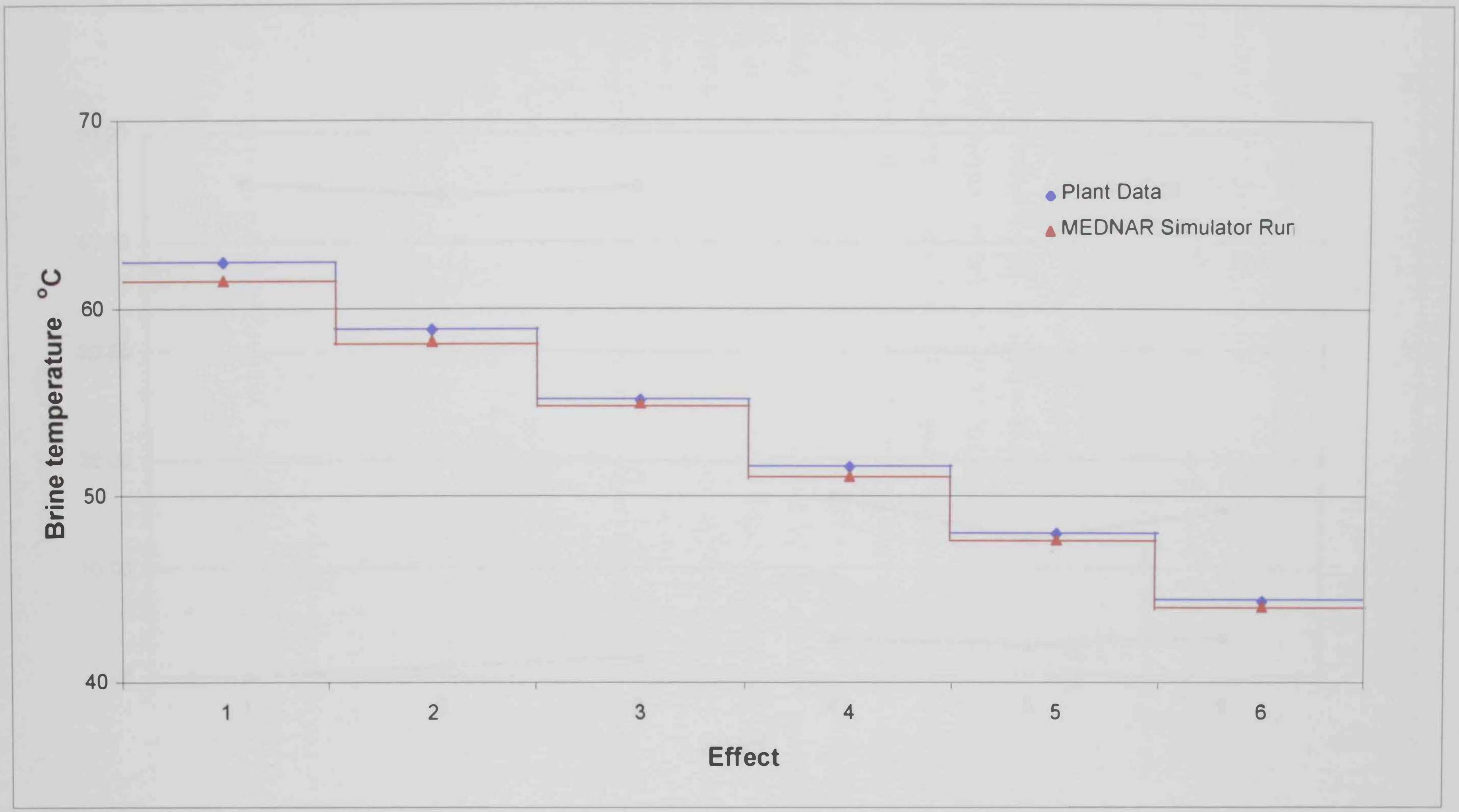


Figure 6-23: Brine temperature profile for plant data and MEDNAR

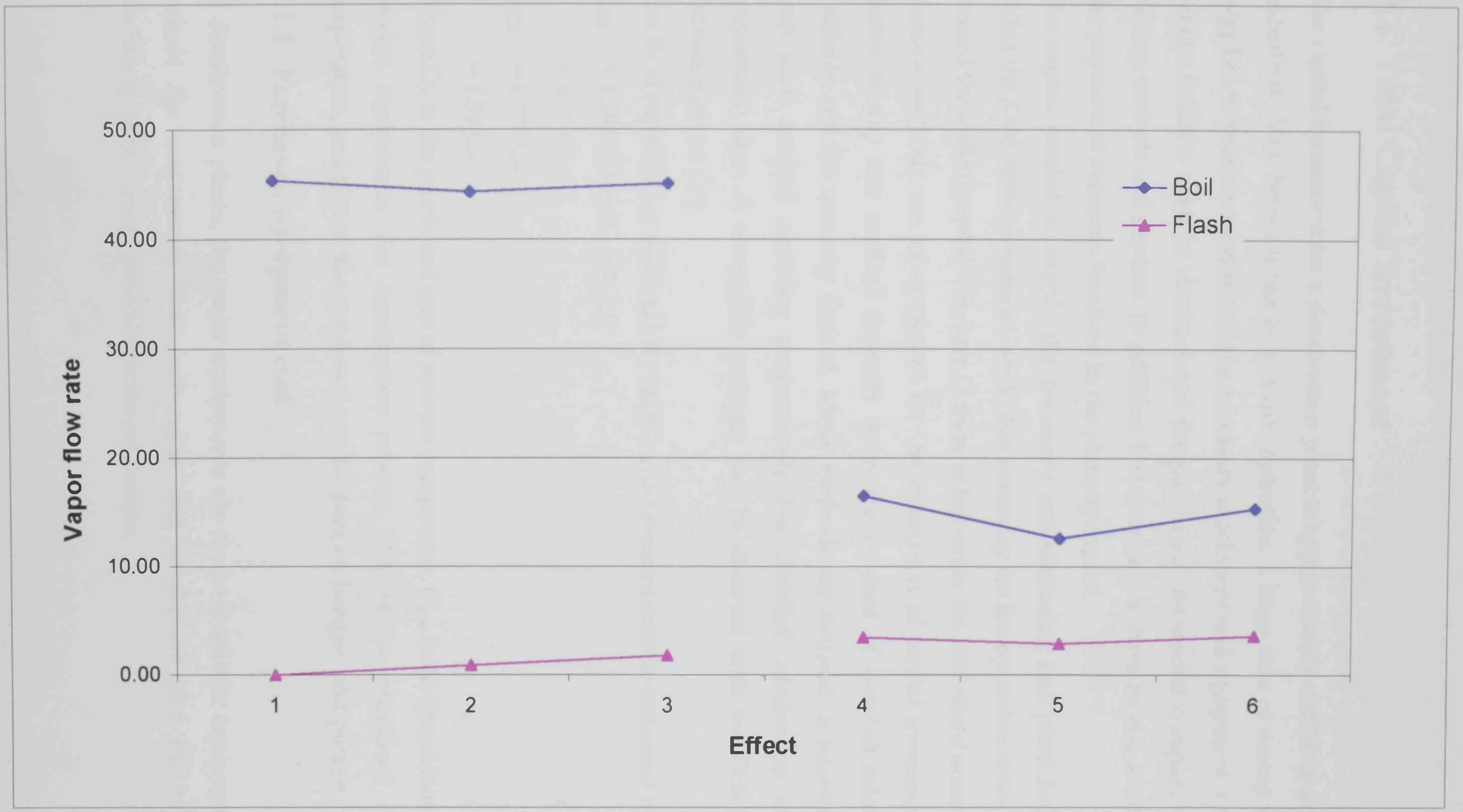


Figure 6-24: The boil and flash vapor flow rates (kg/s) production from effects as predicted from MEDNAR

7 ECONOMIC ANALYSIS

7.1 Total Capital Investment

The capital investment for a desalination plant is approximately similar to any other industrial plant. Before it can be put into operation, a large sum of money must be supplied to purchase and install the necessary machinery and equipment. Land and service facilities must be obtained and the plant must be erected complete with all piping, controls and services. In addition it is necessary to have money available for the payment of expenses involved in the plant operation.

The capital needed to supply the necessary manufacturing and plant facilities is called the *fixed capital investment*, while that necessary for the operation of the plant is termed the *working capital*. The sum of these is known as the *total capital investment*.

Various methods can be employed for the estimation of capital investment. The choice of any one method depends upon the amount of detailed information available and the accuracy desired. Many methods are outlined in reference [27], with each method requiring progressively less detailed information and less preparation time. A reasonable accuracy can be obtained with the *Detailed Item Estimate* method [27].

The fixed capital (C_{FCI}) and total capital (C_{TCI}) investments can be estimated by

$$C_{FCI} = 1.581\,(2.145\,C_{Pr} + C_{Inst})$$

$$= 4.75\,C_{Pr} \qquad (7\text{-}1)$$

$$C_{TCI} = 1.177\,C_{FCI}$$

$$= 5.59\,C_{Pr} \qquad (7\text{-}2)$$

Where C_{Pr} is the purchased cost of process equipments. C_{Inst} is the installation cost of process equipments. For desalination process, 75% of the purchased cost for evaporators, and 45% of the purchased cost for heat exchangers and pumps.

7.1.1 Purchased equipment cost

In desalination plants, the major equipments are the heat transfer equipments that include the evaporators, heat exchangers and condensers and the pumping machinery, which are in general centrifugal pumps.

7.1.1.1 Cost Index

Most available cost data for immediate use in cost estimates are based on conditions at some time due to changes in economic conditions; some method must therefore be used for updating cost data applicable at a past date to costs that are representative of conditions at a later time. This can be done by the use of cost index.

$$C_P = C_O \left(\frac{I_P}{I_O} \right) \tag{7-3}$$

Where C_P is the present cost. C_O is the original cost. I_P and I_O represent the index value at present time and original time, respectively.

7.1.2 The Annual Worth of a Present

In our calculations, we want to find the net annual water cost. Thus, the total capital cost must be put in *annual worth* form, which is an equal annual series of money for a stated period that is equivalent to the cash inflows and outflows at an interest rate.

$$A = P \left(\frac{e^{rN}(e^{r} - 1)}{e^{rN} - 1} \right) \tag{7-4}$$

Where r represents the annual interest rate, N is the number of years, A and P represent the annual and present equivalent amounts respectively [28].

7.2 Total Production Cost

Determination of the necessary capital investment is only part of a complete cost estimate. Another equally important part is the estimation of costs for operating the plant and delivering the product to the distributor. These costs can be grouped under the general heading of total production costs and general expenses. Manufacturing costs are known as operating or production costs. The annual cost basis is probably the best choice for estimation total cost. Total Production Cost (C_{TPC}) equals the sum of the manufacturing cost plus the general expense.

A. Manufacturing cost equals the sum of the following costs:
   1. Utilities (C_{Ut})
   2. Maintenance and repairs (C_{Mnt}); about 4% of the fixed capital investment
   3. Operation (C_{Oprt}); about 15% of the maintenance and repairs
   4. Laboratory charges (C_{LChrg}); about 2.25% of the total production cost

5. Direct supervision (C_{Dsup}); about 2.263% of the total production cost
6. Depreciation (C_{Dpr}); about 8% of the fixed capital investment
7. Local taxes (C_{Tax}); about 3% of the fixed capital investment
8. Insurance (C_{Insur}); about 1% of the fixed capital investment
9. Plant overhead (C_{PlOvh}); about 10% of the total production cost

B. General expenses equal the sum of the following costs:
   1. Administrative costs (C_{Admn}); about 3.75% of the total production cost
   2. Distribution & selling costs (C_{Sell}); about 5% of the total production cost

The general formula is

$$C_{TPC} = 1.629\ (C_{Ut} + 0.189\ C_{FCI}) \tag{7-5}$$

7.2.1 Utility cost

The cost for utilities such as steam, electricity, process cooling water and compressed air varies widely depending on the plant capacity and location. In desalination plants, the most critical economic calculation is the steam cost. The energy consumption in desalting water processes is one of the important parameters that dictate the choice of one desalination method over another and the final unit cost of desalted water. This cost accounting issue related to cogeneration plants is how to equitably split the total cost of owning and operating the plant between its main products, namely, electricity and heat. This is important since the way cost allocation is made will directly influence the cost of electricity generated as well as the cost of desalted water from the desalination plant. Allocating the cost of cogeneration between its two products in a fair and equitable manner not only affects the costs of water and electricity but also influences a host of products and services, which depend on these two products. Therefore, it is crucial to have a rational basis for pricing the two products.

Several methods for cost allocation have been developed; some of them allocate the total cost of owning and operating the plant among the two products direct without having to split the total cost into its components (direct method) and other methods allocate each cost component (e.g. manpower, material, fuel and capital depreciation) among the two products (indirect method). Assumptions regarding cost separation can also have an effect on the resulting product costs. This is why there are number of cogeneration cost allocation methods available [29].

Another method described by El-Nashar [30] in many papers is based on cost accounting using exergy as the commodity of value. A steam cycle with low temperature MED plant has been analyzed by Kamal [31]. Cost allocation methods were proposed in details by Darwish M. et al. [32] and Elsayed M. [33]. The method used in this thesis is called *prorating on the basis of power generated*, which has been mentioned by M. Elsayed. The steam cost is evaluated by the electricity loss in the turbine due to the low energy steam extracted to the desalination plant. The energy consumption and capital investment cost of steam turbine and high-pressure boiler are between water and electricity cost, the percentage is based on the electricity loss.

7.3 Net annual Water Cost

After annualizing the total investment cost by equation (7-4) we obtain the total production cost, which is calculated on annual base. The total yearly water production is calculated from the daily designed production and the plant running factor. The net annual water cost will be the total annual cost divided by the annual water production.

7.4 Economic Results and Discussion

A simple model has been adopted from previous equations to obtain the net annual water cost. The following assumptions used in this model are:

- The desalination plant life is 25 years
- The interest rate is 8%
- The plant running load factor is 0.9
- No local taxes
- The fuel cost is 1 $ per 1 GJ based on the local market

Note: this assumed fuel cost is equivalent to 6 $ per barrel of oil which is too low. A sensitivity analysis for different fuel cost has been done in this chapter.

7.4.1 Cost analysis

The model results give 1.09 $ for each m^3 of water. The annual production cost share is 73% while the annual capital cost share is 27% as shown in Figure 7-1. From the annual percentage cost break down analysis, the largest item is depreciation (23%). Purchased equipments, utility and maintenance come next with more than (10%)

each. Figure 7-2 illustrates the break down cost percentage. The cost break down prices is displayed in table 7-1.

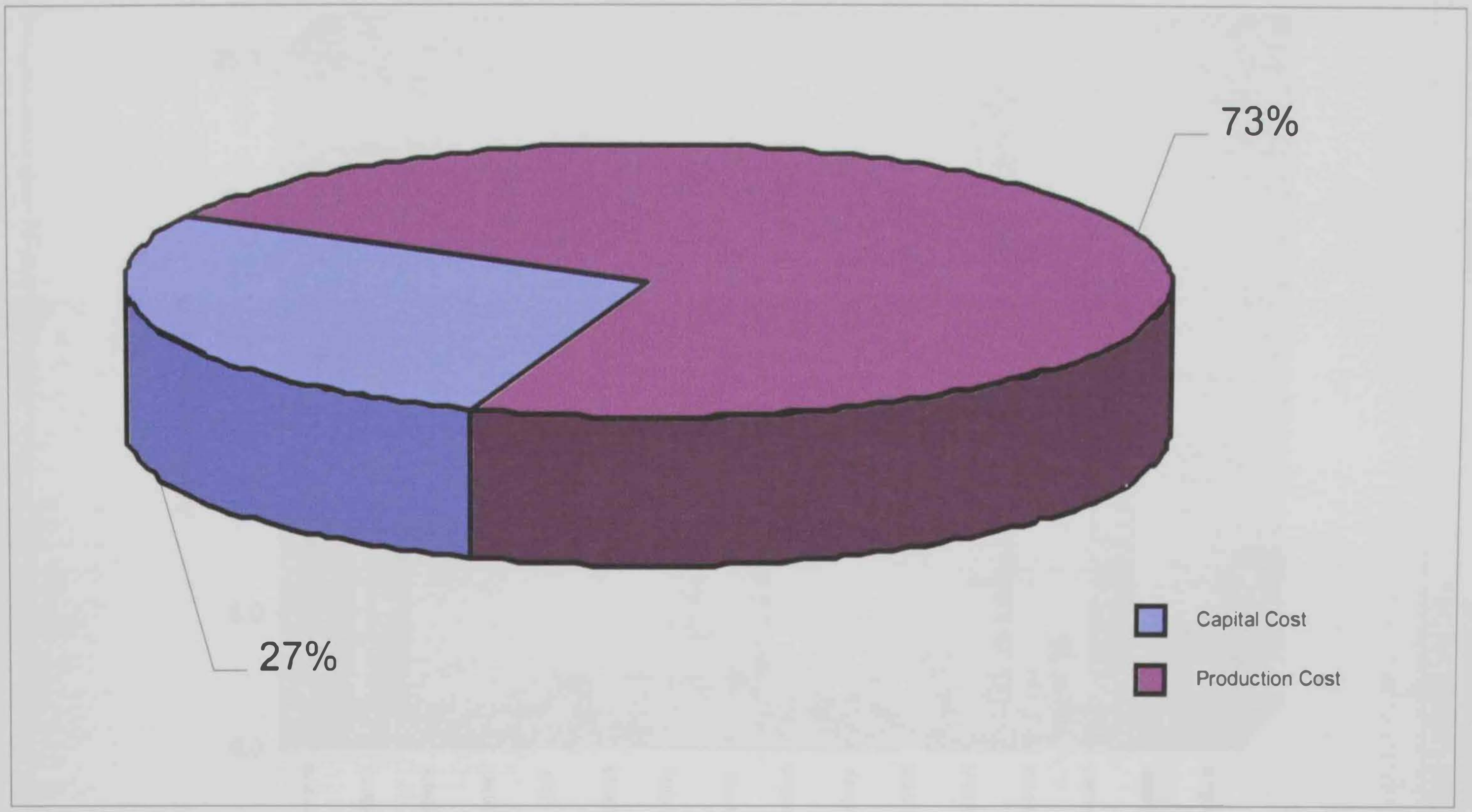


Figure 7-1: Annual share for capital cost and production cost from the net annual water cost

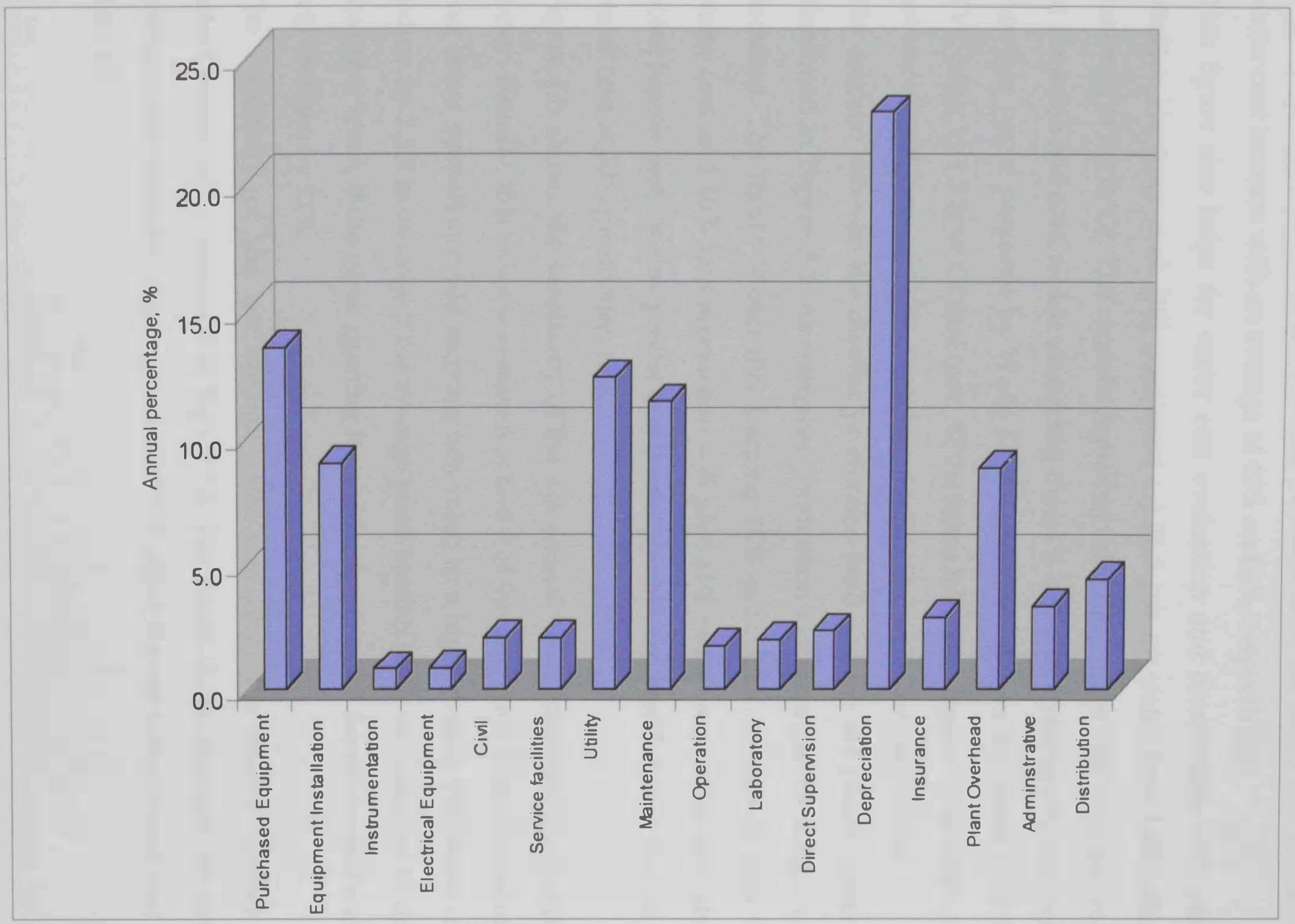


Figure 7-2: Annual cost percentage for the break down items

7.4.2 Cost sensitivity analysis

Figure 7-3 illustrates the steam cost and the net annual water cost based on different fuel cost rates. Both steam cost and net annual water cost increase linearly with increasing fuel cost. When the fuel cost increases 50%; the steam cost and net annual water cost increase with an average of 40% and 6%, respectively.

This figure also helps for easier cost evaluation and comparison with other cost studies. Hanbury et al. [10] mentioned a 1.31 $ per m^3 water for a MED plant with fuel cost of 2.8 $/GJ. The model adopted in this study gives 1.30 $ per m^3 water cost at the same fuel cost, which means less than 1% deviation between the two results.

Another paper prepared by Wade [2] illustrates 0.953 $per m^3 water for LT-MED-TVC plant at 1.5 $per GJ fuel cost. At the same fuel cost, Figure 7-3 shows 1.15 $ per m^3 water, which means 17% deviation observed between the two results.

The relation between the distiller production load and the net annual water cost is illustrated in Figure 7-4. As continues production increases the net annual cost will decrease. The figure shows that keeping 10% production more will give 9% less water cost and 10% less production will give 11% more water cost. The interesting point here is that the low production is uneconomic and could double the net annual water cost at 50% production rate.

Figure 7-5 shows the sensitivity of the net annual water cost with the distiller load factor. Usually, this value is assumed to be 0.9 of the year days. Planned maintenance and better operation could increase this value to a higher value. The water cost will reduce by 2.5% in average if the average plant availability is increased by 10 days. On the other hand, if the plant running factor is reduced to 0.8 the net annual water cost will increase by 11%.

The sensitivity of the net annual water cost from the annual operation and maintenance cost is observed in Figure 7-6. The figure shows that each 5% saving or losing in operation and maintenance cost will affect the net annual water cost by less than 1%.

Table 7-1: Net annual water cost break down with fuel price sensitivity

Plant Gross Output	MW	75	75	75
Auxiliary Loads	MW	7	7	7
Net Output	MW	68	68	68
Load	%	100	100	100
Extr Steam to Desal	T/h	84.25	84.25	84.25
Enthalpy of Desal Steam	kg/kJ			
Enthalpy of Return	kg/kJ			
Desal Production	MIGD	3.5	3.5	3.5
Desal Production	m^3/h	663	663	663
Desal Aux Loads	kW	980	980	980
Desal Plant Type		**MED-TVC**	**MED-TVC**	**MED-TVC**
Desal Plant Life	year	25	25	25
Desal Discount Rate	%	8	8	8
Desal Load Factor		0.9	0.9	0.9
Power Plant Type		**BPST**	**BPST**	**BPST**
Power Plant Life	year	25	25	25
Power Discount Rate	%	8	8	8
Power Plant Factor		0.9	0.9	0.9
H.P.B.Capital Cost	$	$2,650,000	$2,650,000	$2,650,000
Turbine Capital Cost	$	$7,150,000	$7,150,000	$7,150,000
Thermal Efficiency Turbine		0.4	0.4	0.4
Thermal Efficiency Boiler		0.92	0.92	0.92
Desal Capital Investment Cost				
Purchased Equipment Cost, Cpr	$	30,589,711	30,589,711	30,589,711
Equipment Installation Cost, Cinst	$	20,262,214	20,262,214	20,262,214
Instrumentation & Control Cost, Cinstr	$	1,770,194	1,770,194	1,770,194
Electrical Equipment Cost	$	1,836,553	1,836,553	1,836,553
Civil Cost	$	4,550,000	4,550,000	4,550,000
Service facilities Cost	$	4,550,000	4,550,000	4,550,000
TOTAL	**$**	**60,375,960**	**60,375,960**	**60,375,960**
Desal Production Cost				
Fuel Cost	**$/y**	**2,617,249**	**4,424,258**	**6,231,267**
Fuel Price	$/GJ	1.00	1.80	2.60
Extraction Steam Cost	$/T	1.07	1.81	2.56
Manufacturing Cost	**$/y**	**10,976,186**	**10,976,186**	**10,976,186**
Maintenance & Repair Cost	$/y	2,415,038	2,415,038	2,415,038
Operation Cost	$/y	362,256	362,256	362,256
Laboratory Charges	$/y	418,244	418,244	418,244
Direct Supervision Cost	$/y	487,951	487,951	487,951
Depreciation	$/y	4,830,077	4,830,077	4,830,077
Insurance	$/y	603,760	603,760	603,760
Plant Overhead	$/y	1,858,861	1,858,861	1,858,861
General Expenses Costs	**$/y**	**16,846,443**	**18,653,451**	**20,460,460**
Administrative Cost	$/y	697,073	697,073	697,073
Distribution & Selling Cost	$/y	929,431	929,431	929,431
TOTAL	**$/y**	**15,219,939**	**17,026,948**	**18,833,956**
Total Annual Cost	**Dhs/y**	**20,875,886**	**22,682,894**	**24,489,903**
Total Annual Water Production	**m^3/y**	**5226138.364**	**5226138.364**	**5226138.364**
Net Annual Water Cost	**Dhs/m^3**	**3.99**	**4.34**	**4.69**
Net Annual Water Cost	**Dhs/kIG**	**18.16**	**19.73**	**21.30**
Net Annual Water Cost	**$\$/m^3$**	**1.09**	**1.18**	**1.28**

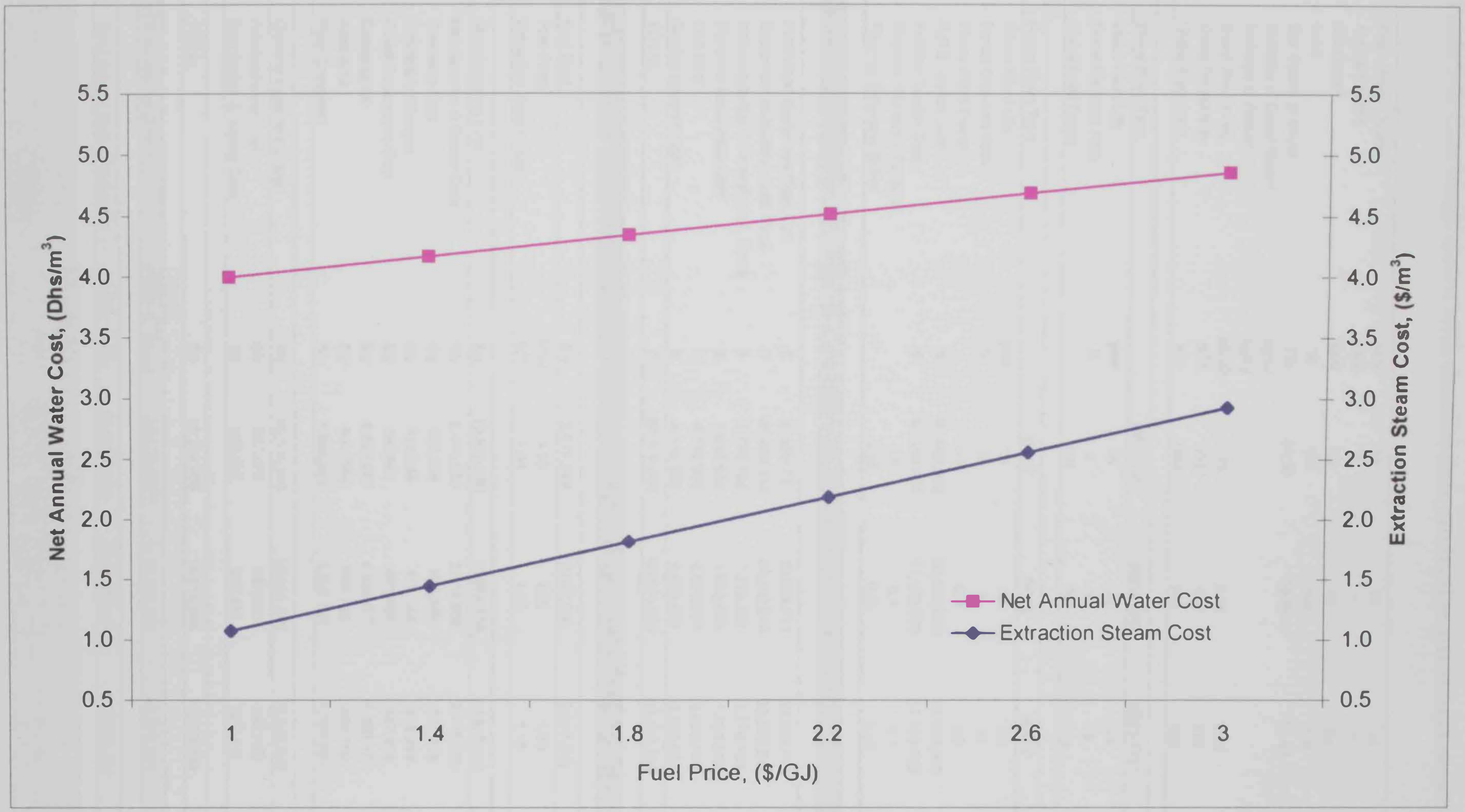


Figure 7-3: Net annual water cost and Extraction steam cost as a function of fuel price

Table 7-2: Cost break down with desalination production plant load sensitivity

Plant Gross Output	MW	75	75	75	75
Auxiliary Loads	MW	7	7	7	7
Net Output	MW	68	68	68	68
Load	%	100	100	100	100
Extr Steam to Desal	T/h	55.32	66.89	78.46	90.04
Enthalpy of Desal Steam	kg/kJ				
Enthalpy of Return	kg/kJ				
Desal Production	MIGD	1.75	2.45	3.15	3.85
Desal Production	m^3/h	331	464	597	729
Desal Aux Loads	kW	980	980	980	980
Desal Plant Type		MED-TVC	MED-TVC	MED-TVC	MED-TVC
Desal Plant Life	year	25	25	25	25
Desal Discount Rate	%	8	8	8	8
Desal Load Factor		0.9	0.9	0.9	0.9
Power Plant Type		BPST	BPST	BPST	BPST
Power Plant Life	year	25	25	25	25
Power Discount Rate	%	8	8	8	8
Power Plant Factor		0.9	0.9	0.9	0.9
H.P.B.Capital Cost	$	$2,650,000	$2,650,000	$2,650,000	$2,650,000
Turbine Capital Cost	$	$7,150,000	$7,150,000	$7,150,000	$7,150,000
Thermal Efficiency Turbine		0.4	0.4	0.4	0.4
Thermal Efficiency Boiler		0.92	0.92	0.92	0.92
Desal Capital Investment Cost					
Purchased Equipment Cost, Cpr	$	30,589,711	30,589,711	30,589,711	30,589,711
Equipment Installation Cost, Cinst	$	20,262,214	20,262,214	20,262,214	20,262,214
Instrumentation & Control Cost, Cinstr	$	1,770,194	1,770,194	1,770,194	1,770,194
Electrical Equipment Cost	$	1,836,553	1,836,553	1,836,553	1,836,553
Civil Cost	$	4,550,000	4,550,000	4,550,000	4,550,000
Service facilities Cost	$	4,550,000	4,550,000	4,550,000	4,550,000
TOTAL	$	60,375,960	60,375,960	60,375,960	60,375,960
Desal Production Cost					
Fuel Cost	$/y	2,617,249	2,617,249	2,617,249	2,617,249
Fuel Price	$/GJ	1.00	1.00	1.00	1.00
Extraction Steam Cost	$/T	1.64	1.35	1.15	1.00
Manufacturing Cost	$/y	10,976,186	10,976,186	10,976,186	10,976,186
Maintenance & Repair Cost	$/y	2,415,038	2,415,038	2,415,038	2,415,038
Operation Cost	$/y	362,256	362,256	362,256	362,256
Laboratory Charges	$/y	418,244	418,244	418,244	418,244
Direct Supervision Cost	$/y	487,951	487,951	487,951	487,951
Depreciation	$/y	4,830,077	4,830,077	4,830,077	4,830,077
Insurance	$/y	603,760	603,760	603,760	603,760
Plant Overhead	$/y	1,858,861	1,858,861	1,858,861	1,858,861
General Expenses Costs	$/y	16,846,443	16,846,443	16,846,443	16,846,443
Administrative Cost	$/y	697,073	697,073	697,073	697,073
Distribution & Selling Cost	$/y	929,431	929,431	929,431	929,431
TOTAL	$/y	15,219,939	15,219,939	15,219,939	15,219,939
Total Annual Cost	Dhs/y	20,875,885	20,875,885	20,875,885	20,875,885
Total Annual Water Production	m^3/y	2613068.182	3658295.455	4703522.727	5748750
Net Annual Water Cost	Dhs/m^3	7.99	5.71	4.44	3.63
Net Annual Water Cost	Dhs/kIG	36.31	25.94	20.17	16.51
Net Annual Water Cost	$\$/m^3$	2.18	1.55	1.21	0.99

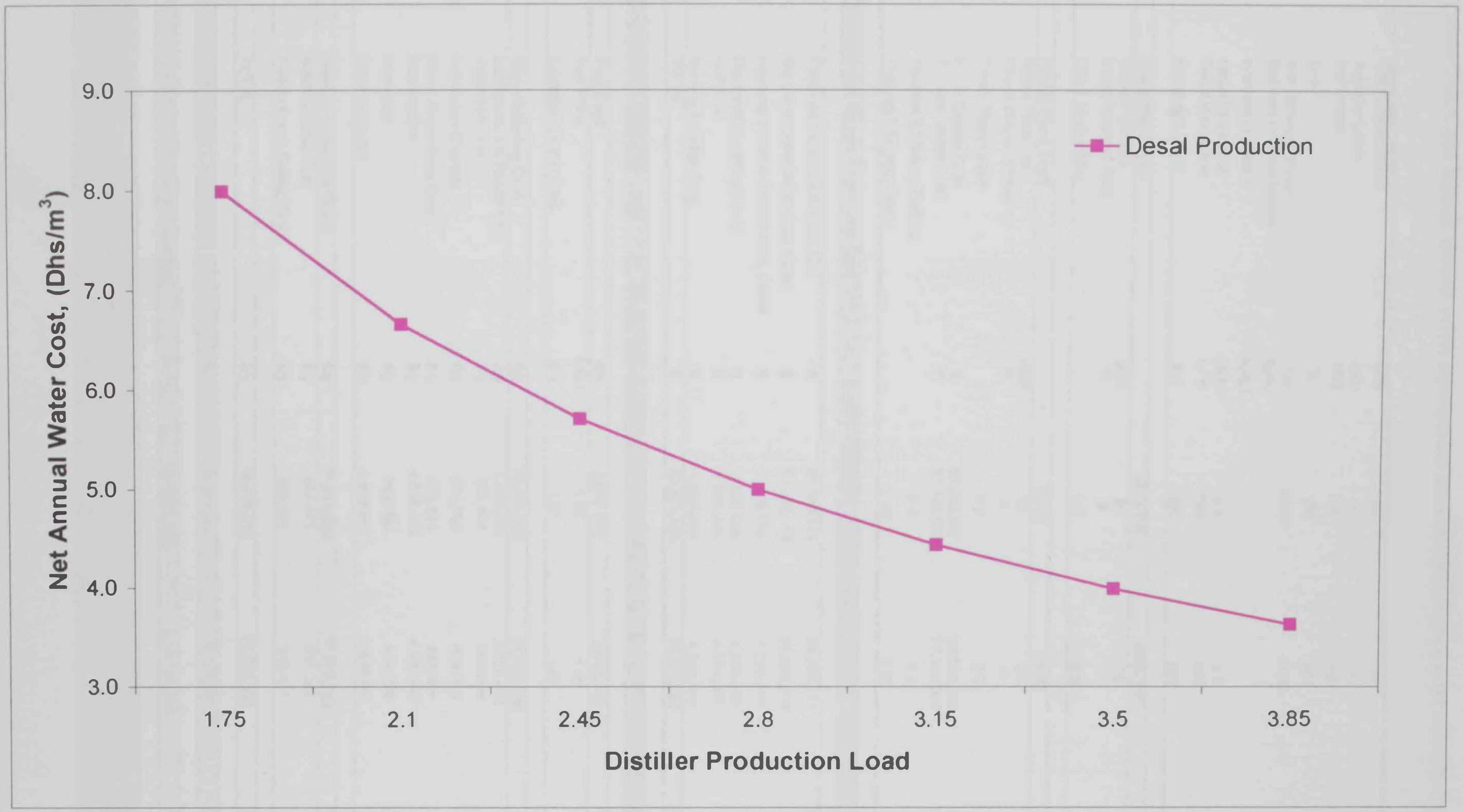


Figure 7-4: Net annual water cost as a function of distiller production load

Table 7-3: Cost break down with operation and maintenance cost sensitivity

Plant Gross Output	MW	75	75	75
Auxiliary Loads	MW	7	7	7
Net Output	MW	68	68	68
Load	%	100	100	100
Extr Steam to Desal	T/h	84.25	84.25	84.25
Enthalpy of Desal Steam	kg/kJ			
Enthalpy of Return	kg/kJ			
Desal Production	MIGD	3.5	3.5	3.5
Desal Production	m^3/h	663	663	663
Desal Aux Loads	kW	980	980	980
Desal Plant Type		**MED-TVC**	**MED-TVC**	**MED-TVC**
Desal Plant Life	year	25	25	25
Desal Discount Rate	%	8	8	8
Desal Load Factor		0.9	0.9	0.9
Power Plant Type		**BPST**	**BPST**	**BPST**
Power Plant Life	year	25	25	25
Power Discount Rate	%	8	8	8
Power Plant Factor		0.9	0.9	0.9
H.P.B.Capital Cost	$	$2,650,000	$2,650,000	$2,650,000
Turbine Capital Cost	$	$7,150,000	$7,150,000	$7,150,000
Thermal Efficiency Turbine		0.4	0.4	0.4
Thermal Efficiency Boiler		0.92	0.92	0.92
Desal Capital Investment Cost				
Purchased Equipment Cost, Cpr	$	30,589,711	30,589,711	30,589,711
Equipment Installation Cost, Cinst	$	20,262,214	20,262,214	20,262,214
Instrumentation & Control Cost, Cinstr	$	1,770,194	1,770,194	1,770,194
Electrical Equipment Cost	$	1,836,553	1,836,553	1,836,553
Civil Cost	$	4,550,000	4,550,000	4,550,000
Service facilities Cost	$	4,550,000	4,550,000	4,550,000
TOTAL	**$**	**60,375,960**	**60,375,960**	**60,375,960**
Desal Production Cost				
Fuel Cost	**$/y**	**2,617,249**	**2,617,249**	**2,617,249**
Fuel Price	$/GJ	1.00	1.00	1.00
Extraction Steam Cost	$/T	1.07	1.07	1.07
Manufacturing Cost	**$/y**	**10,181,528**	**10,976,186**	**11,799,826**
Maintenance & Repair Cost	$/y	1,932,031	2,415,038	2,898,046
Operation Cost	$/y	231,844	362,256	521,648
Laboratory Charges	$/y	334,595	418,244	501,892
Direct Supervision Cost	$/y	390,361	487,951	585,541
Depreciation	$/y	4,830,077	4,830,077	4,830,077
Insurance	$/y	603,760	603,760	603,760
Plant Overhead	$/y	1,858,861	1,858,661	1,858,861
General Expenses Costs	**$/y**	**17,670,082**	**17,670,082**	**17,670,082**
Administrative Cost	$/y	697,073	697,073	697,073
Distribution & Selling Cost	$/y	929,431	929,431	929,431
TOTAL	**$/y**	**14,425,280**	**15,219,939**	**16,043,578**
Total Annual Cost	**Dhs/y**	**20,081,227**	**20,875,885**	**21,699,524**
Total Annual Water Production	**M^3/y**	**5226138.364**	**5226136.364**	**5226136.364**
Net Annual Water Cost	**Dhs/m^3**	**3.84**	**3.99**	**4.15**
Net Annual Water Cost	**Dh/kIG**	**17.47**	**18.16**	**18.87**
Net Annual Water Cost	**$\$/m^3$**	**1.05**	**1.09**	**1.13**

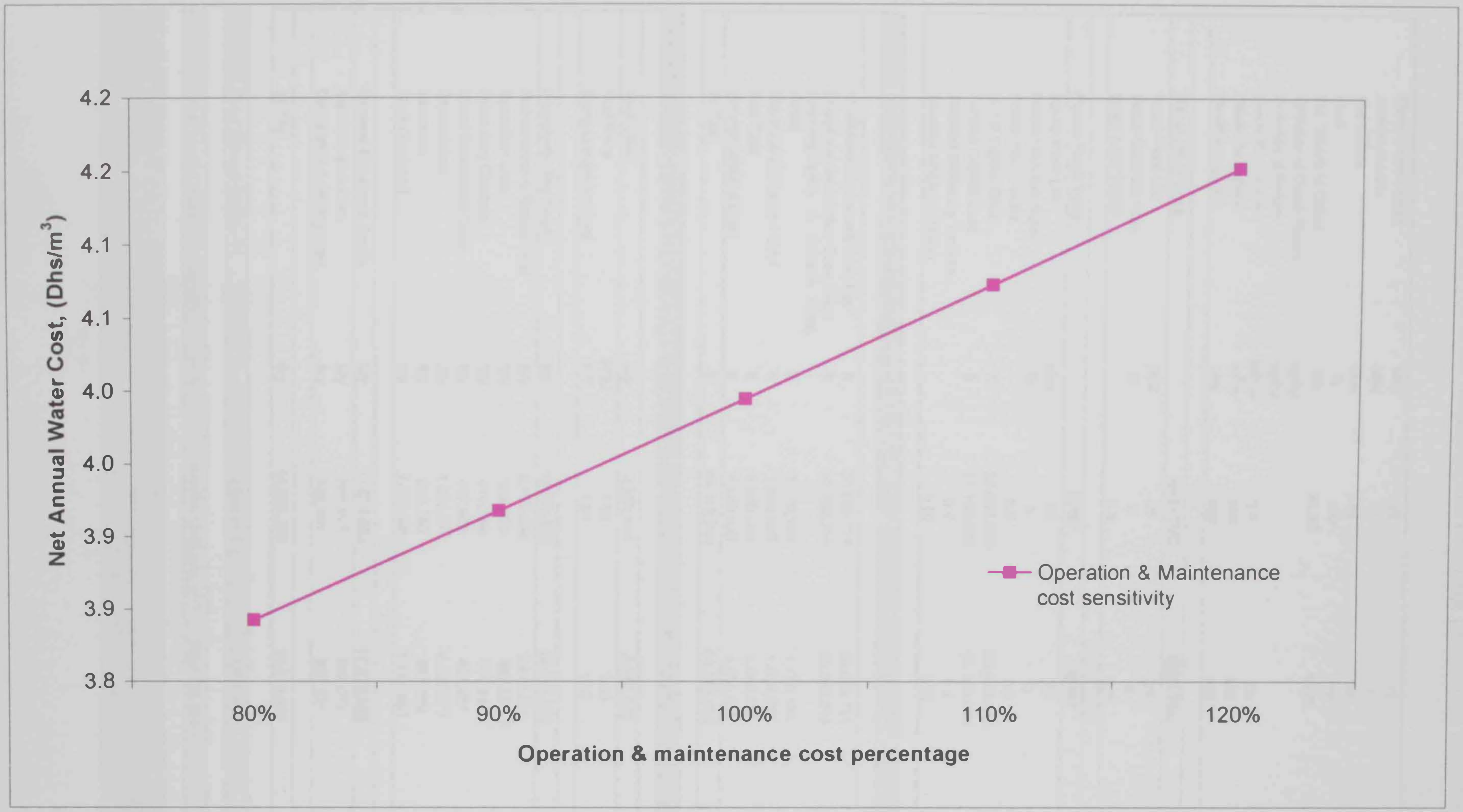


Figure 7-5: Net annual water cost as a function of operation and maintenance cost

Table 7-4: Cost break down with desalination running plant load sensitivity

Plant Gross Output	MW	75	75	75
Auxiliary Loads	MW	7	7	7
Net Output	MW	68	68	68
Load	%	100	100	100
Extr Steam to Desal	T/h	84.25	84.25	84.25
Enthalpy of Desal Steam	kg/kJ			
Enthalpy of Return	kg/kJ			
Desal Production	MIGD	3.5	3.5	3.5
Desal Production	m^3/h	663	663	663
Desal Aux Loads	kW	980	980	980
Desal Plant Type		**MED-TVC**	**MED-TVC**	**MED-TVC**
Desal Plant Life	year	25	25	25
Desal Discount Rate	%	8	8	8
Desal Load Factor		0.8	0.9	1
Power Plant Type		**BPST**	**BPST**	**BPST**
Power Plant Life	year	25	25	25
Power Discount Rate	%	8	8	8
Power Plant Factor		0.9	0.9	0.9
H.P.B.Capital Cost	$	$2,650,000	$2,650,000	$2,650,000
Turbine Capital Cost	$	$7,150,000	$7,150,000	$7,150,000
Thermal Efficiency Turbine		0.4	0.4	0.4
Thermal Efficiency Boiler		0.92	0.92	0.92

Desal Capital Investment Cost				
Purchased Equipment Cost, Cpr	$	30,589,711	30,589,711	30,589,711
Equipment Installation Cost, Cinst	$	20,262,214	20,262,214	20,262,214
Instrumentation & Control Cost, Cinstr	$	1,770,194	1,770,194	1,770,194
Electrical Equipment Cost	$	1,836,553	1,836,553	1,836,553
Civil Cost	$	4,550,000	4,550,000	4,550,000
Service facilities Cost	$	4,550,000	4,550,000	4,550,000
TOTAL	**$**	**60,375,960**	**60,375,960**	**60,375,960**

Desal Production Cost				
Fuel Cost	**$/y**	**2,326,444**	**2,617,249**	**2,908,055**
Fuel Price	$/GJ	1.00	1.00	1.00
Extraction Steam Cost	$/T	1.07	1.07	1.07
Manufacturing Cost	**$/y**	**10,976,186**	**10,976,186**	**10,976,186**
Maintenance & Repair Cost	$/y	2,415,038	2,415,038	2,415,038
Operation Cost	$/y	362,256	362,256	362,256
Laboratory Charges	$/y	418,244	418,244	418,244
Direct Supervision Cost	$/y	487,951	487,951	487,951
Depreciation	$/y	4,830,077	4,830,077	4,830,077
Insurance	$/y	603,760	603,760	603,760
Plant Overhead	$/y	1,858,861	1,858,861	1,858,861
General Expenses Costs	**$/y**	**17,670,082**	**17,670,082**	**17,670,082**
Administrative Cost	$/y	697,073	697,073	697,073
Distribution & Selling Cost	$/y	929,431	929,431	929,431
TOTAL	**$/y**	**14,929,134**	**15,219,939**	**15,510,745**

Total Annual Cost	**Dhs/y**	**20,585,080**	**20,875,885**	**21,166,691**
Total Annual Water Production	**m^3/y**	**4645454.545**	**5226136.364**	**5806818.182**
Net Annual Water Cost	**Dhs/m^3**	**4.43**	**3.99**	**3.65**
Net Annual Water Cost	**Dh/kIG**	**20.14**	**18.16**	**16.57**
Net Annual Water Cost	**$\$/m^3$**	**1.21**	**1.09**	**0.99**

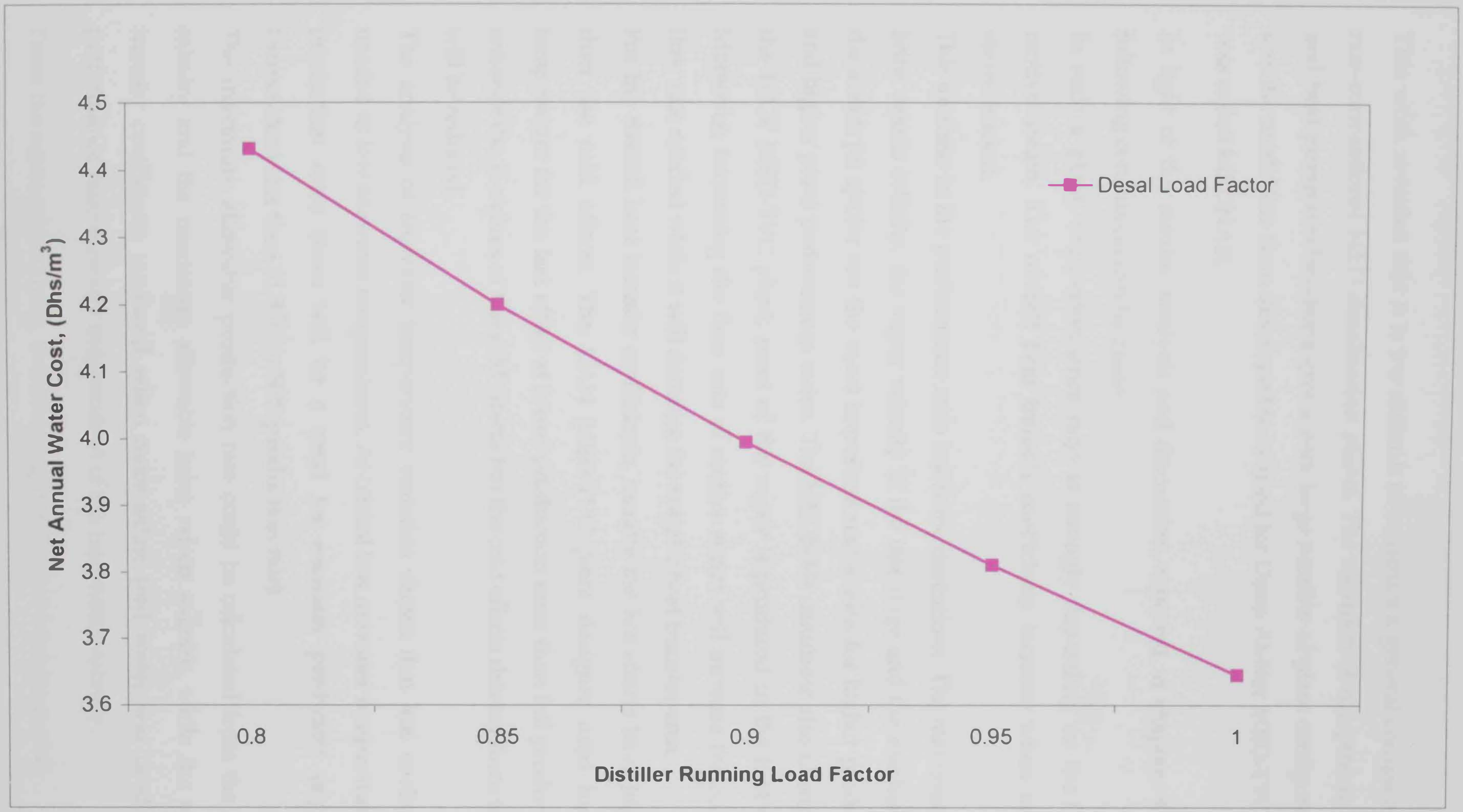


Figure 7-6: Net annual water cost as a function of distiller running load factor

8 CONCLUSIONS

This work revealed that it is too difficult to construct a general computer model for non-conventional MED desalination plants. The variation of equipment, flow sheets and heat pump combinations give a very large number of plant configurations. So, a specific model has been developed to be used for Umm Al-Nar MED-TVC plants and was called MEDNAR.

In light of the results, analysis and discussion, reported in chapters 6 and 7, the following conclusion can be made:

In such a plant, the performance ratio is strongly depending on the flow rate of motive steam. The overall heat transfer coefficients increase when more motive steam is used.

This increase in the performance ratio has some limitations. The maximum allowable brine rejects salinity, the vapor velocity in the last stage and the mechanical size of the steam jet ejector are the most important restrictions for higher production rates and higher plant performance ratios. The MEDNAR simulator also illustrates that in the UAN MED-TVC plant, most of the vapor is produced in the first two effects. Moreover, increasing the flow rate of motive steam will increase the cooling water flow rate needed while it will decrease the specific heat transfer area.

For the overall heat transfer coefficients, usually the hot effects have higher values than the cold effects. The UAN MED-TVC plant designer used higher vapor temperature for the last effect at lower production rates than full production rate to enhance the U-values of the cold effects but the cold effects driving force temperature will be reduced.

The analysis of seawater temperature variation shows that less cooling water is needed at low seawater temperatures. At critical low seawater temperatures and low production rates there will be a need for seawater pre-heater (e.g. seawater temperature less then 31.8 °C at 50% production rate).

The maximum allowable production rate could be calculated from the feed water salinity and the maximum allowable brine rejects salinity, while the overall heat transfer coefficients increased when more saline feed water was used. The plant performance ratio is nearly independent of the feed water salinity.

From the economic analysis, the following points have been proposed:

- The primary energy consumption of any desalination process has to be valued according to the international fuel price. This is to have real cost evaluation and a unified water cost base for clear and better cost analysis.
- The annual production cost share of the net annual water cost is higher than the annual capital cost. The share ratio is around (7:3).
- From the annual cost break down analysis, the depreciation cost has the higher contribution with 23%. Purchased equipment, utility and maintenance are coming next with more than 10% each.
- The net annual water cost could double if the plant is continuously operated at a partial load of 50%.
- The annual plant availability has a direct relation with the net annual water cost. The average running factor is 0.9. Four running days more or less on the average have an impact of 1% on the net annual water cost.
- The operation and maintenance cost deviation of 5% have an impact of 1% on the net annual water cost.
- From the last two points, proper maintenance and operation could give higher plant availability. The relation between these two points needs more optimization to minimize the water unit cost.

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APPENDIX A

Thermal and Physical Properties

The billing point elevation (BPE)
It depends on both the brine salinity X and the brine temperature T. the BPE in K given by

$$BPE = \frac{X*T^2}{13832}\left[\begin{array}{l} 1+1.373*10^{-3}*T-2.72*10^{-3}*\sqrt{X}*T+17.86*X- \\ 1.52*10^{-2}*X*T*\left(\dfrac{T-225.9}{T-236}\right)-\dfrac{2583*X*(1-X)}{T} \end{array}\right]$$

Where

X = salt concentration, wt.frac

T = Temperature, K

The above equation is valid over the following ranges: 20000 < X < 160000 ppm, 20 < T < 180 °C.

Vapor pressure of saturated water
$P = 10.171246 – 0.6167302\ T + 1.832249x10^{2}\ T^{-2} – 1.77376x10^{-4}\ T^{3} + 1.47068x10^{-6}\ T^{4}$
Where P is kPa and T is °C

Saturation temperature
T = (42.6776 – 3892.7/(ln(P/1000)-9.48654)) – 273.15
Where P is kPa and T is °C

Specific volume of water vapor
$V = 163.3453019 – 8.041421773\ T + 0.171021164\ T^{2} – 0.001878124x10^{-3}\ T^{5}$
Where V is m3/kg and T is °C

Latent heat
$\lambda = 2499.5698 – 2.204864\ T – 1.596x10^{-3}\ T^{2}$
Where T is the saturation temperature in °C and λ is the latent heat in kJ/kg

Specific heat of water at constant pressure
$Cp = [A+BT+CT^{2}+DT^{3}]\ x10^{-3}$
T = temperature, °C
X = water salinity, g/kg
$A = 4206.8 – 6.6197\ X + 1.2288x10^{-2}\ X^{2}$
$B = -1.1262 + 5.4178x10^{-2}\ X - 2.2719x10^{-4}\ X^{2}$
$C = 1.2026x10^{-2} - 5.3566x10^{-4}\ X + 1.8906x10^{-6}\ X^{2}$
$D = 6.87774x10^{-7} + 1.517x10^{-6}\ X - 4.4268x10^{-9}\ X^{2}$

APPENDIX B

Umm Al-Nar Plant Information

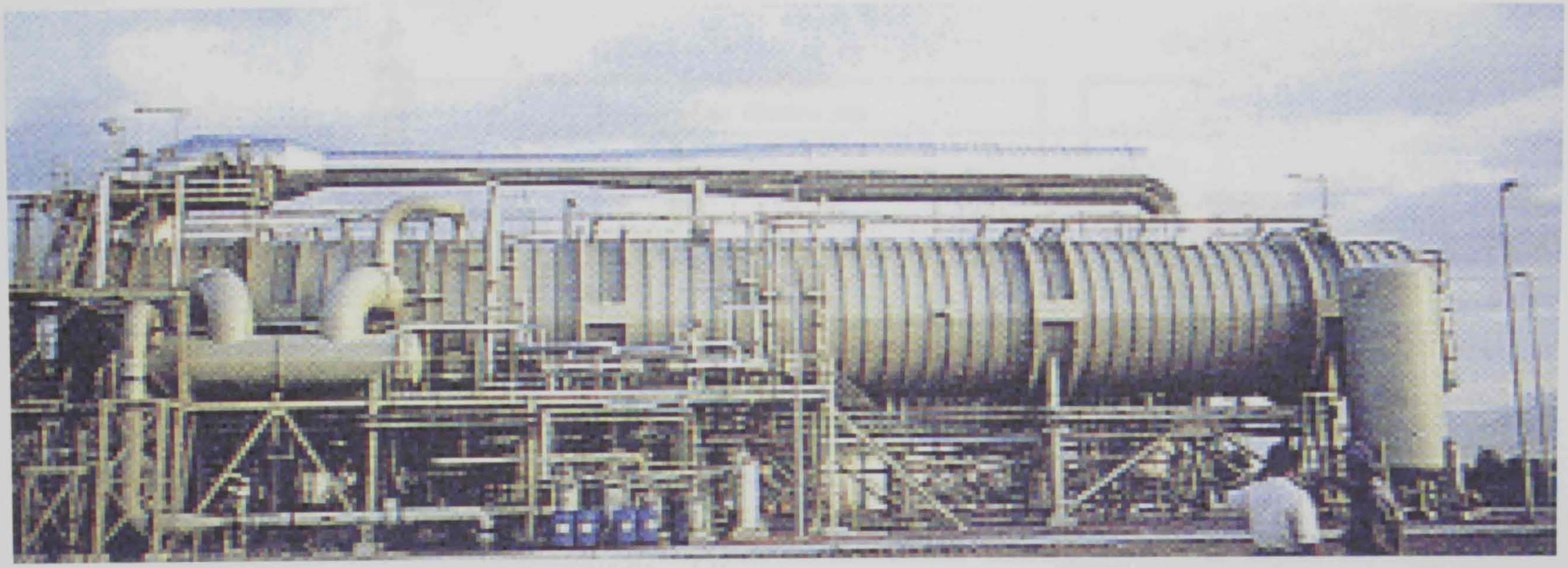


Figure A-1: MED-TVC unit

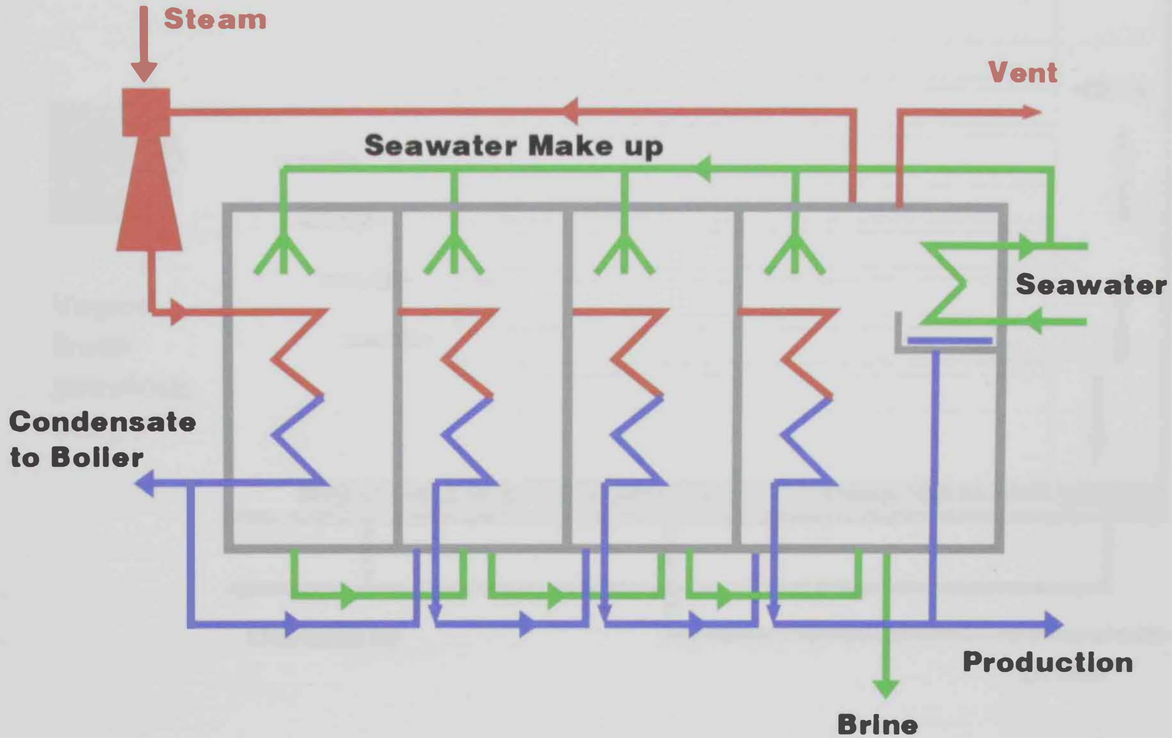


Figure A-2: MED-TVC processes

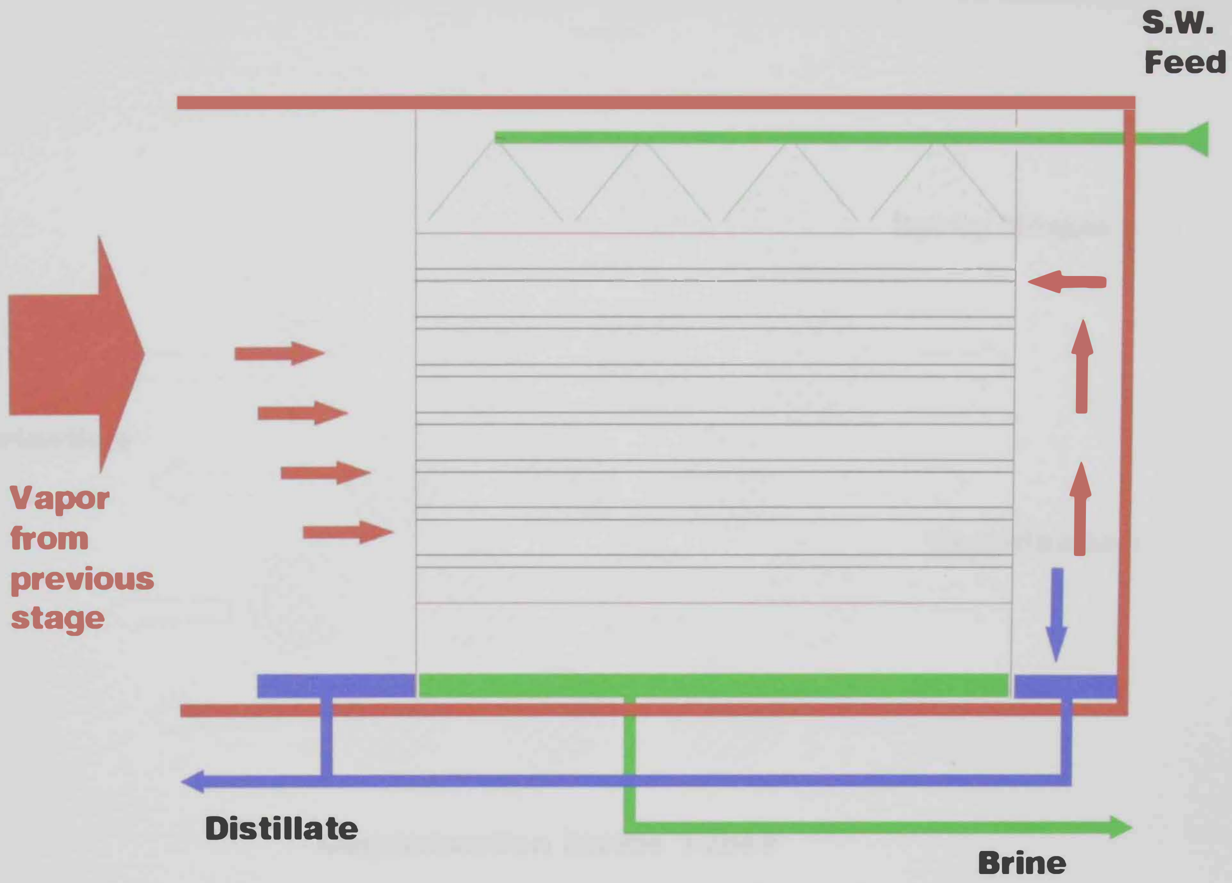


Figure A-3: Evaporator stage process

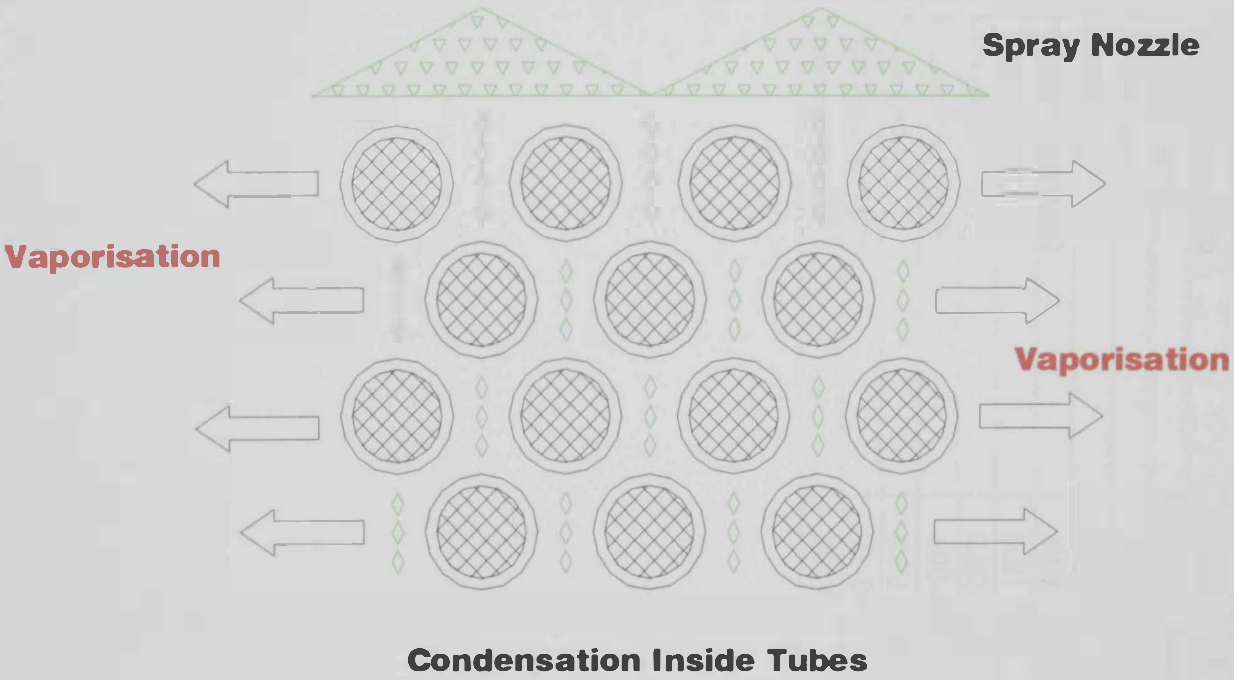


Figure A-4: Horizontal tube falling film process

Table A-1: Umm Al-Nar Seawater Analysis

No.	Parameters	Range	Typical
1	pH @ 25 °C	8.0-8.1	8.1
2	Conductivity @ 25 °C (μS/cm)	66000-70000	69000
3	T.D.S. @ 180 °C	49000-52000	50000
4	Total Hardness (ppm CaCO3)	8900-9500	9000
5	Ca Hardness (ppm)	1400-1500	1500
6	Mg Hardness (ppm)	7400-7900	7700
7	Chloride (ppm)	26000-28000	27500
8	Sulfate (ppm)	3700-3900	3800
9	Calcium (ppm)	580-600	600
10	Magnesium (ppm)	1800-1900	1850
11	Copper (ppm)	0.01-0.02	0.02
12	Iron (ppm)	0.02-0.03	0.02
13	Potassium (ppm)	600-680	650
14	Silica (ppm)	0.02-0.03	0.02
15	Bicarbonate (ppm)	91-94	94
16	Temperature °C	20-35	30

Table 3-2: Main Operating Parameters

Parameter	Unit	Value
"Conditions at rated output and rated fouling factors, when operated at rated top brine temperature and seawater temperature of 33°C and salinity of 52 g/kg"		
Continuous net output of evaporator	m3/day	15911
Minimum controllable output	m3/day	7950
Steam supply to steam transformer from turbine extraction	kg/h	84000
• Pressure of steam supply to steam transformer	ata	2.8
• Temperature of steam supply to steam transformer	°C	130
• Enthalpy	kcal/kg	649.8
Steam supply to ejector from HP/MP reducing station	kg/h	3500
• Pressure of steam supply to ejector	ata	13
• Temperature of steam supply to ejector	°C	210
Specific beat consumption	kJ/kg of distillate	287.5
Performance ratio (product water per 2300 kj net heat input to steam transformer)	-	8
Steam pressure in first effect	ata	0.255
Condensate return flow rate (from steam transformer)	kg/h	84000
Condensate return temperature (from steam transformer)	°C	107.8
Heat transferred in first effect	MW	114,2
Condensate make up steam transformer	kg/h	76500
Maximum brine temperature	°C	63

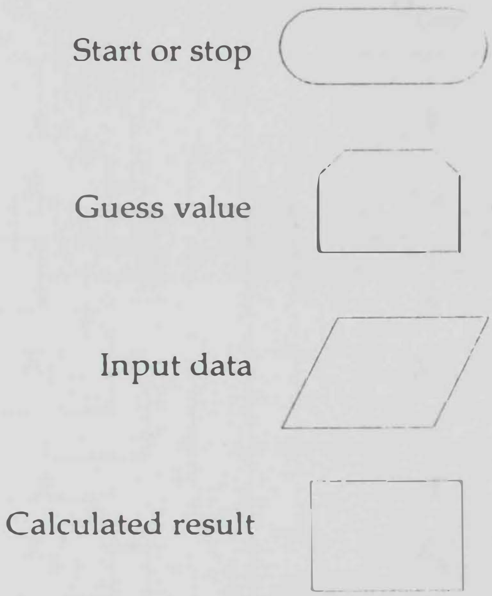
Total requirement of sea water	kg/b	5650000
Minimum brine temperature	°C	44 approx.
Maximum T.D.S. in first effect (brine)	mg/1	73000
Blow down flow rate	kg/h	1641000
Maximum blow down concentration	mg/1	73000
Feed make up flow rate	kg/h	2304000
Cooling water temperature after condenser (reject)	°C	40.5
Seawater flow rate to ejector condensers	kg/h	NA
Scale control chemical		SOKALAN PM 10
Rate of scale control chemical dosing	kg/day	166
Scale control dosing rate related to feed make-up	mg/1	3
Rate of antifoam dosing	kg/day	8.3
Maximum copper content in:		
• Distillate	mg/1	0.04
• Condensate	mg/1	0.04
• Brine blow down	mg/1	0.04

Table A-3: Distiller Materials

Evaporator Vessel	
. Shell in contact with seawater	Stainless steel 316L
. Shell in contact with vapor	Stainless steel 316L
. External reinforcement	Carbon steel
Heat Tubes Bundles	
. Tubes (10 top rows)	Titanium
. Tubes (al l others)	Aluminum brass
. Tube-plates	Stainless steel 316L
Demisters	Stainless steel 316-03
Spray Nozzles	Stainless steel 316L
Distillate Condenser & Seawater Preheaters	
. Shell	Stainless steel 316L
. Tubes	Titanium
. Tube-plates	Stainless steel 316L
. Support plates	Stainless steel 316L
. Water boxes	Stainless steel 316L
Thermo-Compressor	
. Nozzle	Stainless steel 316L
. Diffuser	Stainless steel 316L
Pipe Work Material	
. Seawater	GRP
. Distillate	GRP
. Blow down	GRP
. Ejector Condensate Extraction	Stainless steel 316L

APPENDIX C

MEDNAR Simulator Sub-Flow Charts



Sub-flow charts key

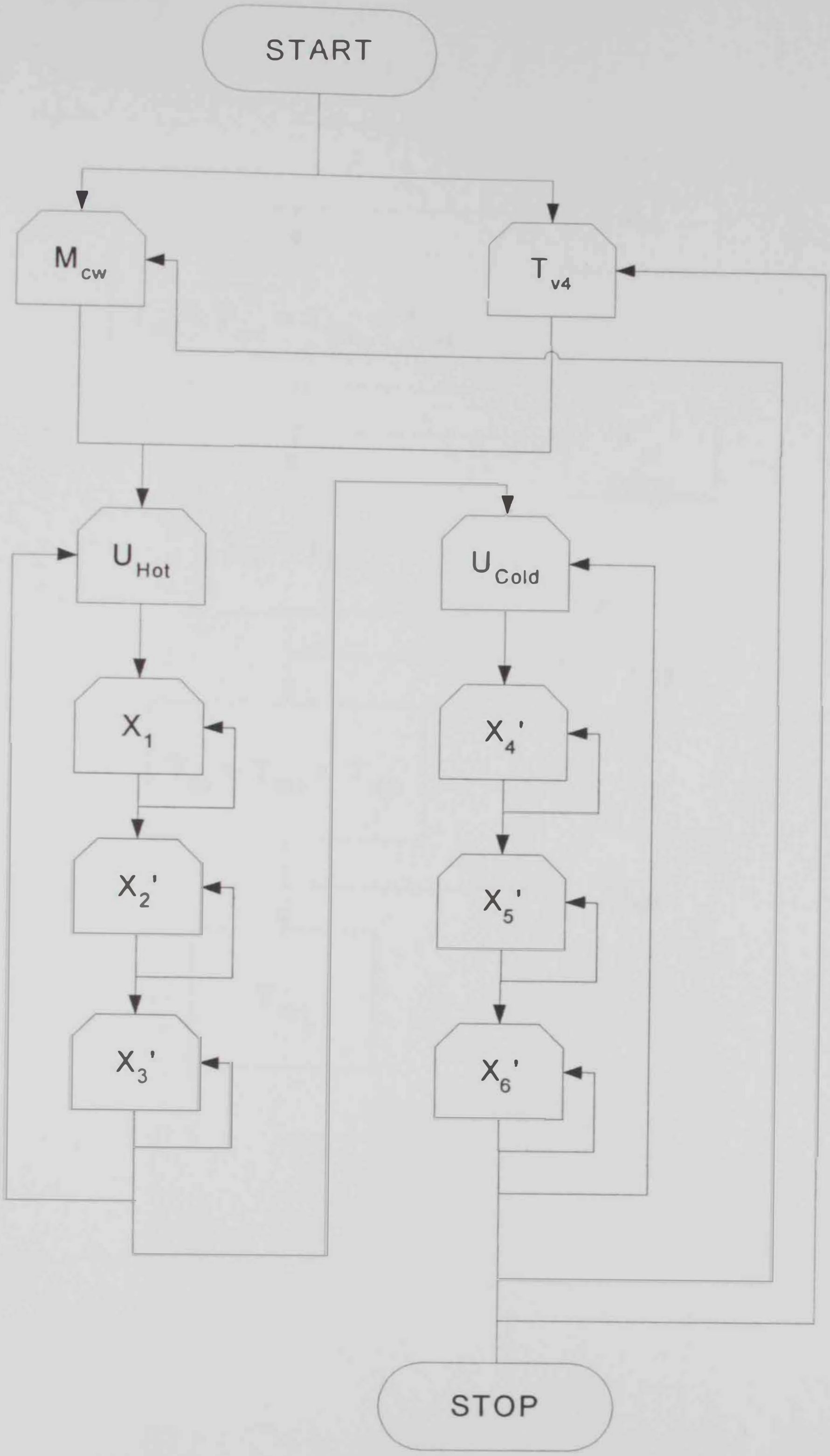


Figure A-4: MEDNAR loops

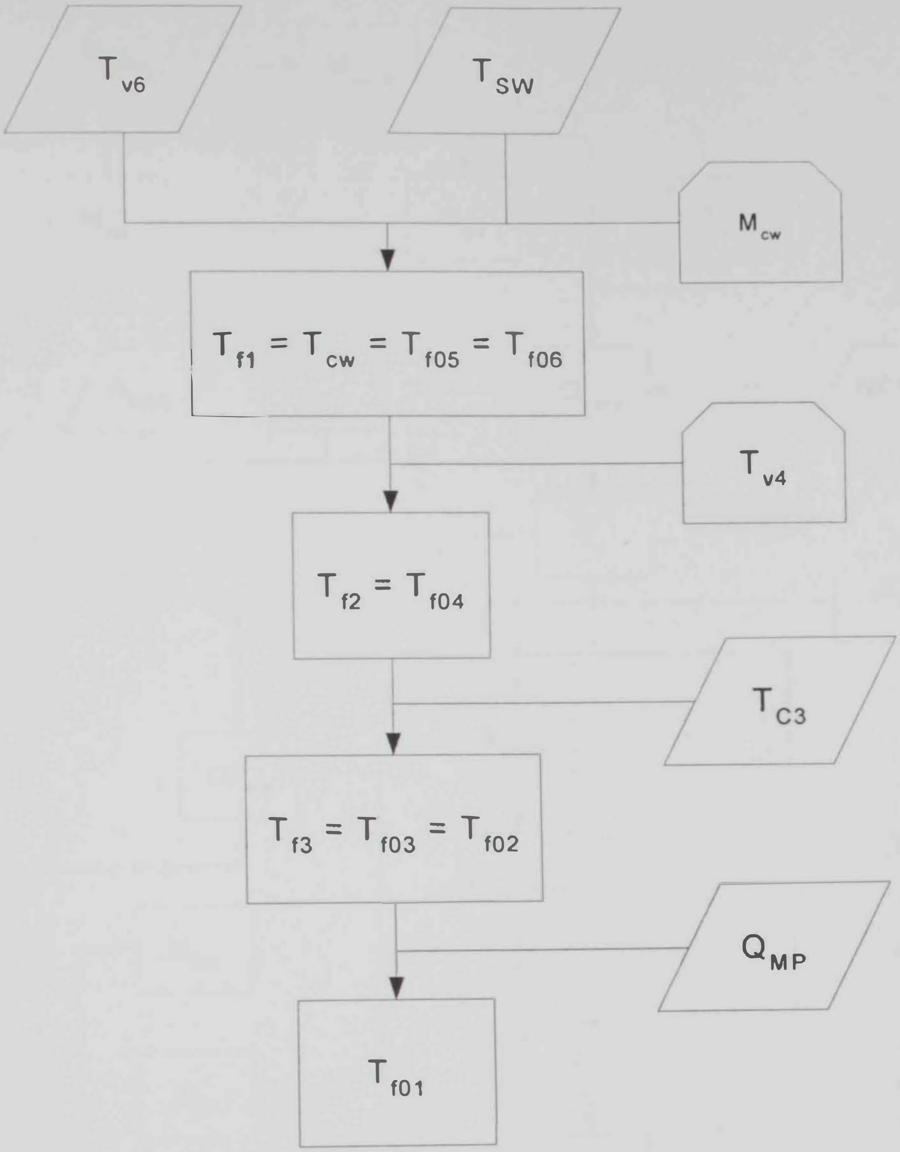


Figure A-5: Feed water temperature profile (Sub-chart A)

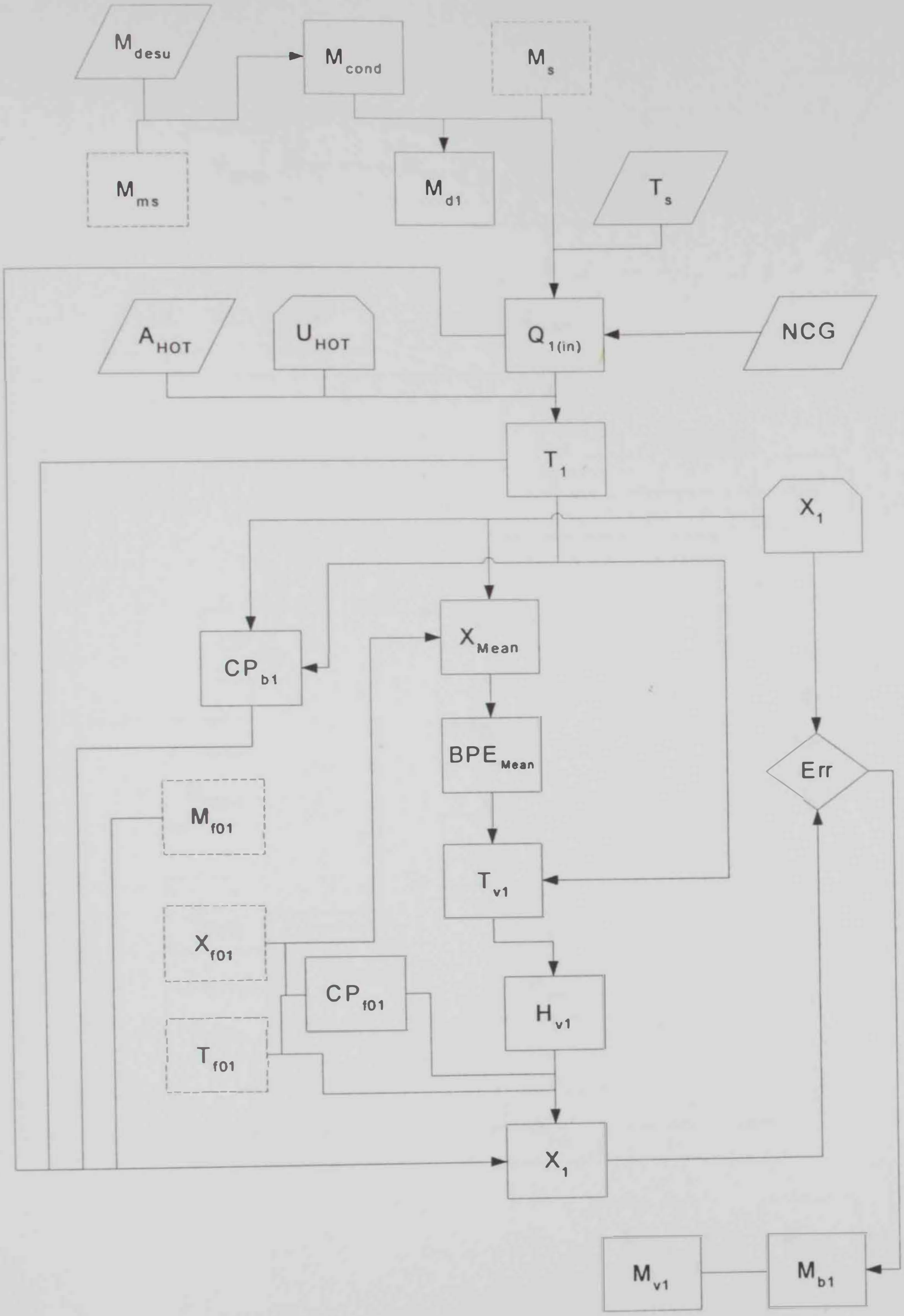


Figure A-6: MEDNAR first effect calculations (Sub-chart B)

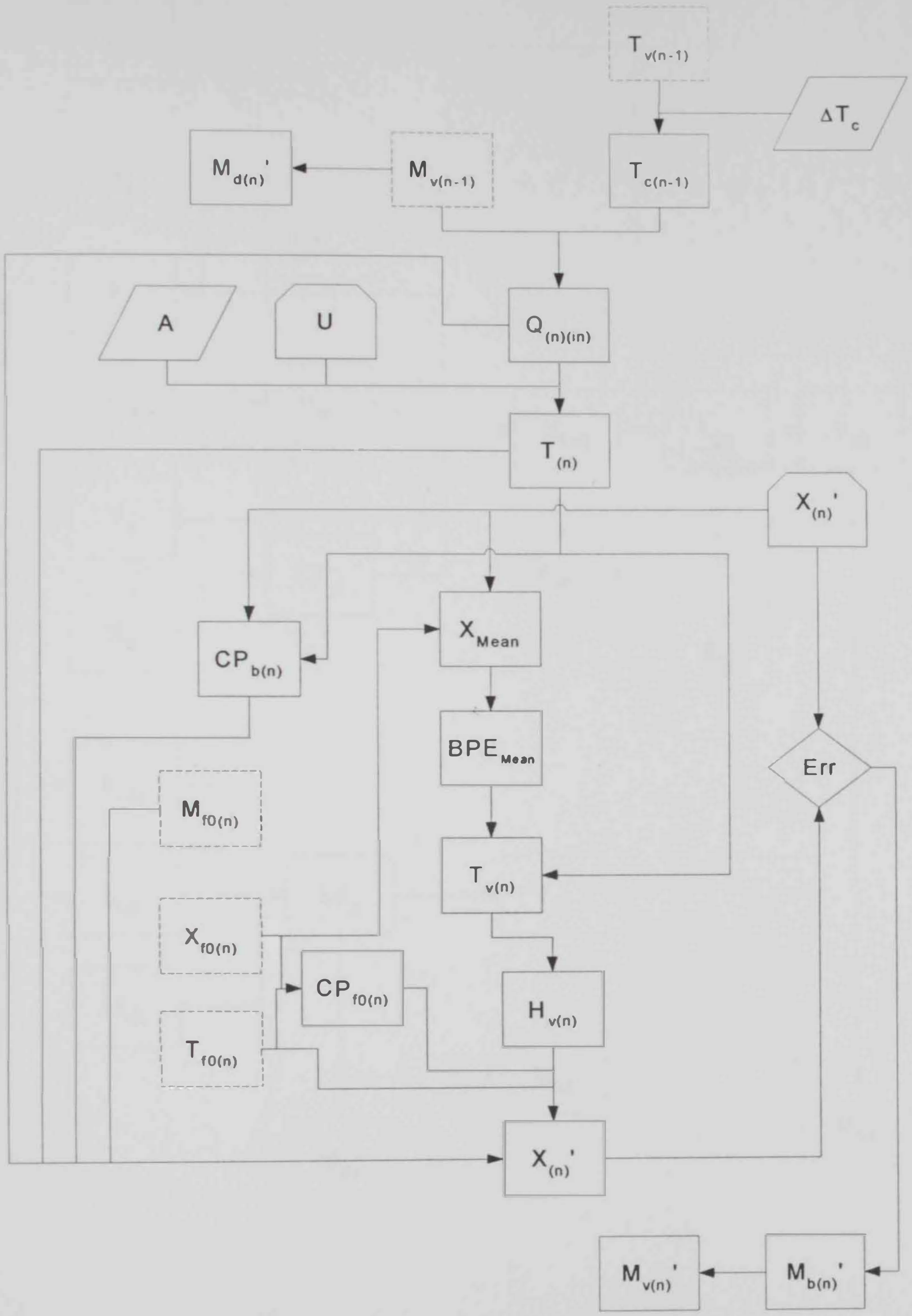


Figure A-7 (Part 1): MEDNAR other effects calculations (Sub-chart C)

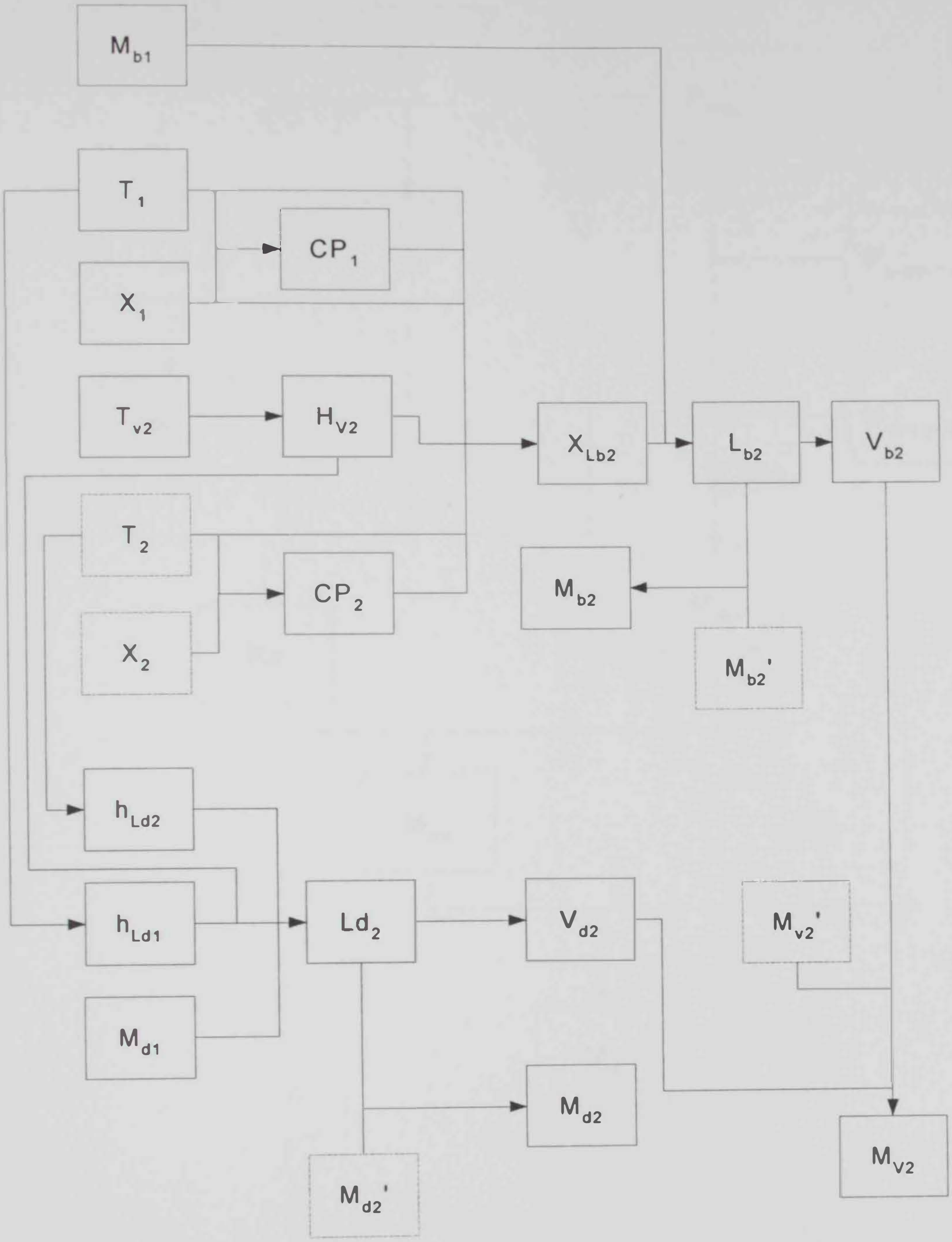


Figure A-7 (Part 2): MEDNAR other effects calculations (Sub-chart C)

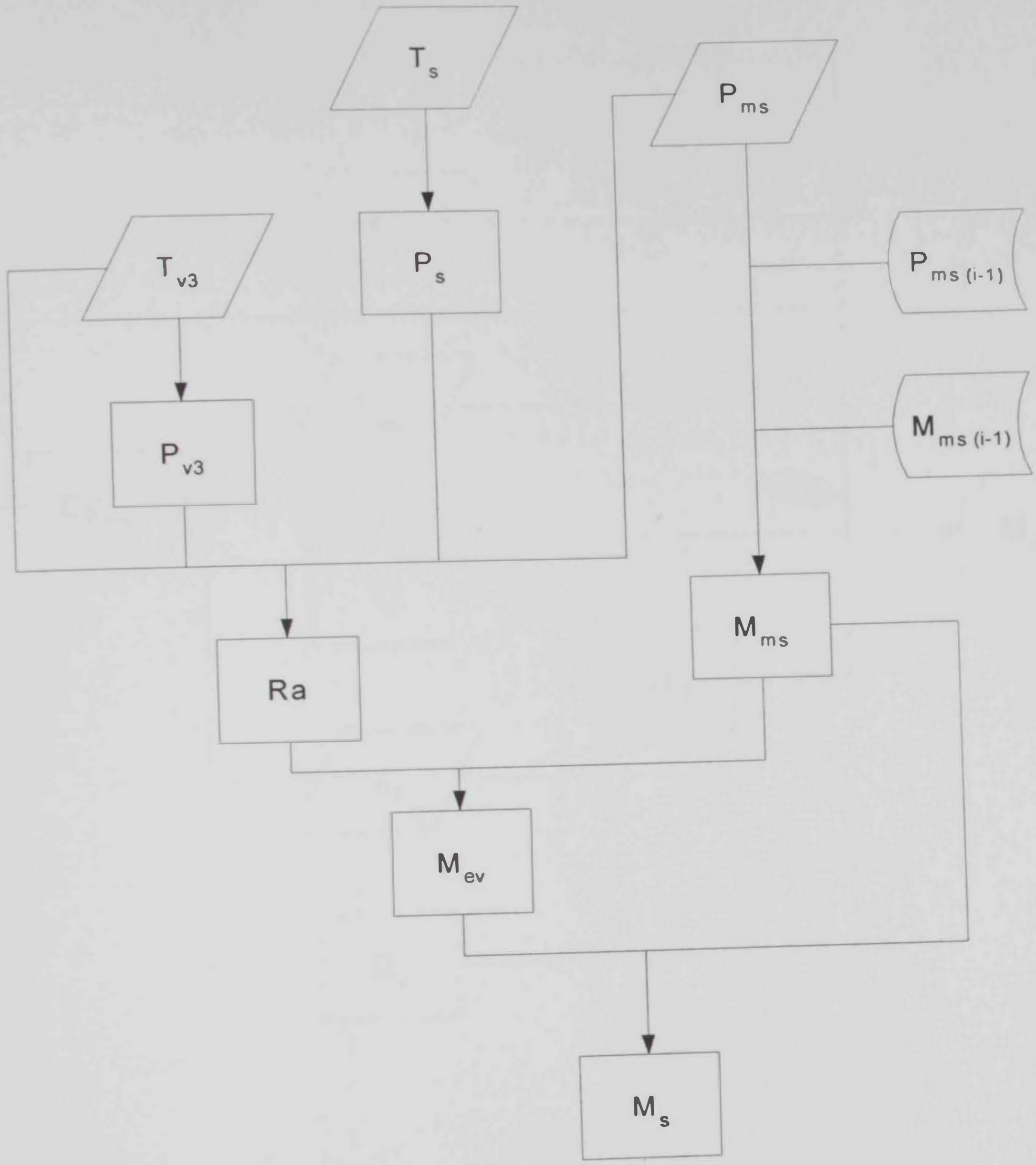


Figure A-8: MEDNAR Thermal vapor compressor calculations

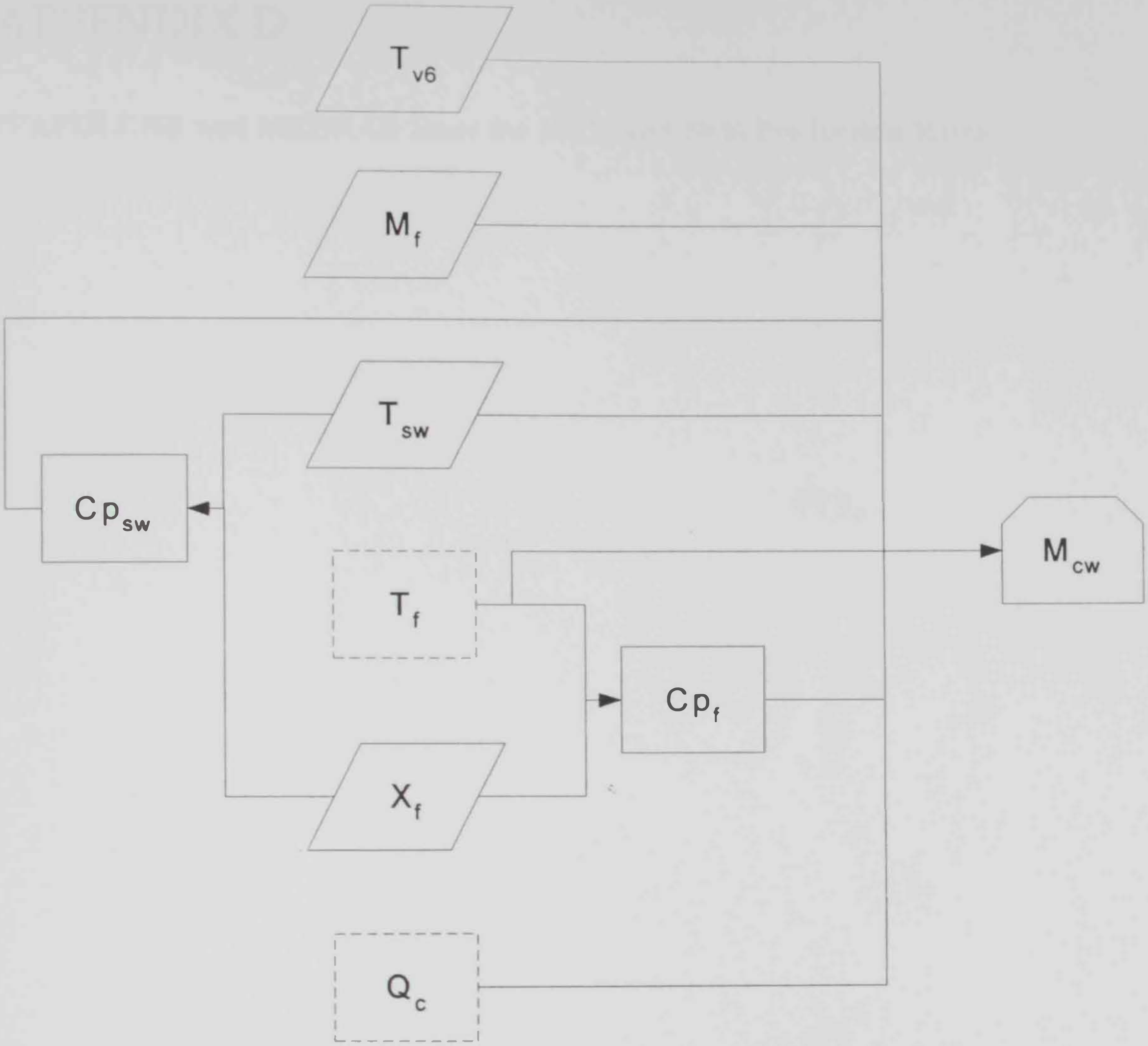


Figure A-9: MEDNAR cooling water flow rate calculation

APPENDIX D

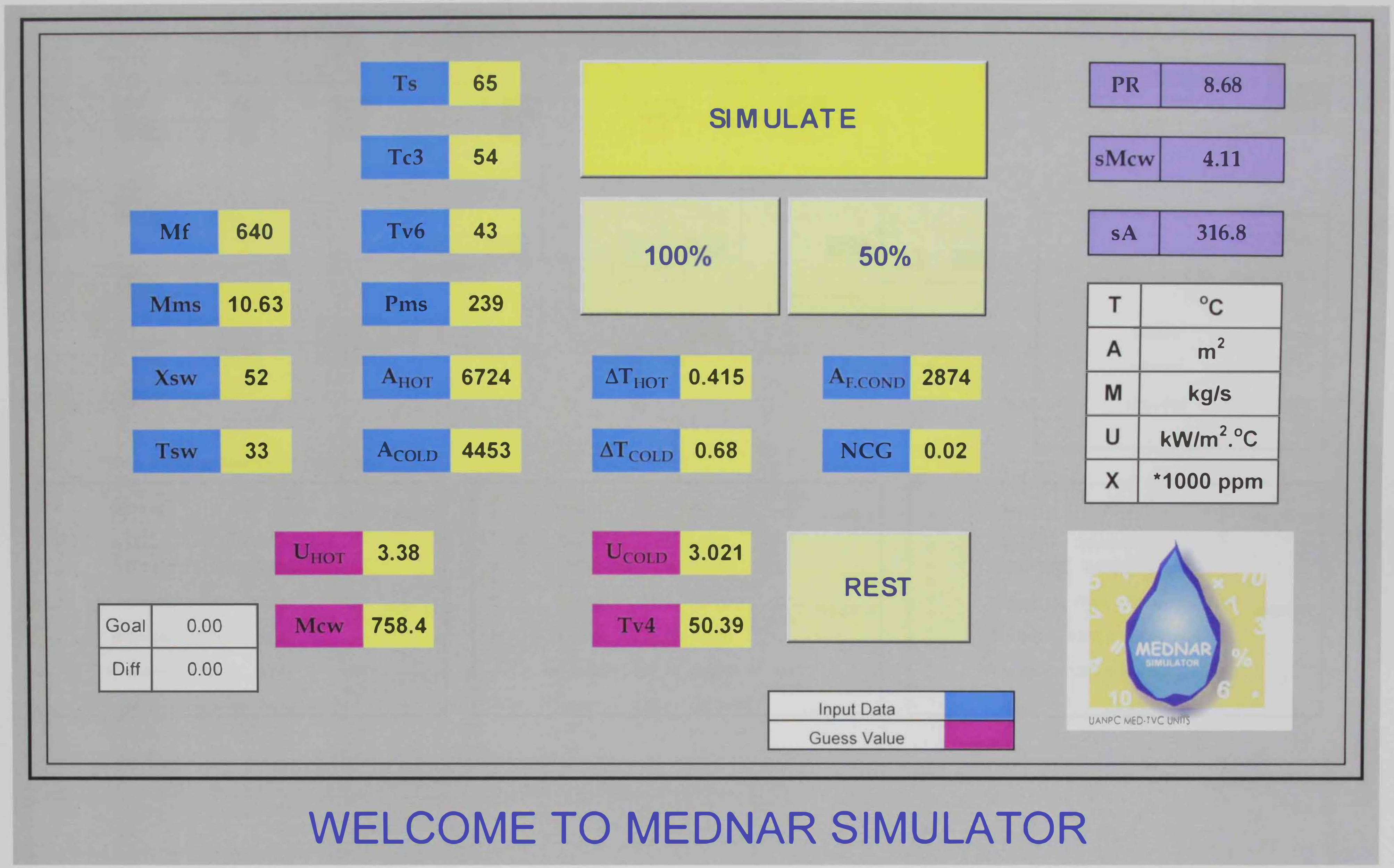
EVAPOLUND and MEDNAR Runs for 100% and 50% Production Rates

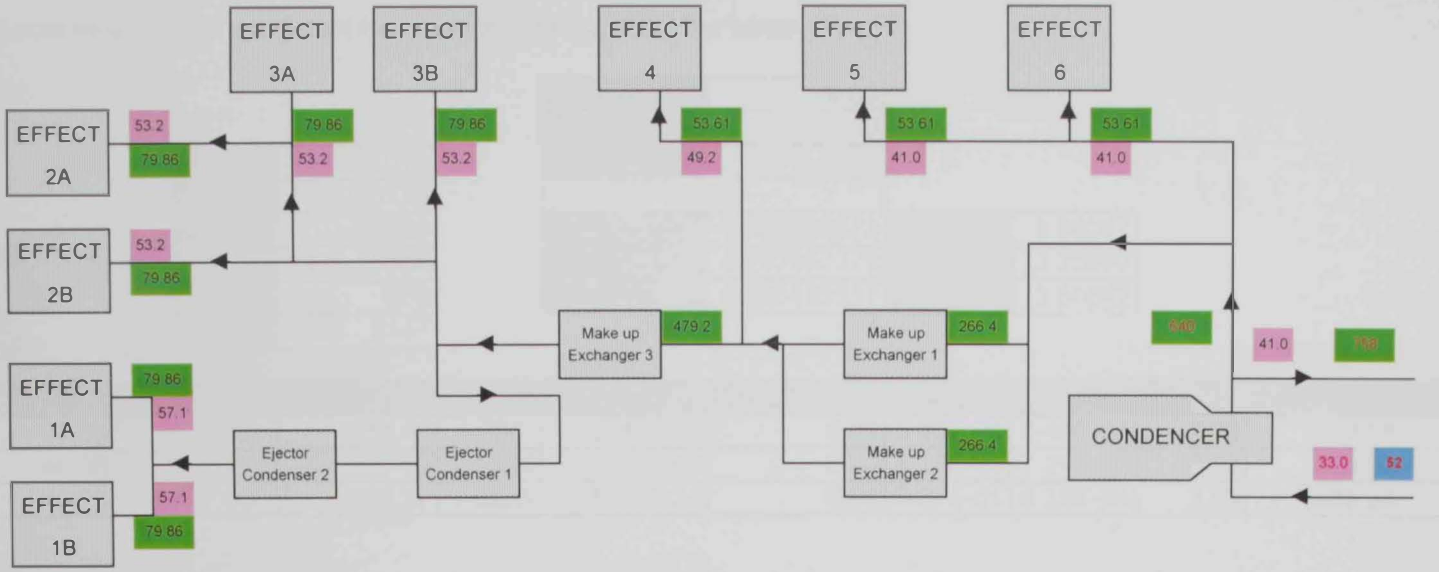
Summarizing Evaporator values

Evaporator No	Vapor out Temp (C)	Vapor out Flow (kg/s)	Area (m2)	U-value kW/(m2,C)	Liquid out Temp (C)	Liquid out Conc (mf)	Liquid out Flow (kg/s)	Liquid out Boil p el (C)
75	43 0	17.28	4,453.0	3.23	43.8	0.073	455.63	0.8
67	47.0	13.96	4,453.0	3.28	47.8	0.073	419.30	0.8
57	50.5	17.30	4,453.0	3.15	51.4	0.073	379.66	0.9
40	54.7	22.68	6,724.0	3.59	55.6	0.073	171.68	0.9
39	54.7	22.68	6,724.0	3.59	55.6	0.073	171.68	0.9
27	58.2	22.33	6,724.0	3.57	59.1	0.073	114.49	0.9
25	58.2	22.33	6,724.0	3.57	59.1	0.073	114.49	0.9
9	61 8	22.91	6,724.0	3.63	62.7	0.073	56.95	0.9
8	61.8	22.91	6,724.0	3.63	62.7	0.073	56.95	0.9
Sum		184 37	53 703 0					

Summarizing Evaporator values

Evaporator No	Vapor out: Temp (C)	Vapor out: Flow (kg/s)	Area (m2)	U-value kW/(m2,C)	Liquid out: Temp (C)	Liquid out: Conc (mf)	Liquid out: Flow (kg/s)	Liquid out: Boil.p. el (C)
75	45.0	12.08	4,453.0	2.52	45.7	0.061	548.76	0.7
67	48.4	9.14	4,453.0	2.60	49.1	0.060	507.23	0.7
57	51.3	11.76	4,453.0	2.52	52.0	0.060	462.77	0.7
40	54.7	9.75	6,724.0	2.31	55.4	0.059	210.46	0.7
39	54.7	9.75	6,724.0	2.31	55.4	0.059	210.46	0.7
27	57.3	9.49	6,724.0	2.29	58.0	0.059	140.34	0.7
25	57.3	9.49	6,724.0	2.29	58.0	0.059	140.34	0.7
9	59.9	9.90	6,724.0	2.33	60.6	0.059	69.96	0.7
8	59.9	9.90	6,724.0	2.33	60.6	0.059	69.96	0.7
Sum		91.24	53,703.0					


Ts 65
Tc3 54
SIMULATE
PR 8.68
sMcw 4.11
sA 316.8
Mf 640
Tv6 43
100%
50%
Mms 10.63
Pms 239
Xsw 52
AHOT 6724
ΔTHOT 0.415
AF.COND 2874
Tsw 33
ACOLD 4453
ΔTCOLD 0.68
NCG 0.02
T °C
A m2
M kg/s
U kW/m2.°C
X *1000 ppm
UHOT 3.38
UCOLD 3.021
REST
Goal 0.00
Diff 0.00
Mcw 758.4
Tv4 50.39
MEDNAR SIMULATOR
UANPC MED-TVC UNITS
Input Data
Guess Value
WELCOME TO MEDNAR SIMULATOR



Temperature *C*	
Pressure *Kpa*	
Flow *Kg/s*	
Salinity **1000 ppm*	

Ejec. Condenser

			2823.237993
	QEC	2475.24	888.0767191
	stream	Flow	Temp
	Mf(inlet)	159.72	53.17
	Mf(outlet)	159.72	57.11

****Mass & Energy balance for Pre-Heaters**

Pre-heater 3

Area m^2	693		Q3	7486.05
U_{HE3}	3.457416621			
3.506418118	3.506609656			
Stream	Flow	Temp	Q3 in	7537.83
Mf2(inlet)	479.17	49.20		
Mf2(outlet)	479.17	53.17	Q3 out	7537.83
Md3"	3.1738	54.68		

Pre-heater 12

Area m^2	1156		Q12	15464.3
U_{HE12}	3.320111218			
3.35697338	3.356356886			
Stream	Flow	Temp	Q12 in	17307.3
Mf1(inlet)	532.78	40.98		
Mf1(outlet)	532.78	49.20	Q12 out	17307.3
Md4"	7.2641	50.39		

F. Condenser

Area m^3	2874		QFC	43750.38
U_{FC}	3.084414825			
3.040172795	3.049596459			
Stream	Flow	Temp	QFC in	44007.51
Mf (inlet)	1398.4	33.00		
Mf (outlet)	1398.4	40.98	QFC out	0
Mdc		43.00		

** Local heat transfer coefficient for seawater flowing inside the tubes

Ri	0.08
Ro	0
K	0.0165

1/UFC	0.328928672	UFC	3.04017
1/UHE12	0.297887379	UHE12	3.35697
1/UHE3	0.285191317	UHE3	3.50642

dout	0.019
dinner	0.0176
salinity	52000
n12	1070.66667
n3	963
nfc	3147

temperature oC	hout	Ditus-Bolter	velocity	Reynolds	Prandtl	Kwater	muwater	cpwater	rhowater
33.00	6.02	654.40	1.768	38106.09	0.01	6.18E-01	8.44E-04	3.92	1033.05
40.98	7.24	777.53	1.986	49689.46	0.00	6.28E-01	7.24E-04	3.93	1029.92
49.20	7.90	839.57	1.9928	57337.34	0.00	6.37E-01	6.28E-04	3.93	1026.34

** Local heat transfer coefficient on the outside surface of tubes

dout	0.019	xncondout	0

con1	1.525535	con2	1
num1	0.816033	dinner	0.0176
mumt	2.093425		

temperature oC	hout	delt	rhovapor	rhocond	kliquid	lamdav	muliquid	mass	velocityout
43.00	52.23	0.1	0.059	990.669	6.33E-04	2399.32	6.20E-04	18.24	36.14153313
50.39	54.27	0.1	0.085	987.507	6.41E-04	2381.58	5.45E-04	7.26	27.79203266
54.68	55.39	0.1	0.103	985.527	6.45E-04	2371.23	5.08E-04	3.17	11.76368426

** STREAMS DATA & RESULTS

		STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B.	E.B
Cell #1		Cell #1		61.78				258.45	2,613.02		
	In	Ms	24.01	65.00					2,618.49	103.87	80827.03
		Mf01	79.86	57.11	52.00		3.93				
	Out	Mb1	57.12	62.57	72.70	0.94	3.85			103.87	79700.04
		Mv1	22.74	61.78					2,613.02		
		Md1	13.39	65.00				271.95			
		Mcon	10.63	65.00				271.95			
										Err %	1.39

U	3.379757453
A	6724
ΔT loss Hot	0.415

Q	T1	delta T	Qcal	Err
55222.455	62.57	2.43	55222.45481	0

NCG	0.02

** BOILING POINT ELEVATION

STREAM #	X	T (K)	BPE	Tv1 ($^{\circ}$C)	Hv1
Mb1(assume)1	0.072702159	335.72	0.9395019		
LOOP 1	0.051	335.72	0.6260898	61.94	2613.294517
LOOP 2	0.062253304	335.72	0.7841983	61.79	2613.024731
LOOP 3	0.062350255	335.72	0.7856006	61.78	2613.022338
LOOP 4	0.062351073	335.72	0.7856124	61.78	2613.022318
LOOP 5	0.06235108	335.72	0.7856125	61.78	2613.022317

** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T	A	B	C	D	Cp
Mf01	52.00	57.11	3895.8024	1.076734	-0.010716138	6.76017E-05	3.934935957
LOOP 1	50	62.57	3906.535	1.014725	-0.0100305	6.54708E-05	3.94679474
LOOP 2	72.50660776	62.57	3791.4286	1.607678	-0.016873612	8.74077E-05	3.847372275
LOOP 3	72.7005107	62.57	3790.491	1.611786	-0.016924246	8.75772E-05	3.846535037
LOOP 4	72.7021453	62.57	3790.4831	1.611821	-0.016924672	8.75786E-05	3.84652798
LOOP 5	72.70215908	62.57	3790.483	1.611821	-0.016924676	8.75786E-05	3.846527921

Cp= (A+BT+CT^2+DT^3)E-3

WHERE

T= TEMPERATURE C

S= WATER SALINITY G/Kg

A= 4206.8-6.6197S+1.2288E-2S^2

B= -1.1262+5.4178E-2S-2.2719E-4S^2'

C= 1.2026E-2-5.3566E-4S+1.8906E-6S^2

D= 6.87774E-7+1.517E-6S-4.4268E-9S^2

**SALINITY

STREAM #	Assumed X	Mb1	Mv1	Cp	Calc X
LOOP 1	50	83.05556	-3.194444	3.946795	72.50660776
LOOP 2	72.50660776	57.27447	22.586638	3.847372	72.7005107
LOOP 3	72.7005107	57.12171	22.739397	3.846535	72.7021453
LOOP 4	72.7021453	57.12043	22.740681	3.846528	72.70215908
LOOP 5	72.70215908	57.12042	22.740692	3.846528	72.7021592

** STREAMS DATA & RESULTS

		STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B.	E.B.
Cell #2		Cell #2		58.20				243.43	2,606.89		
	In	Mf02	79.86	53.17	52.00		3.93			173.11	93,510.85
		Md1	13.39	65.00				271.95			
		Mb1	57.12	62.57	72.70	0.94	3.85				
		Mv1	22.74	61.78				258.45	2,613.02		
	Internal	Md2*	22.74	61.37				256.71	2,612.31		
		Mb2*	57.95	59.01	71.66	0.81	3.85				
		Mv2*	21.91	58.20					2,606.89		
		Ld2	13.23	58.20				243.43			
		Lb2	56.80	59.01	73.12	0.92	3.84				
		Vd2'	0.16	58.20					2,606.89		
		Vb2'	0.32	58.20					2,606.89		
	Out	Md2	35.97	60.20				251.83		173.11	93,489.74
		Mb2	114.75	59.01	72.38	0.91	3.85				
		Mv2	22.40	58.20				243.43	2,606.89		
										Err %	0.02

U	3.379757453
A	6724
dlta loss	0.415

Q	T2	delta T	Qcal	Err
53568.012	59.01	2.36	53568.01214	0

** BOILING POINT ELEVATION

STREAM #	X	T2 (K)	BPE	Tv2 (°C)	Hv2
Mb2	0.072382699	332.16	0.9121081		
Lb2	0.073117883	332.16	0.9230707		
Mb2' (LOOP 1)	0.054506116	332.16	0.6581367	58.35	2607.145091
Mb2' (LOOP 2)	0.065250614	332.16	0.8079357	58.20	2606.887394
Mb2' (LOOP 3)	0.065336515	332.16	0.8091673	58.20	2606.885275
Mb2' (LOOP 4)	0.065337187	332.16	0.8091769	58.20	2606.885259
Mb2' (LOOP 5)	0.065337192	332.16	0.809177	58.20	2606.885258

** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T2	A	B	C	D	Cp
Mf02	52.00	53.17	3895.8024	1.076734	-0.010716138	6.76017E-05	3.932919
Mb2	72.38	59.01	3792.0282	1.605044	-0.016841181	8.72992E-05	[illegible]
Lb2	73.12	59.01	3788.476	1.620572	-0.017032752	8.79409E-05	[illegible]
Mb2' (LOOP 1)	50	59.01	3906.535	1.014725	-0.0100305	6.54708E-05	3.94494
Mb2' (LOOP 2)	71.488996	59.01	3796.3643	1.585836	-0.016605551	8.65126E-05	3.849899
Mb2' (LOOP 3)	71.66079828	59.01	3795.5292	1.589557	-0.016651082	8.66644E-05	3.849156
Mb2' (LOOP 4)	71.66214257	59.01	3795.5227	1.589586	-0.016651438	8.66656E-05	3.84915
Mb2' (LOOP 5)	71.66215309	59.01	3795.5226	1.589586	-0.01665144	8.66656E-05	3.84915

Cp= (A+BT+CT^2+DT^3)E-3

WHERE

T= TEMPERATURE C

S= WATER SALINITY G/Kg

A= 4206.8-6.6197S+1.2288E-2S^2

B= -1.1262+5.4178E-2S-2.2719E-4S^2'

C= 1.2026E-2-5.3566E-4S+1.8906E-6S^2

D= 6.87774E-7+1.517E-6S-4.4268E-9S^2

**MODUL

STREAM #	Assumed X2'	Mb2'	Mv2'	Cp	Calc X2'
Mb2' (LOOP 1)	50	83.05556	-3.19	3.94494	71.488996
Mb2' (LOOP 2)	71.488996	58.08975	21.77	3.849899	71.66079828
Mb2' (LOOP 3)	71.66079828	57.95048	21.91	3.849156	71.66214257
Mb2' (LOOP 4)	71.66214257	57.94939	21.91	3.84915	71.66215309
Mb2' (LOOP 5)	71.66215309	57.94939	21.91	3.84915	71.66215317

** STREAMS DATA & RESULTS

		STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B.	E.B.
Cell #3		Cell #3		54.00				225.81	2,599.62		
	In	Mf03	79.86	53.17	52.00		3.93			252.97	110,189.99
		Md2	35.97	60.20				251.83			
		Mb2	114.75	59.01	72.38	0.91	3.85				
		Mv2	22.40	58.20				243.43	2,606.89		
	Internal	Md3'	22.40	57.79				241.69	2,606.17		
		Mb3'	57.81	55.46	71.83	0.77	3.85				
		Mv3'	22.05	54.69					2,600.82		
		Ld3	35.57	54.00				225.81			
		Lb3	114.09	55.46	72.80	0.90	3.84				
		Vd3	0.39	54.00					2,599.62		
		Vb3	0.66	54.69					2,600.82		
	Out	Md3*	57.97	55.42				231.78		252.97	110,153.13
		Mb3	171.90	55.46	72.47	0.89	3.84				
		Mv3*	9.71	54.68					2,600.80		
		Mev (tvc)in	13.39	54.68					2,600.80		
										Err %	0.03

U	3.379757453
A	6724
dlta loss	0.415

Q	T3	delta T	Qcal	Err
52959.69	55.46	2.33	52959.68992	0

Md3"
3.173816141

** BOILING POINT ELEVATION

STREAM #	X	T3 (K)	BPE	Tv3 (°C)	Hv3
Mb3	0.072473382	328.61	0.8911467		
Lb3	0.07279744	328.61	0.8958767		
Mb3' (LOOP 1)	0.052728823	328.61	0.6186789	54.84	2601.075056
Mb3' (LOOP 2)	0.063562425	328.61	0.7647079	54.69	2600.821848
Mb3' (LOOP 3)	0.063643873	328.61	0.765837	54.69	2600.81989
Mb3' (LOOP 4)	0.063644471	328.61	0.7658453	54.69	2600.819876
Mb3' (LOOP 5)	0.063644476	328.61	0.7658454	54.69	2600.819876

** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T3	A	B	C	D	Cp
Mf03	52.00	53.17	3895.8024	1.076734	-0.010716138	6.76017E-05	3.932919
Mb3	72.47	55.46	3791.5893	1.606972	-0.016864921	8.73786E-05	3.843743
Lb3	72.80	55.46	3790.0163	1.613861	-0.016949843	8.76629E-05	3.842339
Mb3' (LOOP 1)	50	55.46	3906.535	1.014725	-0.0100305	6.54708E-05	3.943127
Mb3' (LOOP 2)	71.66720498	55.46	3795.4981	1.589695	-0.016652777	8.667E-05	3.847225
Mb3' (LOOP 3)	71.83010015	55.46	3794.707	1.59321	-0.016695841	8.68137E-05	3.846521
Mb3' (LOOP 4)	71.83129744	55.46	3794.7012	1.593236	-0.016696157	8.68147E-05	3.846516
Mb3' (LOOP 5)	71.83130624	55.46	3794.7011	1.593236	-0.01669616	8.68147E-05	3.846516

Cp= (A+BT+CT^2+DT^3)E-3

WHERE

T= TEMPERATURE C

S= WATER SALINITY G/Kg

A= 4206.8-6.6197S+1.2288E-2S^2

B= -1.1262+5.4178E-2S-2.2719E-4S^2'

C= 1.2026E-2-5.3566E-4S+1.8906E-6S^2

D= 6.87774E-7+1.517E-6S-4.4268E-9S^2

**MODUL

STREAM #	Assumed X3'	Mb3'	Mv3'	Cp	Calc X3'
Mb3' (LOOP 1)	50	83.05556	-3.19	3.943127	71.66720498
Mb3' (LOOP 2)	71.66720498	57.9453	21.92	3.847225	71.83010015
Mb3' (LOOP 3)	71.83010015	57.81389	22.05	3.846521	71.83129744
Mb3' (LOOP 4)	71.83129744	57.81293	22.05	3.846516	71.83130624
Mb3' (LOOP 5)	71.83130624	57.81292	22.05	3.846516	71.8313063

Steam Jet Ejector Formula: A/B

Ra	Mms	Mev	Ps	Pv	Pms	Tv	PCF	TCF	Ms
0.793572	10.625	13.38882	25.03819	15.38045	239	54.68	1.412136	0.971952	24.01382

** STREAMS DATA & RESULTS

		STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B	E.B
Cell #4		Cell #4		50.40				210.75	2,593.35		
	In	Mf04	53.61	49.20	52.00		3.93			532.78	153,511.72
	In	Md3	119.12	55.42				231.78			
	In	Mb3	343.80	55.46	72.47	0.89	3.84				
	In	Mv3	16.24	54.68				228.66	2,600.80		
	Internal	Md4'	16.24	54.00				225.81	2,599.62		
	Internal	Mb4'	37.58	51.13	74.19	0.73	3.83				
	Internal	Mv4'	16.03	50.40					2,593.35		
	Internal	Ld4	118.07	50.40				210.75			
	Internal	Lb4	341.35	51.13	72.99	0.87	3.84				
	Internal	Vd4	1.05	50.40					2,593.35		
	Internal	Vb4	2.46	50.40					2,593.35		
	Out	Md4*	134.31	50.93				212.94		532.78	153,651.21
	Out	Mb4	378.93	51.13	73.11	0.87	3.84				
	Out	Mv4*	19.54	50.40					2,593.35		
										Err %	-0.09

U	3.021275618
A	4453
ΔT loss Cold	0.68

Q	T4	delta T	Qcal	Err
38560.917	51.13	2.87	38560.91684	0

Md4"
7.264069521

** BOILING POINT ELEVATION

STREAM #	X	T4 (K)	BPE	Tv4 (°C)	Hv3
Mb4	0.073112953	324.28	0.8733238		
Lb4	0.072994665	324.28	0.8716527		
Mb4' (LOOP 1)	0.050566908	324.28	0.5727598	50.56	2593.622969
Mb4' (LOOP 2)	0.062574193	324.28	0.7284482	50.41	2593.350418
Mb4' (LOOP 3)	0.062660041	324.28	0.7295963	50.40	2593.348407
Mb4' (LOOP 4)	0.062660639	324.28	0.7296043	50.40	2593.348393
Mb4' (LOOP 5)	0.062660643	324.28	0.7296043	50.40	2593.348393

1.344207

** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T4	A	B	C	D	Cp
Mf04	52.00	49.20	3895.8024	1.076734	-0.010716138	6.76017E-05	3.930887
Mb4	73.11	51.13	3788.4997	1.620469	-0.017031475	8.79366E-05	3.838586
Lb4	72.99	51.13	3789.0704	1.617986	-0.017000787	8.78337E-05	3.839096
Mb4' (LOOP 1)	50	51.13	3906.535	1.014725	-0.0100305	6.54708E-05	3.940949
Mb4' (LOOP 2)	74.01457154	51.13	3784.1613	1.639179	-0.017263642	8.87172E-05	3.834701
Mb4' (LOOP 3)	74.18626749	51.13	3783.3374	1.6427	-0.017307506	8.8865E-05	3.833963
Mb4' (LOOP 4)	74.18746307	51.13	3783.3317	1.642725	-0.017307811	8.8866E-05	3.833958
Mb4' (LOOP 5)	74.18747139	51.13	3783.3317	1.642725	-0.017307813	8.8866E-05	3.833957

Cp= (A+BT+CT^2+DT^3)E-3

WHERE
T= TEMPERATURE C
S= WATER SALINITY G/Kg
A= 4206.8-6.6197S+1.2288E-2S^2
B= -1.1262+5.4178E-2S-2.2719E-4S^2'
C= 1.2026E-2-5.3566E-4S+1.8906E-6S^2
D= 6.87774E-7+1.517E-6S-4.4268E-9S^2

**MODUL

STREAM #	Assumed X4'	Mb4'	Mv4'	Cp	Calc X4'
Mb4' (LOOP 1)	50	55.75556	-2.14	3.940949	74.01457154
Mb4' (LOOP 2)	74.01457154	37.66526	15.95	3.834701	74.18626749
Mb4' (LOOP 3)	74.18626749	37.57808	16.03	3.833963	74.18746307
Mb4' (LOOP 4)	74.18746307	37.57748	16.03	3.833958	74.18747139
Mb4' (LOOP 5)	74.18747139	37.57747	16.03	3.833957	74.18747145

** STREAMS DATA & RESULTS

		STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B	E.B
Cell #5		Cell #5		46.91				196.13	2,587.21		
Cell #5	In	Mf05	53.61	40.98	52.00		3.93			586.39	144,985.82
Cell #5	In	Md4	141.58	50.93				212.94			
Cell #5	In	Mb4	378.93	51.13	73.11	0.87	3.84				
Cell #5	In	Mv4	12.28	50.40				210.75	2,593.35		
Cell #5	Internal	Md5'	12.28	49.72				207.90	2,592.16		
Cell #5	Internal	Mb5'	41.94	47.55	66.47	0.64	3.87				
Cell #5	Internal	Mv5'	11.67	46.91					2,587.21		
Cell #5	Internal	Ld5	140.58	46.91				196.13			
Cell #5	Internal	Lb5	376.96	47.55	73.49	0.86	3.83				
Cell #5	Internal	Vd5	1.00	46.91					2,587.21		
Cell #5	Internal	Vb5	1.97	46.91					2,587.21		
Cell #5	Out	Md5	152.86	47.29				197.71		586.39	144,535.86
Cell #5	Out	Mb5	418.90	47.55	72.79	0.85	3.84				
Cell #5	Out	Mv5	14.64	46.91					2,587.21		
										Err %	0.31

U	3.021275618
A	4453
dlta loss	0.68

Q	T5	delta T	Qcal	Err
29268.944	47.55	2.18	29268.94381	0

** BOILING POINT ELEVATION

STREAM #	X	T5 (K)	BPE	Tv5 (°C)	Hv3
Mb5	0.07279194	320.70	0.8466569		
Lb5	0.073494789	320.70	0.8563357		
Mb5' (LOOP 1)	0.048774343	320.70	0.5363728	47.01	2587.387055
Mb5' (LOOP 2)	0.056962342	320.70	0.6378716	46.91	2587.207979
Mb5' (LOOP 3)	0.057011232	320.70	0.6384904	46.91	2587.206887
Mb5' (LOOP 4)	0.057011519	320.70	0.638494	46.91	2587.20688
Mb5' (LOOP 5)	0.05701152	320.70	0.6384941	46.91	2587.20688

** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T5	A	B	C	D	Cp
Mf05	52.00	40.98	3895.8024	1.076734	-0.010716138	6.76017E-05	3.926581
Mb5	72.79	47.55	3790.0492	1.613718	-0.016948072	8.7657E-05	3.837885
Lb5	73.49	47.55	3786.66	1.628438	-0.017130173	8.82681E-05	3.83485
Mb5' (LOOP 1)	50	47.55	3906.535	1.014725	-0.0100305	6.54708E-05	3.939144
Mb5' (LOOP 2)	66.37599805	47.55	3821.5489	1.468971	-0.015199412	8.18767E-05	3.865834
Mb5' (LOOP 3)	66.47377833	47.55	3821.0613	1.471318	-0.01522723	8.19675E-05	3.865405
Mb5' (LOOP 4)	66.47435121	47.55	3821.0584	1.471331	-0.015227393	8.1968E-05	3.865403
Mb5' (LOOP 5)	66.47435457	47.55	3821.0584	1.471331	-0.015227394	8.1968E-05	3.865403

Cp= (A+BT+CT^2+DT^3)E-3

WHERE

T= TEMPERATURE C

S= WATER SALINITY G/Kg

A= 4206.8-6.6197S+1.2288E-2S^2

B= -1.1262+5.4178E-2S-2.2719E-4S^2'

C= 1.2026E-2-5.3566E-4S+1.8906E-6S^2

D= 6.87774E-7+1.517E-6S-4.4268E-9S^2

**MODUL

STREAM #	Assumed X5'	Mb5'	Mv5'	Cp	Calc X5'
Mb5' (LOOP 1)	50	55.75556	-2.14	3.939144	66.37599805
Mb5' (LOOP 2)	66.37599805	41.99979	11.61	3.865834	66.47377833
Mb5' (LOOP 3)	66.47377833	41.93801	11.67	3.865405	66.47435121
Mb5' (LOOP 4)	66.47435121	41.93765	11.67	3.865403	66.47435457
Mb5' (LOOP 5)	66.47435457	41.93764	11.67	3.865403	66.47435459

** STREAMS DATA & RESULTS

		STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B.	E.B.
Cell #6		Cell #6		43.00				179.78	2,580.28		
	In	Mf06	53.61	40.98	52.00		3.93			640.00	153,161.89
		Md5	152.86	47.29				197.71			
		Mb5	418.90	47.55	72.79	0.85	3.84				
		Mv5	14.64	46.91				196.13	2,587.21		
	Internal	Md6'	14.64	46.23				193.28	2,586.01		
		Mb6'	39.24	43.63	71.05	0.62	3.84				
		Mv6'	14.37	43.00					2,580.28		
		Ld6	151.71	43.00				179.78			
		Lb6	416.32	43.63	73.24	0.83	3.83				
		Vd6	1.14	43.00					2,580.28		
		Vb6	2.57	43.00					2,580.28		
	Out	Md6	166.35	43.50				181.86		640.00	153,137.42
		Mb6	455.56	43.63	73.05	0.83	3.83				
		Mv6	18.09	43.00					2,580.28		
										Err %	0.02

U	3.021275618
A	4453
dlta loss	0.68

Q	T6	delta T	Qcal	Err
35024.811	43.63	2.60	35024.81051	0

** BOILING POINT ELEVATION

STREAM #	X	T6 (K)	BPE	Tv6 (°C)	Hv3
Mb6	0.073053274	316.78	0.8261825		
Lb6	0.073242051	316.78	0.8287076		
Mb6' (LOOP 1)	0.046813421	316.78	0.4982749	43.13	2580.506309
Mb6' (LOOP 2)	0.057276881	316.78	0.6237633	43.00	2580.283005
Mb6' (LOOP 3)	0.057338191	316.78	0.6245183	43.00	2580.281662
Mb6' (LOOP 4)	0.057338541	316.78	0.6245226	43.00	2580.281654
Mb6' (LOOP 5)	0.057338543	316.78	0.6245226	43.00	2580.281654

** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T6	A	B	C	D	Cp
Mf06	52.00	40.98	3895.8024	1.076734	-0.010716138	6.76017E-05	3.926581
Mb6	73.05	43.63	3788.7876	1.619217	-0.017015999	8.78847E-05	3.83434
Lb6	73.24	43.63	3787.8773	1.62317	-0.017064906	8.80488E-05	3.833523
Mb6' (LOOP 1)	50	43.63	3906.535	1.014725	-0.0100305	6.54708E-05	3.93715
Mb6' (LOOP 2)	70.92692135	43.63	3799.1014	1.57357	-0.016455809	8.60143E-05	3.843573
Mb6' (LOOP 3)	71.04954027	43.63	3798.5036	1.576258	-0.016488578	8.61233E-05	3.843039
Mb6' (LOOP 4)	71.05024062	43.63	3798.5002	1.576274	-0.016488765	8.61239E-05	3.843036
Mb6' (LOOP 5)	71.05024461	43.63	3798.5002	1.576274	-0.016488766	8.61239E-05	3.843036

Cp= (A+BT+CT^2+DT^3)E-3

WHERE

T=	TEMPERATURE C
S=	WATER SALINITY G/Kg
A=	4206.8-6.6197S+1.2288E-2S^2
B=	-1.1262+5.4178E-2S-2.2719E-4S^2'
C=	1.2026E-2-5.3566E-4S+1.8906E-6S^2
D=	6.87774E-7+1.517E-6S-4.4268E-9S^2

**MODUL

STREAM #	Assumed X6'	Mb6'	Mv6'	Cp	Calc X6'
Mb6' (LOOP 1)	50	55.75556	-2.14	3.93715	70.92692135
Mb6' (LOOP 2)	70.92692135	39.30493	14.31	3.843573	71.04954027
Mb6' (LOOP 3)	71.04954027	39.2371	14.37	3.843039	71.05024062
Mb6' (LOOP 4)	71.05024062	39.23671	14.37	3.843036	71.05024461
Mb6' (LOOP 5)	71.05024461	39.23671	14.37	3.843036	71.05024464

** STREAMS DATA & RESULTS

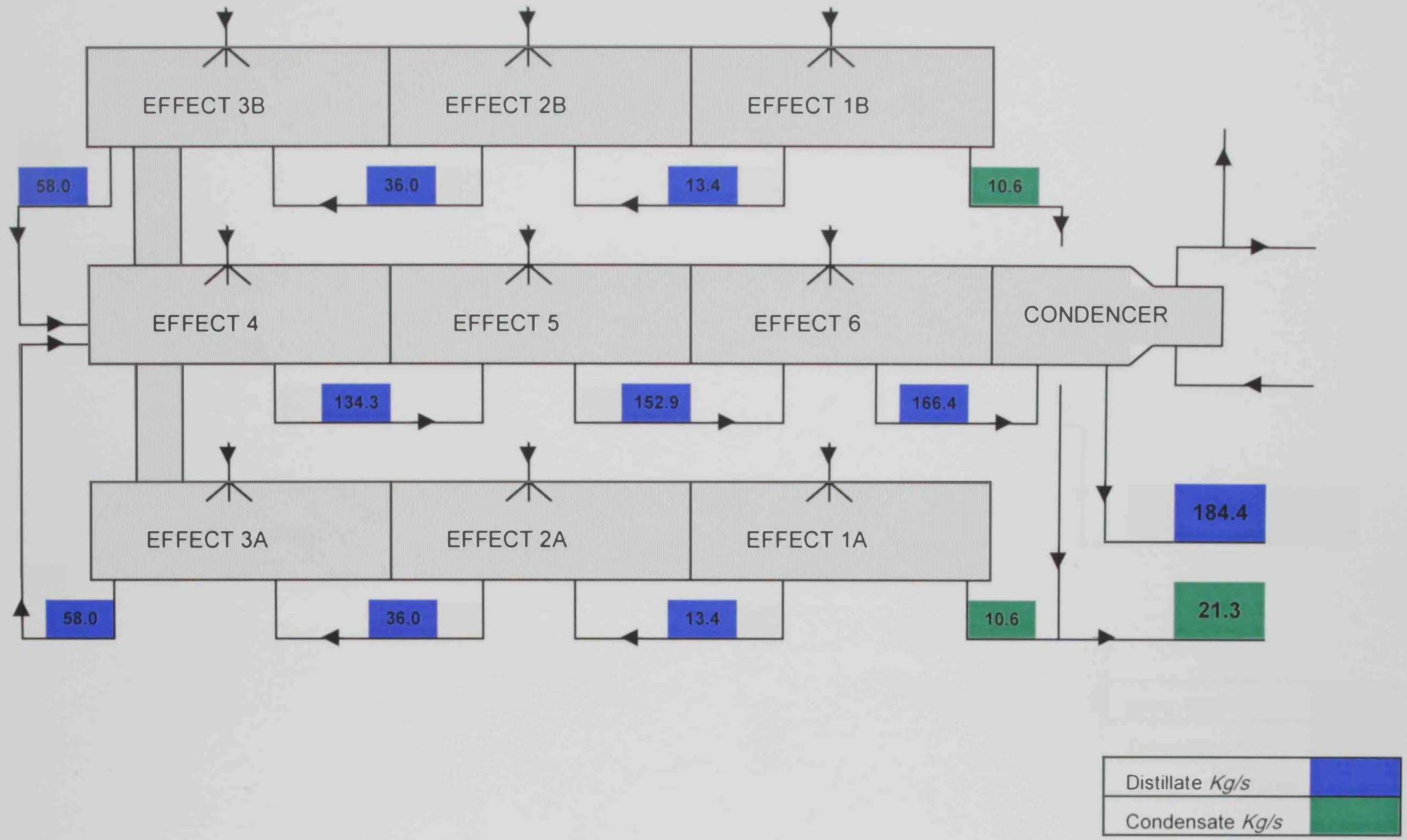
		STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B.	E.B.
Condenser		F.Condenser		43.00				179.78	2,580.28		
	In	Mf+Mcw	1,398.36	33.00	52.00		3.92			1,582.80	257919.7
		Md6	166.35	43.50				181.86			
		Mv6	18.09	43.00					2,580.28		
	Internal	Vdc'	0.14	43.00					2,580.28		
		Ldc	166.21	43.00				179.78			
		Mdc	18.24	43.00				179.78			
	Out	Mcw	758.36	40.98	52.00		3.93			1,582.80	258153.8
		Mf	640.00	40.98	52.00		3.93				
		Distillate	184.44	43.00				179.78	2,580.28		
										Err %	-0.09

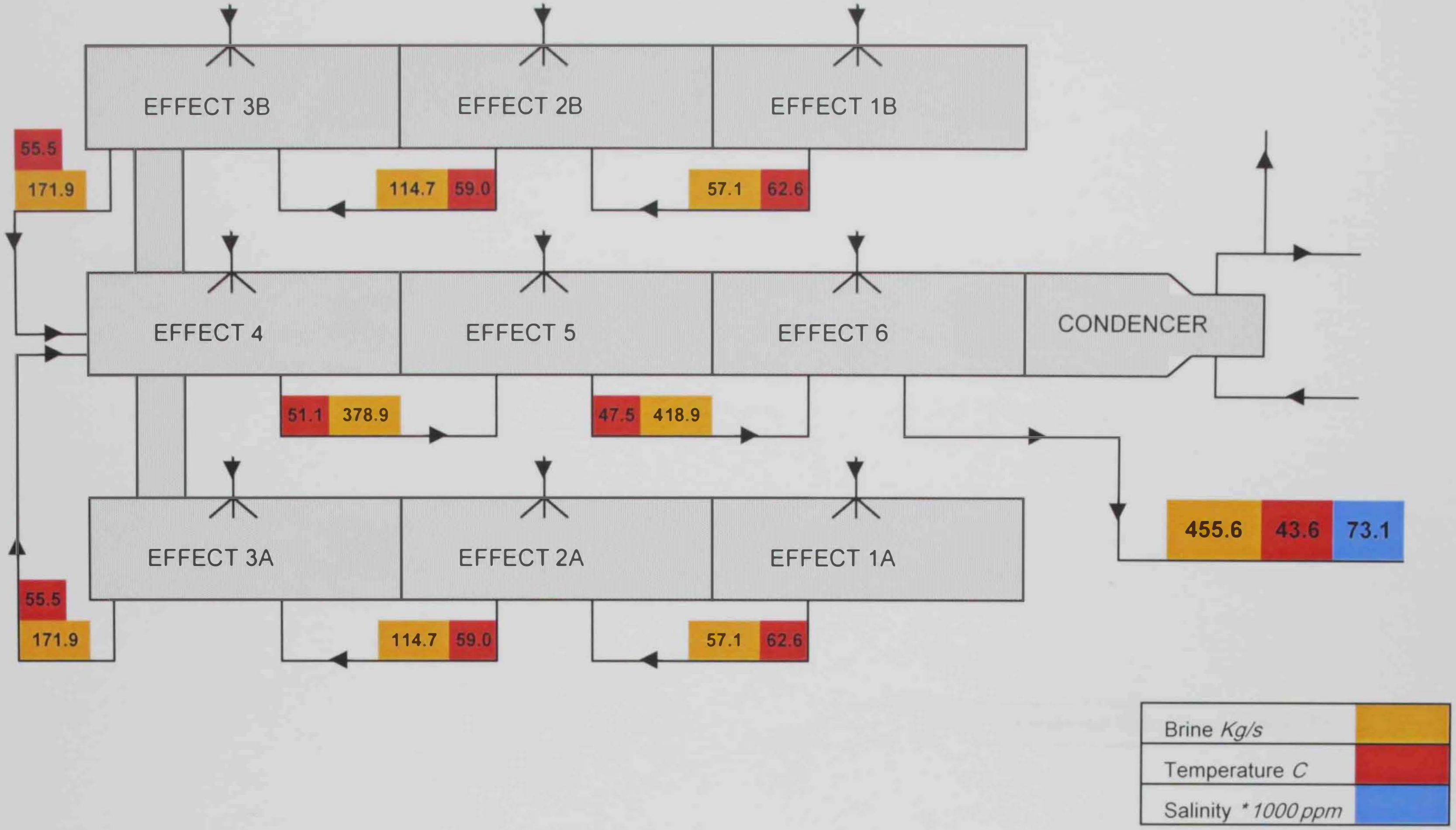
U	3.049596459
A	2874

Q	Tvc	delta T lm	Qcal	Err
43773.423	43.00	4.99	43775.40543	-1.982654161

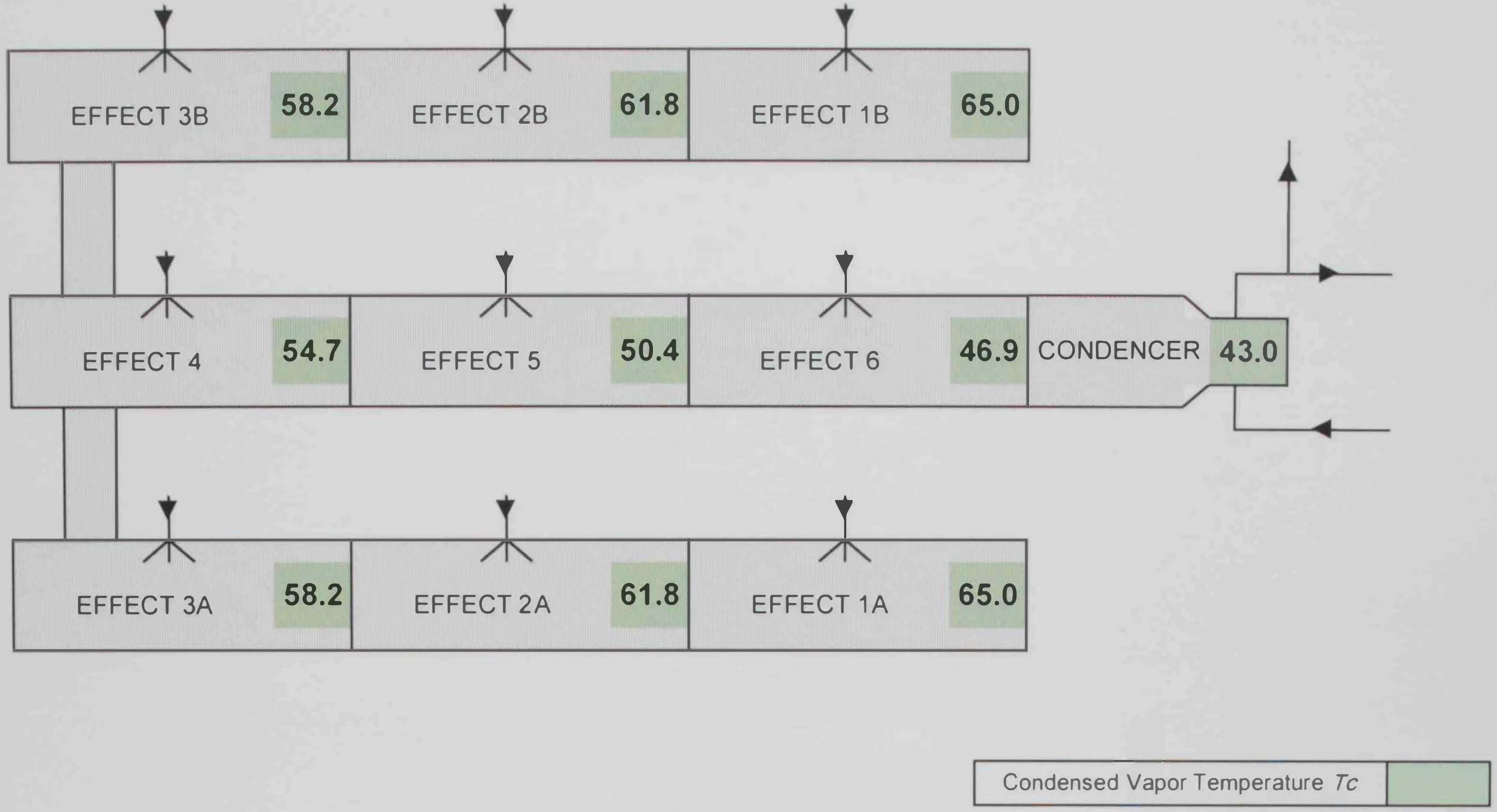
** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T	A	B	C	D	Cp
Mf+Mcw in	52.00	33.00	3895.8024	1.076734	-0.010716138	6.76017E-05	3.922094111
Mf & Mcw out	52.00	40.98	3895.8024	1.076734	-0.010716138	6.76017E-05	3.926581471


EFFECT 3B
EFFECT 2B
EFFECT 1B
58.0
36.0
13.4
10.6
EFFECT 4
EFFECT 5
EFFECT 6
CONDENCER
134.3
152.9
166.4
184.4
EFFECT 3A
EFFECT 2A
EFFECT 1A
58.0
36.0
13.4
10.6
21.3
Distillate Kg/s
Condensate Kg/s



EFFECT 3B
EFFECT 2B
EFFECT 1B
55.5
171.9
114.7
59.0
57.1
62.6
EFFECT 4
EFFECT 5
EFFECT 6
CONDENCER
51.1
378.9
47.5
418.9
455.6
43.6
73.1
EFFECT 3A
EFFECT 2A
EFFECT 1A
55.5
171.9
114.7
59.0
57.1
62.6
Brine Kg/s
Temperature C
Salinity *1000 ppm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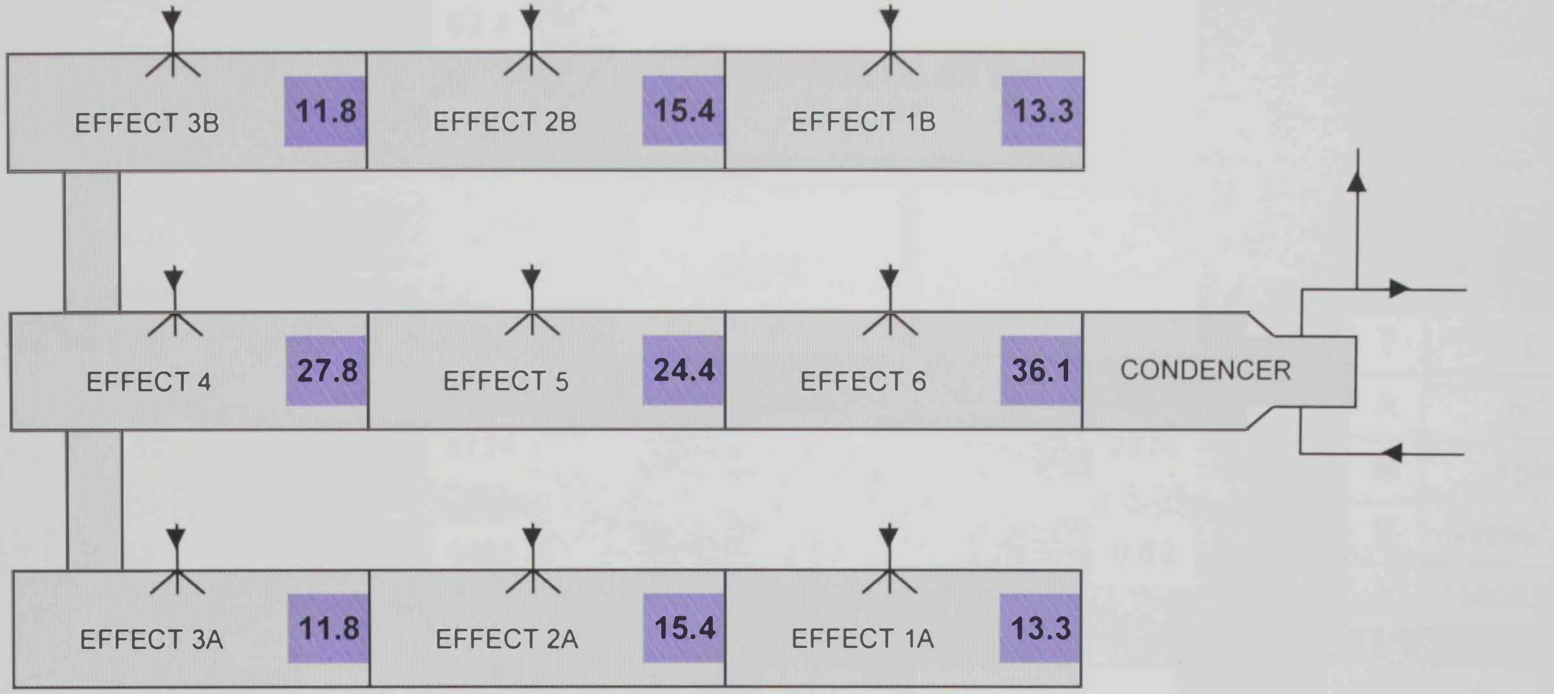
EFFECT 2B
61.8
EFFECT 1B
65.0
EFFECT 4
54.7
EFFECT 5
50.4
EFFECT 6
46.9
CONDENCER
43.0
EFFECT 3A
58.2
EFFECT 2A
61.8
EFFECT 1A
65.0
Condensed Vapor Temperature Tc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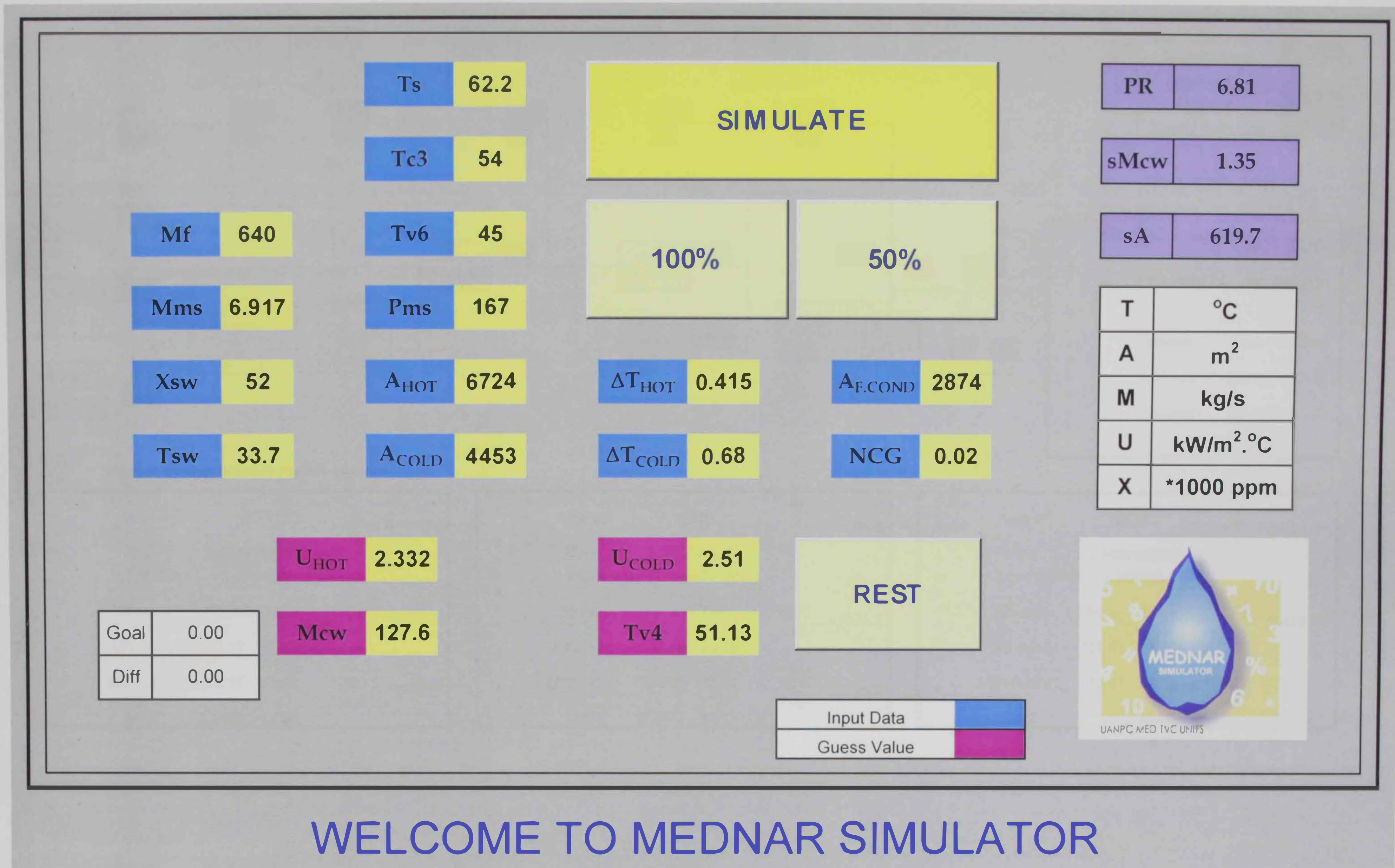
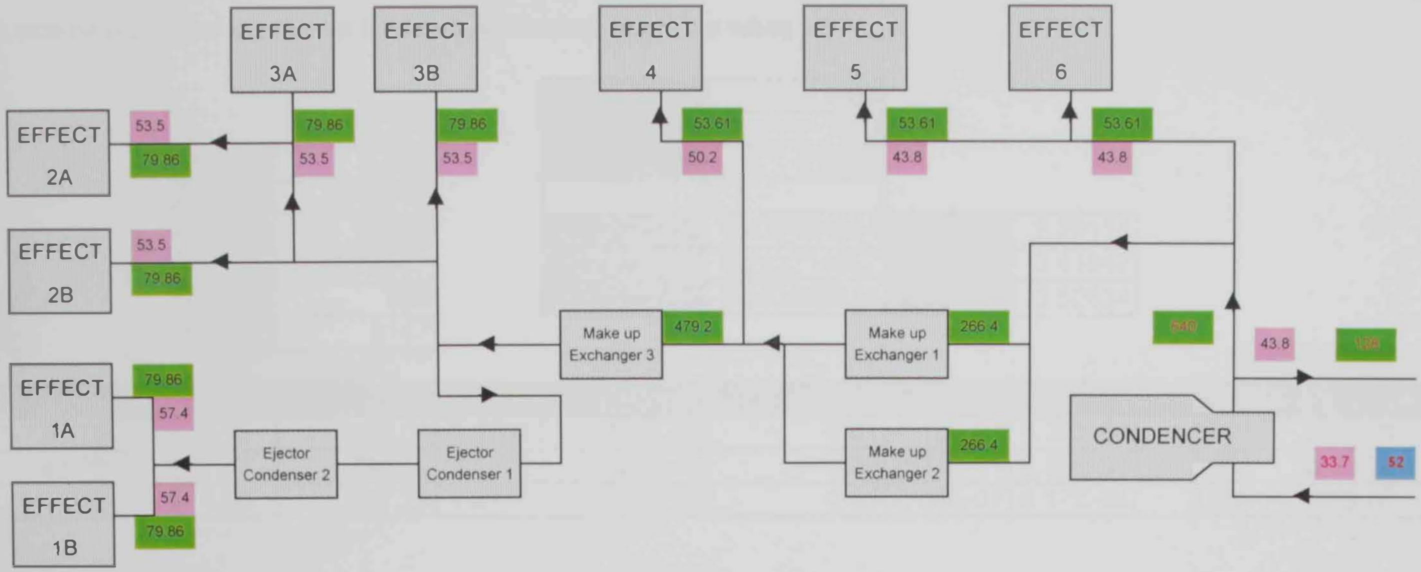
EFFECT 4
27.8
EFFECT 5
24.4
EFFECT 6
36.1
CONDENCER
EFFECT 3A
11.8
EFFECT 2A
15.4
EFFECT 1A
13.3
Vapor Velocity m/s


Ts 62.2
Tc3 54
SIMULATE
PR 6.81
sMcw 1.35
sA 619.7
Mf 640
Tv6 45
100%
50%
Mms 6.917
Pms 167
T °C
A m2
M kg/s
U kW/m2.°C
X *1000 ppm
Xsw 52
AHOT 6724
ΔTHOT 0.415
AF.COND 2874
Tsw 33.7
ACOLD 4453
ΔTCOLD 0.68
NCG 0.02
UHOT 2.332
UCOLD 2.51
REST
Goal 0.00
Diff 0.00
Mcw 127.6
Tv4 51.13
Input Data
Guess Value
MEDNAR
SIMULATOR
UANPC MED-TVC UNITS
WELCOME TO MEDNAR SIMULATOR



Temperature *C*	
Pressure *Kpa*	
Flow *Kg/s*	
Salinity **1000 ppm*	

Ej. Condenser			
			2823.237993
	QEC	2475.24	888.0767191
	stream	Flow	Temp
	Mf(inlet)	159.72	53.46
	Mf(outlet)	159.72	57.40

**Mass & Energy balance for Pre-Heaters

Pre-heater 3					
Area m^2	693		Q3	6101.31
U_{HE3}	3.457416621			
3.525339613	3.5257317			
Stream	Flow	Temp	Q3 in	6143.7
Mf2(inlet)	479.17	50.22		
Mf2(outlet)	479.17	53.46	Q3 out	6143.7
Md3"	2.5868	54.68		

Pre-heater 12				
Area m^2	1156		Q12	12182.4
U_{HE12}	3.343722937			
3.410471931	3.409477155			
Stream	Flow	Temp	Q12 in	13635.2
Mf1(inlet)	532.78	43.75		
Mf1(outlet)	532.78	50.22	Q12 out	13635.2
Md4"	5.7272	51.13		

F. Condenser				
Area m^3	2874		QFC	30261.14
U_{FC}	3.148049669			
2.291136745	2.3067454			
Stream	Flow	Temp	QFC in	30447.65
Mf (inlet)	767.61	33.70		
Mf (outlet)	767.61	43.75	QFC out	0
Mdc		45.00		

** Local heat transfer coefficient for seawater flowing inside the tubes

dout	0.019
dinner	0.0176
salinity	52000
n12	1070.66667
n3	963
nfc	3147

Ri	0.08		
Ro	0		
K	0.0165		
1/UFC	0.436464564	UFC	2.29114
1/UHE12	0.293214552	UHE12	3.41047
1/UHE3	0.283660614	UHE3	3.52534

temperature oC	hout	Ditus-Bolter	velocity	Reynolds	Prandtl	Kwater	muwater	cpwater	rhowater
33.70	3.76	408.05	0.9708	21211.08	0.01	6.19E-01	8.32E-04	3.92	1032.79
43.75	7.47	798.59	1.9882	52225.49	0.00	6.31E-01	6.89E-04	3.93	1028.75
50.22	7.98	847.25	1.9937	58320.95	0.00	6.38E-01	6.17E-04	3.93	1025.86

** Local heat transfer coefficient on the outside surface of tubes

dout	0.019	xncondout	0

con1	1.538923	con2	1
num1	0.854182	dinner	0.0176
mumt	2.293731		

temperature oC	hout	delt	rhovapor	rhocond	kliquid	lamdav	muliquid	mass	velocityout
45.00	53.26	0.1	0.066	989.845	6.35E-04	2394.53	5.98E-04	12.64	22.88340011
51.13	54.94	0.1	0.088	987.173	6.42E-04	2379.80	5.38E-04	5.73	17.96367406
54.68	55.87	0.1	0.103	985.527	6.45E-04	2371.23	5.08E-04	2.59	7.781995123

** STREAMS DATA & RESULTS

		STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B.	E.B.
Cell #1		Cell #1		59.92				250.65	2,609.84		
	In	Ms	10.83	62.20					2,613.73	90.69	46332.98
	In	Mf01	79.86	57.40	52.00		3.94				
	Out	Mb1	69.70	60.61	59.58	0.74	3.90			90.69	45823.43
	Out	Mv1	10.16	59.92					2,609.84		
	Out	Md1	3.91	62.20				260.20			
	Out	Mcon	6.92	62.20				260.20			
										Err %	1.10

U	2.332132721
A	6724
ΔT loss Hot	0.415

Q	T1	delta T	Qcal	Err
24967.662	60.61	1.59	24967.66185	-5.09317E-11

NCG	0.02

** BOILING POINT ELEVATION

STREAM #	X	T (K)	BPE	Tv1 (°C)	Hv1
Mb1(assume)1	0.059581015	333.76	0.7358878		
LOOP 1	0.051	333.76	0.6177527	59.99	2609.95329
LOOP 2	0.055757684	333.76	0.6826125	59.92	2609.842136
LOOP 3	0.055790285	333.76	0.6830624	59.92	2609.841365
LOOP 4	0.055790506	333.76	0.6830655	59.92	2609.84136
LOOP 5	0.055790507	333.76	0.6830655	59.92	2609.84136

** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T	A	B	C	D	Cp
Mf01	52.00	57.40	3895.8024	1.076734	-0.010716138	6.76017E-05	3.935085544
LOOP 1	50	60.61	3906.535	1.014725	-0.0100305	6.54708E-05	3.945765853
LOOP 2	59.51536793	60.61	3856.3512	1.293499	-0.013157347	7.52925E-05	3.903178653
LOOP 3	59.58056921	60.61	3856.015	1.295267	-0.013177592	7.5357E-05	3.902889634
LOOP 4	59.58101179	60.61	3856.0127	1.295279	-0.01317773	7.53575E-05	3.902887672
LOOP 5	59.58101479	60.61	3856.0127	1.295279	-0.013177731	7.53575E-05	3.902887659

Cp= (A+BT+CT^2+DT^3)E-3

WHERE
T= TEMPERATURE C
S= WATER SALINITY G/Kg
A= 4206.8-6.6197S+1.2288E-2S^2
B= -1.1262+5.4178E-2S-2.2719E-4S^2'
C= 1.2026E-2-5.3566E-4S+1.8906E-6S^2
D= 6.87774E-7+1.517E-6S-4.4268E-9S^2

**SALINITY

STREAM #	Assumed X	Mb1	Mv1	Cp	Calc X
LOOP 1	50	83.05556	-3.194444	3.945766	59.51536793
LOOP 2	59.51536793	69.77656	10.084549	3.903179	59.58056921
LOOP 3	59.58056921	69.7002	10.160908	3.90289	59.58101179
LOOP 4	59.58101179	69.69969	10.161426	3.902888	59.58101479
LOOP 5	59.58101479	69.69968	10.161429	3.902888	59.58101481

** STREAMS DATA & RESULTS

		STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B.	E.B.
Cell #2		Cell #2		57.27				239.53	2,605.28		
	In	Mf02	79.86	53.46	52.00		3.93			163.63	60,816.01
		Md1	3.91	62.20				260.20			
		Mb1	69.70	60.61	59.58	0.74	3.90				
		Mv1	10.16	59.92				250.65	2,609.84		
	Internal	Md2*	10.16	59.51				248.91	2,609.13		
		Mb2*	70.32	57.98	59.06	0.71	3.90				
		Mv2*	9.54	57.27					2,605.28		
		Ld2	3.87	57.27				239.53			
		Lb2	69.40	57.98	59.84	0.73	3.90				
		Vd2'	0.03	57.27					2,605.28		
		Vb2'	0.30	57.27					2,605.28		
	Out	Md2	14.04	58.89				246.32		163.63	60,802.09
		Mb2	139.72	57.98	59.45	0.72	3.90				
		Mv2	9.88	57.27				239.53	2,605.28		
										Err %	0.02

U	2.332132721
A	6724
dlta loss	0.415

Q	T2	delta T	Qcal	Err
23983.202	57.98	1.53	23983.20245	4.73E-11

** BOILING POINT ELEVATION

STREAM #	X	T2 (K)	BPE	Tv2 (°C)	Hv2
Mb2	0.059445535	331.13	0.7208063		
Lb2	0.059837595	331.13	0.7262256		
Mb2' (LOOP 1)	0.053990035	331.13	0.6465181	57.33	2605.387738
Mb2' (LOOP 2)	0.058489934	331.13	0.707643	57.27	2605.282354
Mb2' (LOOP 3)	0.058519135	331.13	0.7080443	57.27	2605.281662
Mb2' (LOOP 4)	0.058519323	331.13	0.7080469	57.27	2605.281657
Mb2' (LOOP 5)	0.058519324	331.13	0.7080469	57.27	2606.281657

** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T2	A	B	C	D	Cp
Mf02	52.00	53.46	3895.8024	1.076734	-0.010716138	6.76017E-05	3.933068
Mb2	59.45	57.98	3856.7114	1.291603	-0.013135646	7.52233E-05	3.902102
Lb2	5984	57.98	3854.6907	1.302219	-0.013257241	7.56111E-05	3.900364
Mb2' (LOOP 1)	50	57.98	3906.535	1.014725	-0.0100305	6.54708E-05	3.94441
Mb2' (LOOP 2)	58.99979802	57.98	3859.0133	1.279448	-0.012996698	7.47809E-05	3.904081
Mb2' (LOOP 3)	59.05820059	57.98	3858.7114	1.281046	-0.013014947	7.4839E-05	3.903821
Mb2' (LOOP 4)	59.05857612	57.98	3858.7094	1.281056	-0.013015064	7.48393E-05	3.90382
Mb2' (LOOP 5)	59.05857853	57.98	3858.7094	1.281056	-0.013015065	7.48393E-05	3.90382

Cp= (A+BT+CT^2+DT^3)E-3

WHERE

T= TEMPERATURE C
S= WATER SALINITY G/Kg
A= 4206.8-6.6197S+1.2288E-2S^2
B= -1.1262+5.4178E-2S-2.2719E-4S^2'
C= 1.2026E-2-5.3566E-4S+1.8906E-6S^2
D= 6.87774E-7+1.517E-6S-4.4268E-9S^2

**MODUL

STREAM #	Assumed X2'	Mb2'	Mv2'	Cp	Calc X2'
Mb2' (LOOP 1)	50	83.05556	-3.19	3.94441	58.99979802
Mb2' (LOOP 2)	58.99979802	70.3863	9.47	3.904081	59.05820059
Mb2' (LOOP 3)	59.05820059	70.3167	9.54	3.903821	59.05857612
Mb2' (LOOP 4)	59.05857612	70.31625	9.54	3.90382	59.05857853
Mb2' (LOOP 5)	59.05857853	70.31625	9.54	3.90382	59.05857855

** STREAMS DATA & RESULTS

		STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B.	E.B.
Cell #3		Cell #3		54.00				225.81	2,599.62		
	In	Mf03	79.86	53.46	52.00		3.93			243.49	77,594.35
		Md2	14.04	58.89				246.32			
		Mb2	139.72	57.98	59.45	0.72	3.90				
		Mv2	9.88	57.27				239.53	2,605.28		
	Internal	Md3'	9.88	56.86				237.79	2,604.57		
		Mb3'	70.25	55.37	59.12	0.68	3.90				
		Mv3'	9.61	54.69					2,600.81		
		Ld3	13.91	54.00				225.81			
		Lb3	139.12	55.37	59.70	0.71	3.90				
		Vd3	0.12	54.00					2,599.62		
		Vb3	0.60	54.69					2,600.81		
	Out	Md3*	23.79	55.14				230.58		243.49	77,569.23
		Mb3	209.37	55.37	59.50	0.71	3.90				
		Mv3*	6.42	54.68					2,600.80		
		Mev (tvc)in	3.91	54.68					2,600.80		
										Err %	0.03

U	2.332132721
A	6724
dlta loss	0.415

Q	T3	delta T	Qcal	Err
23378.728	55.37	1.49	23378.72764	0

Md3"
2.586822138

** BOILING POINT ELEVATION

STREAM #	X	T3 (K)	BPE	Tv3 (°C)	Hv3
Mb3	0.05950444	328.52	0.7085986		
Lb3	0.059700759	328.52	0.7112625		
Mb3' (LOOP 1)	0.052683076	328.52	0.6176838	54.75	2600.918147
Mb3' (LOOP 2)	0.057212654	328.52	0.6776992	54.69	2600.814072
Mb3' (LOOP 3)	0.057240728	328.52	0.6780753	54.69	2600.81342
Mb3' (LOOP 4)	0.057240901	328.52	0.6780777	54.69	2600.813416
Mb3' (LOOP 5)	0.057240902	328.52	0.6780777	54.69	2600.813416

** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T3	A	B	C	D	Cp
Mf03	52.00	53.46	3895.8024	1.076734	-0.010716138	6.76017E-05	3.933068
Mb3	59.50	55.37	3856.4075	1.293202	-0.013153953	7.52817E-05	3.900462
Lb3	59.70	55.37	3855.3955	1.298522	-0.013214869	7.54759E-05	3.89959
Mb3' (LOOP 1)	50	55.37	3906.535	1.014725	-0.0100305	6.54708E-05	3.943081
Mb3' (LOOP 2)	59.05917496	55.37	3858.7064	1.281072	-0.013015251	7.48399E-05	3.902439
Mb3' (LOOP 3)	59.11530404	55.37	3858.4163	1.282606	-0.013032777	7.48957E-05	3.90219
Mb3' (LOOP 4)	59.11564851	55.37	3858.4145	1.282616	-0.013032884	7.48961E-05	3.902188
Mb3' (LOOP 5)	59.11565062	55.37	3858.4145	1.282616	-0.013032885	7.48961E-05	3.902188

Cp= (A+BT+CT^2+DT^3)E-3

WHERE
T= TEMPERATURE C
S= WATER SALINITY G/Kg
A= 4206.8-6.6197S+1.2288E-2S^2
B= -1.1262+5.4178E-2S-2.2719E-4S^2'
C= 1.2026E-2-5.3566E-4S+1.8906E-6S^2
D= 6.87774E-7+1.517E-6S-4.4268E-9S^2

**MODUL

STREAM #	Assumed X3'	Mb3'	Mv3'	Cp	Calc X3'
Mb3' (LOOP 1)	50	83.05556	-3.19	3.943081	59.05917496
Mb3' (LOOP 2)	59.05917496	70.31554	9.55	3.902439	59.11530404
Mb3' (LOOP 3)	59.11530404	70.24878	9.61	3.90219	59.11564851
Mb3' (LOOP 4)	59.11564851	70.24837	9.61	3.902188	59.11565062
Mb3' (LOOP 5)	59.11565062	70.24836	9.61	3.902188	59.11565064

Steam Jet Ejector Formula: A/B

Ra	Mms	Mev	Ps	Pv	Pms	Tv	PCF	TCF	Ms
1.769689	6.91667	3.908411	22.02731	15.38045	166.9936	54.68	1.468172	0.971952	10.82508

** STREAMS DATA & RESULTS

		STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B	E.B
Cell #4		Cell #4		51.15				213.88	2,594.65		
	In	Mf04	53.61	50.22	52.00		3.93			532.78	139,262.46
		Md3	50.17	55.14				230.58			
		Mb3	418.74	55.37	59.50	0.71	3.90				
		Mv3	10.26	54.68				228.66	2,600.80		
	Internal	Md4'	10.26	54.00				225.81	2,599.62		
		Mb4'	43.51	51.82	64.07	0.67	3.88				
		Mv4'	10.10	51.15					2,594.65		
		Ld4	49.82	51.15				213.88			
		Lb4	416.12	51.82	59.88	0.70	3.90				
		Vd4	0.35	51.15					2,594.65		
		Vb4	2.62	51.15					2,594.65		
	Out	Md4*	60.08	51.76				216.44		532.78	139,693.55
		Mb4	459.63	51.82	60.28	0.70	3.90				
		Mv4*	13.07	51.15					2,594.65		
										Err %	-0.31

U	2.510295657
A	4453
ΔT loss Cold	0.68

Q	T4	delta T	Qcal	Err
24352.436	51.82	2.18	24352.4358	0

Md4"
5.727203609

** BOILING POINT ELEVATION

STREAM #	X	T4 (K)	BPE	Tv4 (°C)	Hv3
Mb4	0.060275668	324.97	0.7013211		
Lb4	0.059878934	324.97	0.6960587		
Mb4' (LOOP 1)	0.050910734	324.97	0.5799125	51.24	2594.813156
Mb4' (LOOP 2)	0.057901501	324.97	0.6699897	51.15	2594.655715
Mb4' (LOOP 3)	0.057945353	324.97	0.6705649	51.15	2594.65471
Mb4' (LOOP 4)	0.057945624	324.97	0.6705685	51.15	2594.654703
Mb4' (LOOP 5)	0.057945626	324.97	0.6705685	51.15	2594.654703

1.30524

** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T4	A	B	C	D	Cp
Mf04	52.00	50.22	3895.8024	1.076734	-0.010716138	6.76017E-05	3.931413
Mb4	60.28	51.82	3852.4374	1.313998	-0.013392419	7.50427E-05	3.895148
Lb4	59.88	51.82	3854.4779	1.303334	-0.013270029	7.56519E-05	3.89691
Mb4' (LOOP 1)	50	51.82	3906.535	1.014725	-0.0100305	6.54708E-05	3.941294
Mb4' (LOOP 2)	63.98153378	51.82	3833.564	1.410158	-0.014506919	7.9626E-05	3.878764
Mb4' (LOOP 3)	64.06923873	51.82	3833.1215	1.412358	-0.014532666	7.97094E-05	3.878378
Mb4' (LOOP 4)	64.06978054	51.82	3833.1187	1.412372	-0.014532825	7.97099E-05	3.878375
Mb4' (LOOP 5)	64.06978389	51.82	3833.1187	1.412372	-0.014532826	7.97099E-05	3.878375

Cp= (A+BT+CT^2+DT^3)E-3

WHERE

T=	TEMPERATURE C
S=	WATER SALINITY G/Kg
A=	4206.8-6.6197S+1.2288E-2S^2
B=	-1.1262+5.4178E-2S-2.2719E-4S^2'
C=	1.2026E-2-5.3566E-4S+1.8906E-6S^2
D=	6.87774E-7+1.517E-6S-4.4268E-9S^2

**MODUL

STREAM #	Assum d X4'	Mb4'	Mv4'	Cp	Calc X4'
Mb4' (LOOP 1)	50	55.75556	-2.14	3.941294	63.98153378
Mb4' (LOOP 2)	63.98153378	43.5716	10.04	3.878764	64.06923873
Mb4' (LOOP 3)	64.06923873	43.51195	10.10	3.878378	64.06978054
Mb4' (LOOP 4)	64.06978054	43.51159	10.10	3.878375	64.06978389
Mb4' (LOOP 5)	64.06978389	43.51158	10.10	3.878375	64.06978391

** STREAMS DATA & RESULTS

		STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B	E.B
Cell #5		Cell #5		48.30				201.92	2,589.65		
	In	Mf05	53.61	43.75	52.00		3.93			586.39	135,286.33
		Md4	65.81	51.76				216.44			
		Mb4	459.63	51.82	60.28	0.70	3.90				
		Mv4	7.34	51.15				213.88	2,594.65		
	Internal	Md5'	7.34	50.47				211.03	2,593.47		
		Mb5'	46.73	48.91	59.65	0.61	3.90				
		Mv5'	6.88	48.30					2,589.65		
		Ld5	65.41	48.30				201.92			
		Lb5	457.46	48.91	60.56	0.69	3.89				
		Vd5	0.40	48.30					2,589.65		
		Vb5	2.16	48.30					2,589.65		
	Out	Md5	72.75	48.73				203.73		586.39	135,260.37
		Mb5	504.20	48.91	60.48	0.69	3.89				
		Mv5	9.44	48.30					2,589.65		
										Err %	0.02

U	2.510295657
A	4453
dlta loss	0.68

Q	T5	delta T	Qcal	Err
17494.348	48.91	1.57	17494.34846	0

** BOILING POINT ELEVATION

STREAM #	X	T5 (K)	BPE	Tv5 (°C)	Hv3
Mb5	0.060476727	322.06	0.6894685		
Lb5	0.060560881	322.06	0.690564		
Mb5' (LOOP 1)	0.049452939	322.06	0.5499701	48.36	2589.753807
Mb5' (LOOP 2)	0.054252777	322.06	0.6097496	48.30	2589.648654
Mb5' (LOOP 3)	0.054279268	322.06	0.6100836	48.30	2589.648067
Mb5' (LOOP 4)	0.054279413	322.06	0.6100854	48.30	2589.648064
Mb5' (LOOP 5)	0.054279414	322.06	0.6100854	48.30	2589.648064

** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T5	A	B	C	D	Cp
Mf05	52.00	43.75	3895.8024	1.076734	-0.010716138	6.76017E-05	3.928059
Mb5	60.48	48.91	3851.4048	1.319376	-0.013454218	7.62402E-05	3.892668
Lb5	60.56	48.91	3850.9729	1.321621	-0.013480039	7.63228E-05	3.892294
Mb5' (LOOP 1)	50	48.91	3906.535	1.014725	-0.0100305	6.54708E-05	3.939828
Mb5' (LOOP 2)	59.59967572	48.91	3855.9165	1.295785	-0.013183522	7.5376E-05	3.896573
Mb5' (LOOP 3)	59.65265827	48.91	3855.6434	1.29722	-0.013199957	7.54284E-05	3.896337
Mb5' (LOOP 4)	59.65294751	48.91	3855.6419	1.297228	-0.013200047	7.54286E-05	3.896335
Mb5' (LOOP 5)	59.65294909	48.91	3855.6419	1.297228	-0.013200047	7.54286E-05	3.896335

Cp= (A+BT+CT^2+DT^3)E-3

WHERE

T=	TEMPERATURE C
S=	WATER SALINITY G/Kg
A=	4206.8-6.6197S+1.2288E-2S^2
B=	-1.1262+5.4178E-2S-2.2719E-4S^2'
C=	1.2026E-2-5.3566E-4S+1.8906E-6S^2
D=	6.87774E-7+1.517E-6S-4.4268E-9S^2

**MODUL

STREAM #	Assumed X5'	Mb5'	Mv5'	Cp	Calc X5'
Mb5' (LOOP 1)	50	55.75556	-2.14	3.939828	59.59967572
Mb5' (LOOP 2)	59.59967572	46.77505	6.84	3.896573	59.65265827
Mb5' (LOOP 3)	59.65265827	46.7335	6.88	3.896337	59.65294751
Mb5' (LOOP 4)	59.652 4751	46.73328	6.88	3.896335	59.65294909
Mb5' (LOOP 5)	59.65294909	46.73328	6.88	3.896335	59.6529491

** STREAMS DATA & RESULTS

		STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B	E.B
Cell #6		Cell #6		45.00				188.14	2,583.83		
	In	Mf06	53.61	43.75	52.00		3.93			640.00	144,473.65
		Md5	72.75	48.73				203.73			
		Mb5	504.20	48.91	60.48	0.69	3.89				
		Mv5	9.44	48.30				201.92	2,589.65		
	Internal	Md6'	9.44	47.62				199.08	2,588.45		
		Mb6'	44.35	45.60	62.86	0.60	3.88				
		Mv6'	9.27	45.00					2,583.83		
		Ld6	72.28	45.00				188.14			
		Lb6	501.38	45.60	60.82	0.68	3.89				
		Vd6	0.47	45.00					2,583.83		
		Vb6	2.82	45.00					2,583.83		
	Out	Md6	81.72	45.59				190.61		640.00	144,780.17
		Mb6	545.73	45.60	60.98	0.68	3.89				
		Mv6	12.56	45.00					2,583.83		
										Err %	-0.21

U	2.510295657
A	4453
dlta loss	0.68

Q	T6	delta T	Qcal	Err
22561.871	45.60	2.02	22561.87147	0

** BOILING POINT ELEVATION

STREAM #	X	T6 (K)	BPE	Tv6 (°C)	Hv3
Mb6	0.060982951	318.75	0.6795769		
Lb6	0.060816507	318.75	0.6774581		
Mb6' (LOOP 1)	0.047798719	318.75	0.5171977	45.08	2583.97138
Mb6' (LOOP 2)	0.054196234	318.75	0.5946369	45.00	2583.834174
Mb6' (LOOP 3)	0.054230935	318.75	0.5950637	45.00	2583.833418
Mb6' (LOOP 4)	0.05423112	318.75	0.595066	45.00	2583.833414
Mb6' (LOOP 5)	0.054231121	318.75	0.595066	45.00	2583.833414

** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T6	A	B	C	D	Cp
Mf06	52.00	43.75	3895.8024	1.076734	-0.010716138	6.76017E-05	3.928059
Mb6	60.98	45.60	3848.8093	1.332833	-0.013609137	7.6736E-05	3.888563
Lb6	60.82	45.60	3849.6619	1.328421	-0.013558307	7.65732E-05	3.889304
Mb6' (LOOP 1)	50	45.60	3906.535	1.014725	-0.0100305	6.54708E-05	3.938156
Mb6' (LOOP 2)	62.79503082	45.60	3839.57	1.38005	-0.014155742	7.8492E-05	3.880506
Mb6' (LOOP 3)	62.86443179	45.60	3839.2177	1.381829	-0.01417643	7.85587E-05	3.880199
Mb6' (LOOP 4)	62.86480253	45.60	3839.2158	1.381838	-0.01417654	7.8559E-05	3.880197
Mb6' (LOOP 5)	62.86480451	45.60	3839.2158	1.381838	-0.014176541	7.8559E-05	3.880197

Cp= (A+BT+CT^2+DT^3)E-3

WHERE

T= TEMPERATURE C

S= WATER SALINITY G/Kg

A= 4206.8-6.6197S+1.2288E-2S^2

B= -1.1262+5.4178E-2S-2.2719E-4S^2'

C= 1.2026E-2-5.3566E-4S+1.8906E-6S^2

D= 6.87774E-7+1.517E-6S-4.4268E-9S^2

**MODUL

STREAM #	Assumed X6'	Mb6'	Mv6'	Cp	Calc X6'
Mb6' (LOOP 1)	50	55.75556	-2.14	3.938156	62.79503082
Mb6' (LOOP 2)	62.79503082	44.39488	9.22	3.880506	62.86443179
Mb6' (LOOP 3)	62.86443179	44.34587	9.27	3.880199	62.86480253
Mb6' (LOOP 4)	62.86480253	44.34561	9.27	3.880197	62.86480451
Mb6' (LOOP 5)	62.86480451	44.3456	9.27	3.880197	62.86480452

** STREAMS DATA & RESULTS

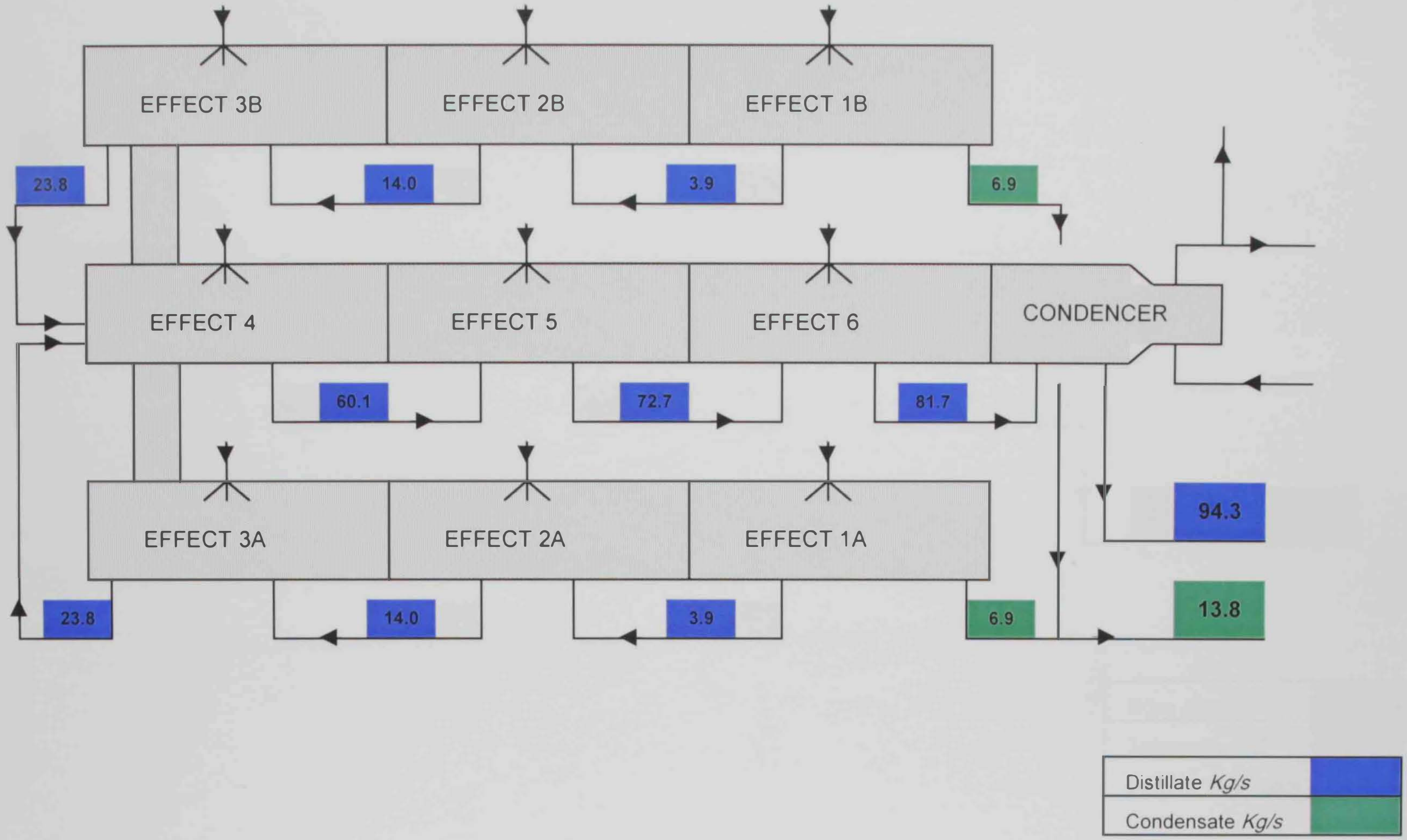
		STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B.	E.B.
Condenser		F.Condenser		45.00				188.14	2,583.83		
	In	Mf+Mcw	767.61	33.70	52.00		3.92				
		Md6	81.72	45.59				190.61		861.88	149487.2
		Mv6	12.56	45.00					2,583.83		
	Internal	Vdc'	0.08	45.00					2,583.83		
		Ldc	81.63	45.00				188.14			
		Mdc	12.64	45.00				188.14			
	Out	Mcw	127.61	43.75	52.00		3.93				
		Mf	640.00	43.75	52.00		3.93			861.88	149653.6
		Distillate	94.27	45.00				188.14	2,583.83		
										Err %	-0.11

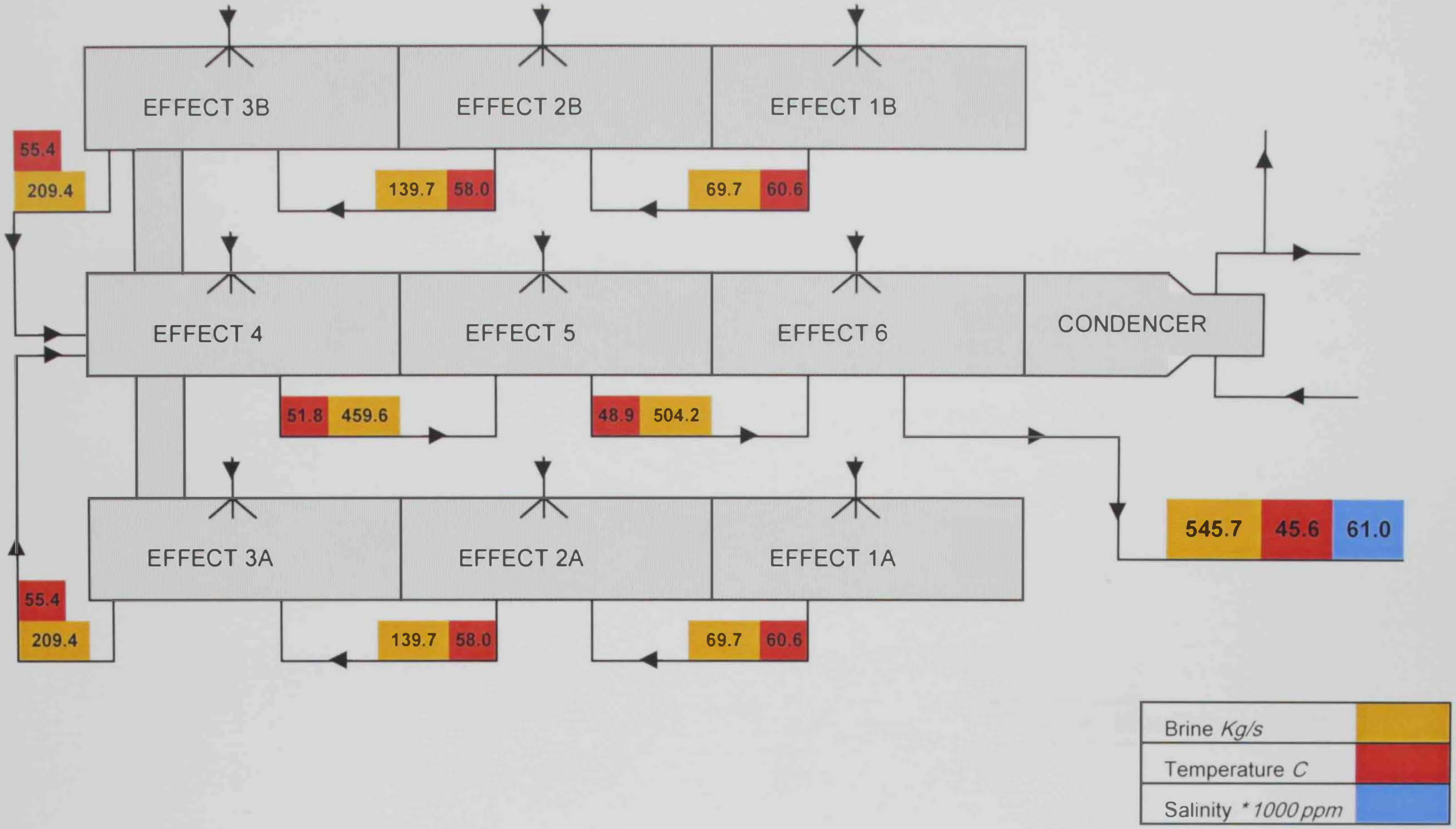
U	2.3067454
A	2874

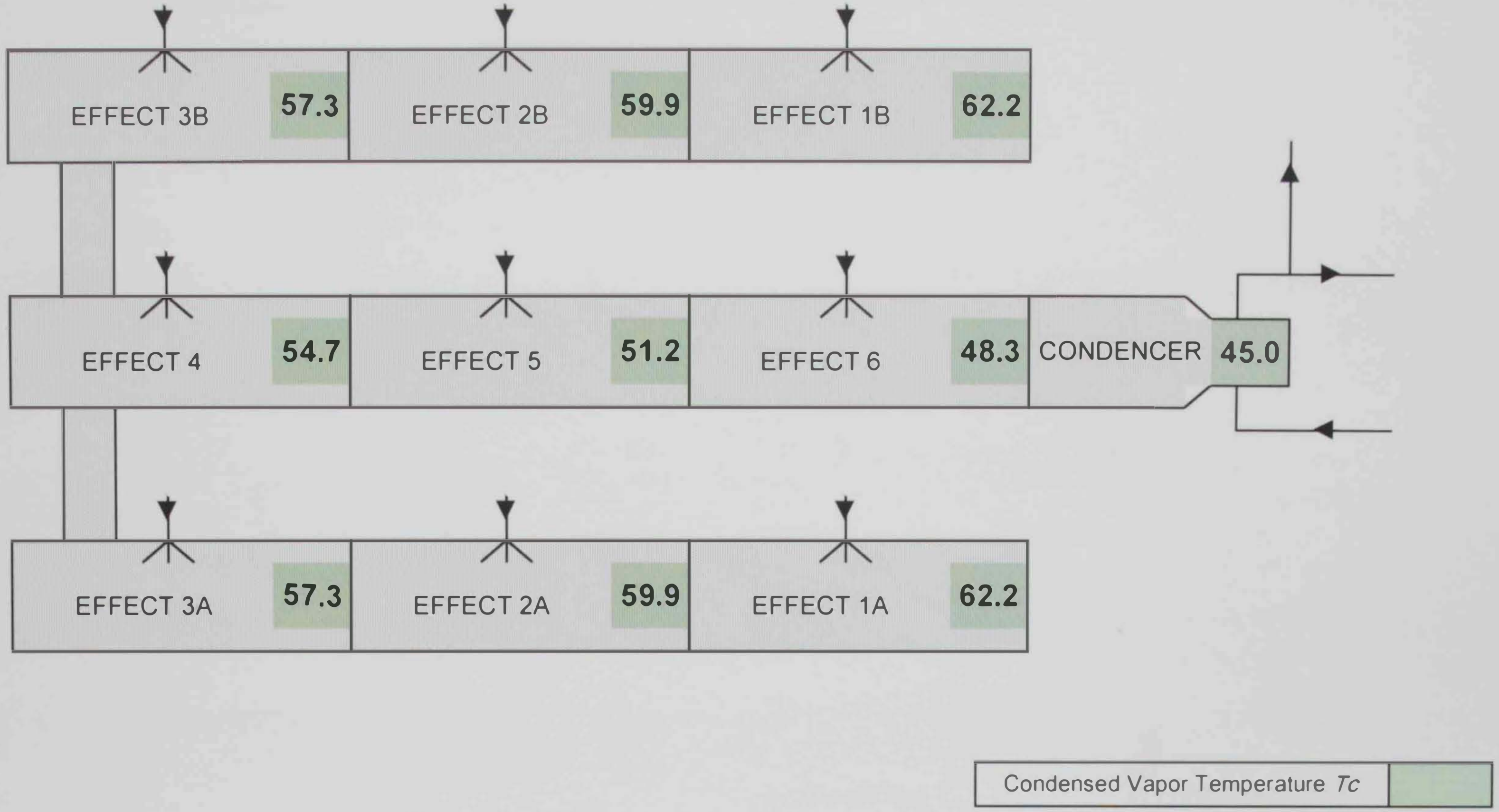
Q	Tvc	delta T lm	Qcal	Err
30281.418	45.00	4.57	30282.58517	-1.166970827

** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T	A	B	C	D	Cp
Mf+Mcw in	52.00	33.70	3895.8024	1.076734	-0.010716138	6.76017E-05	3.922505389
Mf & Mcw out	52.00	43.75	3895.8024	1.076734	-0.010716138	6.76017E-05	3.928059295


EFFECT 3B
EFFECT 2B
EFFECT 1B
23.8
14.0
3.9
6.9
EFFECT 4
EFFECT 5
EFFECT 6
CONDENCER
60.1
72.7
81.7
EFFECT 3A
EFFECT 2A
EFFECT 1A
94.3
23.8
14.0
3.9
6.9
13.8
Distillate Kg/s
Condensate Kg/s


EFFECT 3B
EFFECT 2B
EFFECT 1B
55.4
209.4
139.7
58.0
69.7
60.6
EFFECT 4
EFFECT 5
EFFECT 6
CONDENCER
51.8
459.6
48.9
504.2
EFFECT 3A
EFFECT 2A
EFFECT 1A
545.7
45.6
61.0
55.4
209.4
139.7
58.0
69.7
60.6
Brine Kg/s
Temperature C
Salinity *1000 ppm


EFFECT 3B
57.3
EFFECT 2B
59.9
EFFECT 1B
62.2
EFFECT 4
54.7
EFFECT 5
51.2
EFFECT 6
48.3
CONDENCER
45.0
EFFECT 3A
57.3
EFFECT 2A
59.9
EFFECT 1A
62.2
Condensed Vapor Temperature Tc


EFFECT 3B
7.8
EFFECT 2B
7.1
EFFECT 1B
6.5
EFFECT 4
18.0
EFFECT 5
14.8
EFFECT 6
22.9
CONDENCER
EFFECT 3A
7.8
EFFECT 2A
7.1
EFFECT 1A
6.5
Vapor Velocity m/s

APPENDIX E

Plant data and MEDNAR plant evaluation Run at Clean Condition

AUTHORITY | CONSULTING ENGINEER: MOTT MACDONALD

OJECT TITLE : ADDITION OF DESALINATION ANTS TO UNITS 9 & 10 IN UMM AL NAR WEST & RATING OF EXISTING UMM AL NAR EAST ODUCTION	PROJECT NO. ADWEA/PS/AD-135/268/98
IT NO. 9	LOCATION : UMM AL NAR WEST

Unit 9

Conditions:		
	Steam flow:	79.97 T/h
	Make-up flow:	2359.25 T/h
	Distillate flow:	674.34 T/h

Brine temperature for each effect:

Item	AKS	Temperature
Effect 1A	9WF11T1T	62.08
Effect 1B	9WF12T1T	61.67
Effect 2A	9WF21T1T	58.11
Effect 2B	9WF22T1T	57.70
Effect 3A	9WF31T1T	55.04
Effect 3B	9WF32T1T	55.00
Effect 4	9WF04T1T	51.77
Effect 5	9WF05T1T	48.93
Effect 6	9WF06T1T	45.22

Brine & Distillate level at last effect:

Item	AKS	Temperature
Cell 6 Brine level	9WF00L1I	152.14
Cell 6 Distillate level	9WP00L1I	227.73

r and on behalf of	NAME	SIGNATURE	DATE
JB-CONTRACTOR			
ONTRACTOR	LAPOINTE . R.		04/10/2000
IGINEER REP.	Osama Heikal	O. Heik	
LIENT	Atef Al-Baghdadi		

T. Ghanem 14/10/2000

idem	Revisions	0	1	2	3	Designation
	Date	5/97				Temperatures & levels at 100% load.
RUE DE CLICHY	Dess-Drawn	MB				
75009 PARIS	Appro	JPQ				

Dossier / File: P 7685

N°PLAN-DRWGN'

Revision

re - Job: 98401

Echelle-Scale

Page

100%

Date : 4-Oct-2000
Time : 10:08:31

REAL TIME VALUE LIST

DESALINATION UNIT 9

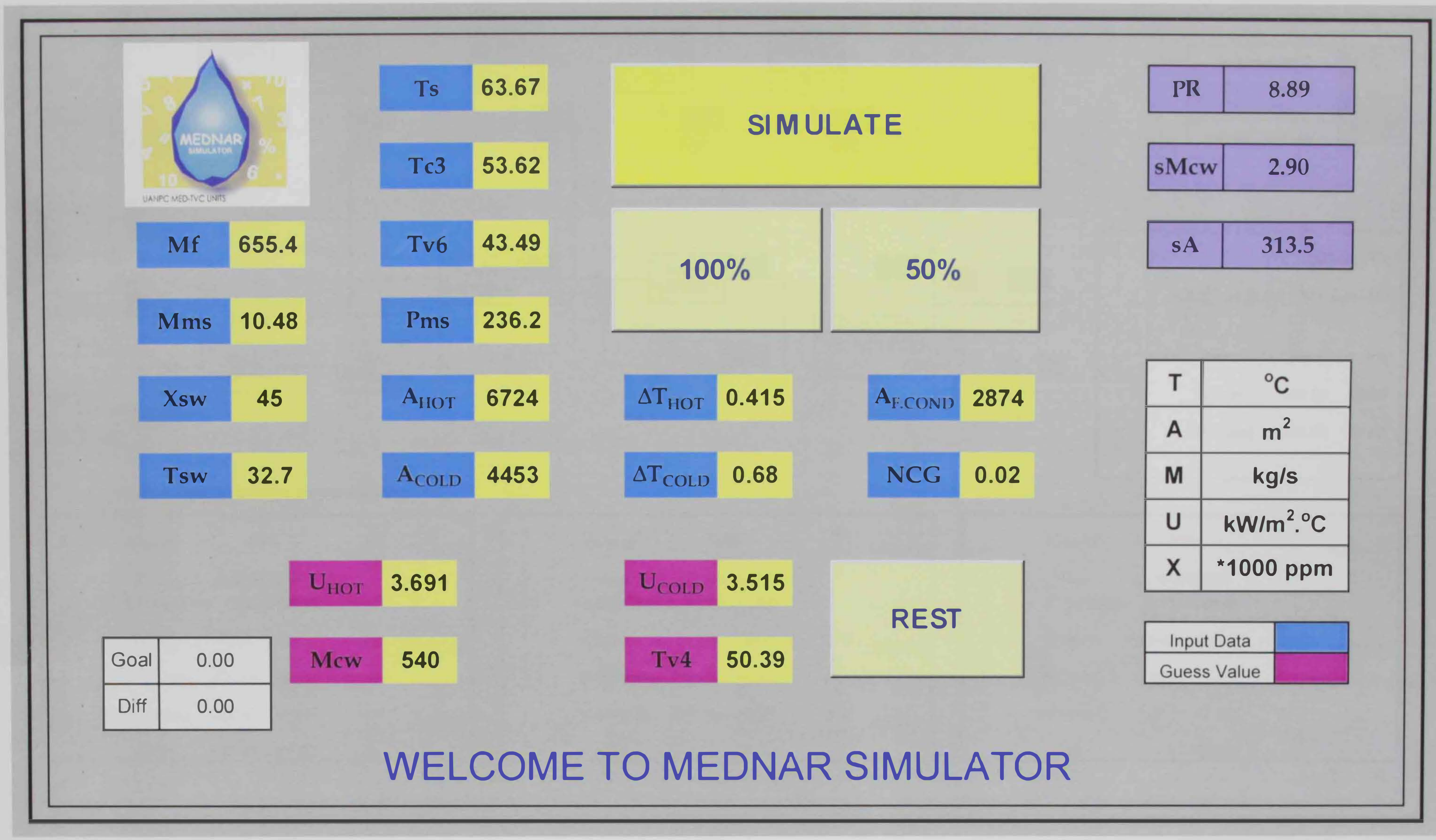
TAG	DESCRIPTION	TYPE	VALUE	UNIT
9WC10F1I	SEA WATER INLET FLOW	REAL	5219.47	t/h
9WC20T3T	SW INLET BEFORE PREHTR TEMP	REAL	32.90	°C
9WC20T4T	SW INLET TO CONDENSER TEMP	REAL	32.70	°C
9WC30T2T	SW AFTER CONDENSER TEMP	REAL	40.30	°C
9WE01T2T	EJECTOCOMP A OUTLET TEMP	REAL	63.77	°C
9WE02T2T	EJECTOCOMP B OUTLET TEMP	REAL	63.57	°C
9WE07P3T	CONDENSER PRESSURE	REAL	89.36	mbar_a
9WE07T1I	CONDENSER TEMPERATURE	REAL	43.22	°C
9WF00L1I	CELL 6 BRINE LEVEL	REAL	152.14	mm
9WF04T1T	CELL 4 BRINE OUTLET TEMP	REAL	51.77	°C
9WF05T1T	CELL 5 BRINE OUTLET TEMP	REAL	48.93	°C
9WF10T1T	CELL 6 BRINE OUTLET TEMP	REAL	45.22	°C
9WF11T1T	CELL 1A BRINE OUTLET TEMP	REAL	62.08	°C
9WF12T1T	CELL 1B BRINE OUTLET TEMP	REAL	61.67	°C
9WF20E1T	BRINE PUMP MOTOR CURRENT	REAL	140.62	A
9WF20F1T	BRINE DISCHARGE FLOW	REAL	1576.00	t/h
9WF20P3T	BRINE DISCHARGE PRESSURE	REAL	0.81	bar
9WF20T1T	BRINE PP DRV END BEAR TEMP	REAL	39.63	°C
9WF20T4T	BRINE PP SHAFT END BEAR TEMP	REAL	45.02	°C
9WF21T1T	CELL 2A BRINE OUTLET TEMP	REAL	58.11	°C
9WF22T1T	CELL 2B BRINE OUTLET TEMP	REAL	57.70	°C
9WF30T1I	CELL 3 BRINE OUTLET TEMP	REAL	55.03	°C
9WF31T1T	CELL 3A BRINE OUTLET TEMP	REAL	55.04	°C
9WF32T1T	CELL 3B BRINE OUTLET TEMP	REAL	55.00	°C
9WH10F1I	LP STEAM TO STEAM TRFR FLOW	REAL	79.97	t/h
9WH10P1T	LP STEAM TO STM TRFR PRES	REAL	1.65	bar
9WH10T1T	LP STEAM TO STM TRFR TEMP	REAL	135.78	°C
9WH11F1I	LP STEAM TO EJECTOCMP A FLOW	REAL	37.86	t/h
9WH11P1T	VAPOUR EJECTOCMP A DIFF PRES	REAL	-11.96	mbar
9WH11P3T	LP STEAM TO EJECTOCMP A PRES	REAL	1.35	bar ←
9WH11T3T	LP STEAM TO EJECTOCMP A TEMP	REAL	126.11	°C ←
9WH12F1I	LP STEAM TO EJECTOCMP B FLOW	REAL	37.58	t/h
9WH12P1T	VAPOUR EJECTOCMP B DIFF PRES	REAL	9.79	mbar
9WH12P3T	LP STEAM TO EJECTOCMP B PRES	REAL	1.33	bar
9WH12T3T	LP STEAM TO EJECTOCMP B TEMP	REAL	126.00	°C
9WJ11F1T	TRISODIUM PHOSPHATE FLOW	REAL	0.39	l/h
9WJ12F1T	SODIUM SULFITE FLOW	REAL	0.02	l/h
9WK01L3T	ANTISCALE TANK 1 LEVEL	REAL	0.44	m
9WK02L3T	ANTISCALE TANK 2 LEVEL	REAL	0.99	m
9WK11F1T	ANTISCALE TO SEA WATER FLOW	REAL	11.01	l/h
9WL01L3T	ANTIFOAM TANK 1 LEVEL	REAL	0.71	m
9WL02L3T	ANTIFOAM TANK 2 LEVEL	REAL	0.41	m
9WL11F1T	ANTIFOAM TO SEA WATER FLOW	REAL	9.94	l/h
9WM00L1I	CONDENSATE STEAM TRFR LVL	REAL	199.71	mm
9WM20E1T	CONDENSATE PP MTR CURRENT	REAL	28.24	A
9WM20F1T	CONDENSATE FLOW	REAL	81.31	t/h
9WM20P3T	CONDENSATE DISCHARGE PRES	REAL	5.61	bar
9WM20T1T	CONDENSATE TEMPERATURE	REAL	96.37	°C
9WM22Q1T	CONDENSATE CONDUCTIVITY	REAL	0.09	µS/cm
9WN00L1I	STM TRFR LVL	REAL	249.69	mm
9WN01T1T	CELL1A CONDENSATE MK UP TEMP	REAL	63.51	°C
9WN02T1T	CELL1B CONDENSATE MK UP TEMP	REAL	63.72	°C
9WN20E1T	COND MK UP PP MTR CURRENT	REAL	25.72	A
9WN20F1T	CONDENSATE MAKE UP FLOW	REAL	81.15	t/h

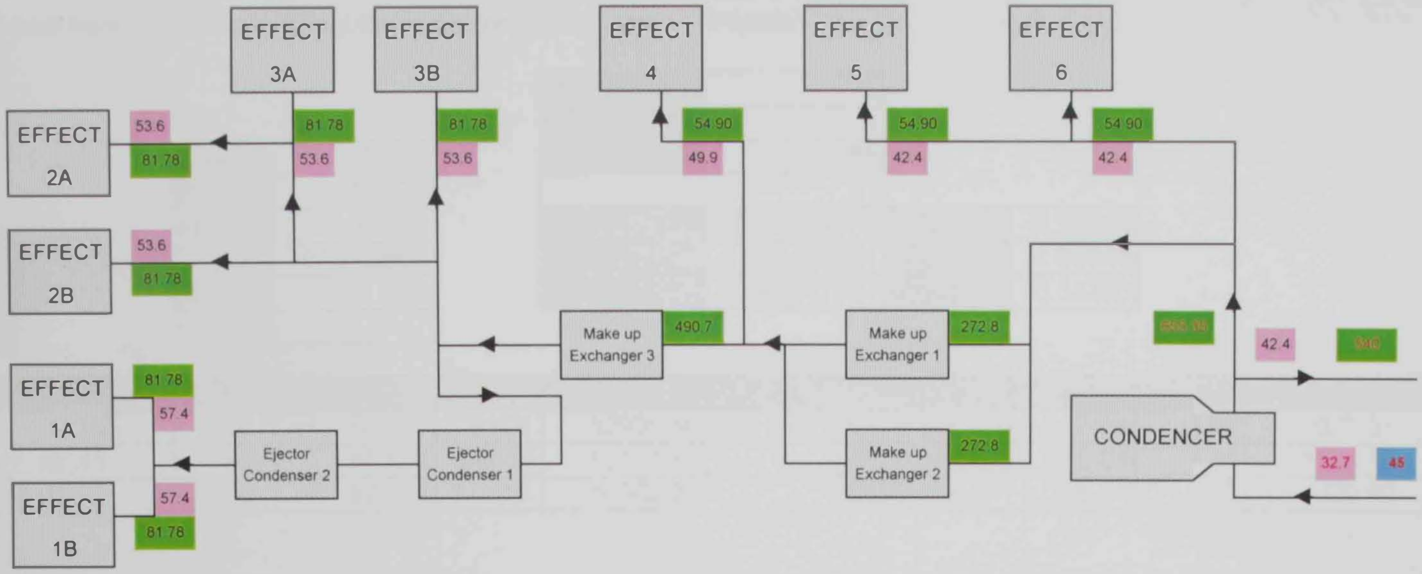
Date : 4-Oct-2000
Time : 10:08:31

REAL TIME VALUE LIST

DESALINATION UNIT 9

TAG	DESCRIPTION	TYPE	VALUE	UNIT
9WN20P3T	CONDENSATE MAKE UP PRES	REAL	4.30	bar
9WN20Q1T	CONDENSATE MK UP CONDUCT	REAL	1.62	µS/cm
9WN20Q2T	CONDENSATE MAKE UP PH	REAL	6.99	pH
9WN20T2T	CONDENSATE MAKE UP TEMP	REAL	63.16	°C
9WN60F1I	STM TRFR RECIRC FLOW	REAL	204.85	t/h
9WN61Q1T	STM TRFR CONDUCT	REAL	232.54	µS/cm
9WN61Q2T	STM TRFR PH	REAL	10.45	pH
9WN80E1T	STM TRFR RECIRC PP MTR CUR	REAL	22.68	A
9WN80F1I	STM TRFR RECIRC PP FLOW	REAL	124.26	t/h
9WN80P3T	STM TRFR RECIRC PRES	REAL	3.10	bar
9WP00L1I	CONDENSER DISTILLATE LEVEL	REAL	227.73	mm
9WP20E1T	DISTILLATE PUMP MTR CURRENT	REAL	147.42	A
9WP20F1T	DISTILLATE FLOW	REAL	674.34 ✓	t/h
9WP20P3T	DISTILLATE PRESSURE	REAL	[redacted]	bar
9WP20T1T	DISTILL PP DRV END BEAR TEMP	REAL	36.82	°C
9WP20T4T	DISTILL PP SHFT END BEAR TEM	REAL	37.68	°C
9WP20T6T	DISTILL BEFORE SW PHTR TEMP	REAL	43.49 ✓	°C
9WP40T7T	DISTILLATE AFTER SW PHT TEMP	REAL	43.58	°C
9WP40Q1T	DISTILLATE CONDUCTIVITY	REAL	94.04	µS/cm
9WQ10F1I	MP STEAM TO EJECTORS FLOW	REAL	3.07	t/h
9WQ10P1T	MP STEAM TO EJECTORS PRES	REAL	11.77	bar
9WQ10T1T	MP STEAM TO EJECTORS TEMP	REAL	189.36	°C
9WW30T1	BOTTOM TEMPERATURE	REAL	55.03	°C
9WZ01T7T	SW MK UP BFR EJEC COND TEMP	REAL	51.52	°C
9WZ01T8T	SW MK UP AFT EJEC COND TEMP	REAL	55.00	°C
9WZ04F1T	SEA WATER TO CELL 4 FLOW	REAL	201.58 ✓	t/h
9WZ05F1T	SEA WATER TO CELL 5 FLOW	REAL	186.00 ✓	t/h
9WZ06F1T	SEA WATER TO CELL 6 FLOW	REAL	217.74 ✓	t/h
9WZ11F1T	SEA WATER TO CELL 1A FLOW	REAL	289.80 ✓	t/h
9WZ12F1T	SEA WATER TO CELL 1B FLOW	REAL	275.00 ✓	t/h
9WZ21F1T	SEA WATER TO CELL 2A FLOW	REAL	303.36 ✓	t/h
9WZ22F1T	SEA WATER TO CELL 2B FLOW	REAL	298.87 ✓	t/h
9WZ30T1T	SW MK UP BTW EXCHANGER TEMP	REAL	47.33	°C
9WZ31F1T	SEA WATER TO CELL 3A FLOW	REAL	312.81 ✓	t/h
9WZ32F1T	SEA WATER TO CELL 3B FLOW	REAL	298.43 ✓	t/h
9WZ40E1T	SW MK UP CIRC MTR CURRENT	REAL	29.71	A
9WZ40F1I	SEA WATER MAKE UP FLOW	REAL	2359.25 ✓	t/h
9WZ40T4T	SWMK UP CIR MTR SHT BRG TEMP	REAL	43.64	°C
9WZ40T6T	SWMK UP CIR MTR DRV BRG TEMP	REAL	52.53	°C
9WZ40T7T	SW MK UP CIRC DRV BEAR TEMP	REAL	44.58	°C
9WZ40T10T	SW MK UP CIRC MTR WDING TEMP	REAL	69.96	°C
9WZ40T11T	SW MK UP CIRC MTR WDING TEMP	REAL	70 37	°C
9WZ40T12T	SW MK UP CIRC MTR WDING TEMP	REAL	69.76	°C
9WZ40T14T	SW MK UP CIRC SHFT BEAR TEMP	REAL	45 76	°C


MEDNAR
SIMULATOR
UANPC MED-TVC UNITS
Ts 63.67
Tc3 53.62
Tv6 43.49
Pms 236.2
SIMULATE
100%
50%
PR 8.89
sMcw 2.90
sA 313.5
Mf 655.4
Mms 10.48
Xsw 45
Tsw 32.7
AHOT 6724
ACOLD 4453
ΔTHOT 0.415
ΔTCOLD 0.68
AF.COND 2874
NCG 0.02
T °C
A m2
M kg/s
U kW/m2.°C
X *1000 ppm
UHOT 3.691
Mcw 540
UCOLD 3.515
Tv4 50.39
REST
Input Data
Guess Value
Goal 0.00
Diff 0.00
WELCOME TO MEDNAR SIMULATOR



Temperature *C*	
Pressure *Kpa*	
Flow *Kg/s*	
Salinity **1000 ppm*	

Ejec. Condenser

			2823.237993
QEC	2475.24		888.0767191
	stream	Flow	Temp
	Mf(inlet)	163.55	53.60
	Mf(outlet)	163.55	57.42

**Mass & Energy balance for Pre-Heaters

Pre-heater 3

Area m^2	693		Q3	7111.24
U_{HE3}	3.445219883			
5.143560194	5.143878096			
Stream	Flow	Temp	Q3 in	7157.41
Mf2(inlet)	490.66	49.95		
Mf2(outlet)	490.66	53.60	Q3 out	7157.41
Md3"	3.0125	54.30		

Pre-heater 12

Area m^2	1156		Q12	14651.7
U_{HE12}	3.319938448			
4.866435078	4.865222299			
Stream	Flow	Temp	Q12 in	16388.5
Mf1(inlet)	545.56	42.41		
Mf1(outlet)	545.56	49.95	Q12 out	16388.5
Md4"	6.8784	50.39		

F. Condenser

Area m^3	2874		QFC	45882.19
U_{FC}	3.099996413			
3.760151657	3.779248493			
Stream	Flow	Temp	QFC in	46121.93
Mf (inlet)	1195.3	32.70		
Mf (outlet)	1195.3	42.41	QFC out	0
Mdc		43.49		

** Local heat transfer coefficient for seawater flowing inside the tubes

dout	0.019
dinner	0.0176
salinity	45000
n12	1070.66667
n3	963
nfc	3147

Ri	0
Ro	0
K	0.0165

1/UFC	0.26594672	UFC	3.76015
1/UHE12	0.205489231	UHE12	4.86644
1/UHE3	0.194417867	UHE3	5.14356

temperature oC	hout	Ditus-Bolter	velocity	Reynolds	Prandtl	Kwater	muwater	cpwater	rhowater
32.70	5.32	581.81	1.5189	32936.80	0.01	6.18E-01	8.34E-04	3.95	1027.91
42.41	7.55	812.54	2.0452	53127.67	0.00	6.30E-01	6.94E-04	3.96	1024.10
49.95	8.16	871.12	2.0516	60502.30	0.00	6.38E-01	6.09E-04	3.96	1020.80

** Local heat transfer coefficient on the outside surface of tubes

dout	0.019	xncondout	0

con1	1.528671	con2	1
num1	0.824968	dinner	0.0176
mumt	2.139519		

temperature oC	hout	delt	rhovapor	rhocond	kliquid	lamdav	muliquid	mass	velocityout
43.49	52.48	0.1	0.061	990.469	6.34E-04	2398.15	6.14E-04	19.13	37.11217912
50.39	54.38	0.1	0.085	987.509	6.41E-04	2381.59	5.45E-04	6.88	28.48938017
54.30	55.40	0.1	0.101	985.707	6.45E-04	2372.15	5.11E-04	3.01	12.23813533

** STREAMS DATA & RESULTS

		STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B.	E.B.
Cell #1		Cell #1		60.82				254.42	2,611.38		
	In	Ms	23.60	63.67					2,616.23	105.38	80377.48
		Mf01	81.78	57.42	45.00		3.97				
	Out	Mb1	59.26	61.48	62.10	0.78	3.89			105.38	79268.21
		Mv1	22.52	60.82					2,611.38		
		Md1	13.12	63.67				266.37			
		Mcon	10.48	63.67				266.37			
										Err %	1.38

U	3.690603926
A	6724
ΔT loss Hot	0.415

Q	T1	delta T	Qcal	Err
54354.145	61.48	2.19	54354.14495	0

NCG	0.02

** BOILING POINT ELEVATION

STREAM #	X	T (K)	BPE	Tv1 (°C)	Hv1
Mb1(assume)1	0.062098964	334.63	0.7761731		
LOOP 1	0.0475	334.63	0.5744381	60.91	2611.520386
LOOP 2	0.053505331	334.63	0.6556146	60.82	2611.381556
LOOP 3	0.053549163	334.63	0.6562161	60.82	2611.380527
LOOP 4	0.05354948	334.63	0.6562205	60.82	2611.38052
LOOP 5	0.053549482	334.63	0.6562205	60.82	2611.38052

** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T	A	B	C	D	Cp
Mf01	45.00	57.42	3933.7967	0.85175	-0.008250235	5.99885E-05	3.966859977
LOOP 1	50	61.48	3906.535	1.014725	-0.0100305	6.54708E-05	3.946221135
LOOP 2	62.01066116	61.48	3843.5593	1.359795	-0.013920665	7.77355E-05	3.892606456
LOOP 3	62.09832664	61.48	3843.1127	1.362073	-0.013947054	7.78203E-05	3.892219831
LOOP 4	62.09895914	61.48	3843.1095	1.362089	-0.013947244	7.78209E-05	3.892217041
LOOP 5	62.0989637	61.48	3843.1095	1.362089	-0.013947245	7.78209E-05	3.892217021

Cp= (A+BT+CT^2+DT^3)E-3

WHERE

T= TEMPERATURE C
S= WATER SALINITY G/Kg
A= 4206.8-6.6197S+1.2288E-2S^2
B= -1.1262+5.4178E-2S-2.2719E-4S^2'
C= 1.2026E-2-5.3566E-4S+1.8906E-6S^2
D= 6.87774E-7+1.517E-6S-4.4268E-9S^2

**SALINITY

STREAM #	Assumed X	Mb1	Mv1	Cp	Calc X
LOOP 1	50	73.59888	8.177653	3.946221	62.01066116
LOOP 2	62.01066116	59.34373	22.432801	3.892606	62.09832664
LOOP 3	62.09832664	59.25995	22.516578	3.89222	62.09895914
LOOP 4	62.09895914	59.25935	22.517182	3.892217	62.0989637
LOOP 5	62.0989637	59.25934	22.517186	3.892217	62.09896373

** STREAMS DATA & RESULTS

		STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B	E.B
Cell #2		Cell #2		57.54				240.66	2,605.75		
	In	Mf02	81.78	53.60	45.00		3.96			176.68	93,857.39
		Md1	13.12	63.67				266.37			
		Mb1	59.26	61.48	62.10	0.78	3.89				
		Mv1	22.52	60.82				254.42	2,611.38		
	Internal	Md2*	22.52	60.41				252.68	2,610.67		
		Mb2*	59.94	58.27	61.39	0.73	3.89				
		Mv2*	21.83	57.54					2,605.75		
		Ld2	12.98	57.54				240.66			
		Lb2	58.95	58.27	62.42	0.76	3.89				
		Vd2'	0.14	57.54					2,605.75		
		Vb2'	0.31	57.54					2,605.75		
	Out	Md2	35.50	59.36				248.28		176.68	93,841.68
		Mb2	118.89	58.27	61.90	0.76	3.89				
		Mv2	22.29	57.54				240.66	2,605.75		
										Err %	0.02

U	3.690603926
A	6724
dlta loss	0.415

Q	T2	delta T	Qcal	Err
53095.287	58.27	2.14	53095.28702	5.82E-11

** BOILING POINT ELEVATION

STREAM #	X	T2 (K)	BPE	Tv2 (°C)	Hv2
Mb2	0.061903308	331.42	0.7564739		
Lb2	0.062424723	331.42	0.7637908		
Mb2' (LOOP 1)	0.054134434	331.42	0.6497556	57.62	2605.879876
Mb2' (LOOP 2)	0.059790794	331.42	0.7270309	57.54	2605.74673
Mb2' (LOOP 3)	0.059829438	331.42	0.7275666	57.54	2605.745807
Mb2' (LOOP 4)	0.059829699	331.42	0.7275703	57.54	2605.745801
Mb2' (LOOP 5)	0.0598297	331.42	0.7275703	57.54	2605.745801

** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T2	A	B	C	D	Cp
Mf02	45.00	53.60	3933.7967	0.85175	-0.008250235	5.99885E-05	3.964987
Mb2	61.90	58.27	3844.1065	1.357001	-0.01388831	7.76315E-05	3.891381
Lb2	62.42	58.27	3841.4515	1.370522	-0.01404505	7.81355E-05	3.889082
Mb2' (LOOP 1)	50	58.27	3906.535	1.014725	-0.0100305	6.54708E-05	3.944558
Mb2' (LOOP 2)	61.31271922	58.27	3847.1219	1.341537	-0.013709534	7.70577E-05	3.893989
Mb2' (LOOP 3)	61.39000674	58.27	3846.7268	1.343569	-0.013733005	7.7133E-05	3.893648
Mb2' (LOOP 4)	61.39052885	58.27	3846.7241	1.343583	-0.013733163	7.71335E-05	3.893645
Mb2' (LOOP 5)	61.39053237	58.27	3846.7241	1.343583	-0.013733164	7.71335E-05	3.893645

Cp= (A+BT+CT^2+DT^3)E-3

WHERE
T= TEMPERATURE C
S= WATER SALINITY G/Kg
A= 4206.8-6.6197S+1.2288E-2S^2
B= -1.1262+5.4178E-2S-2.2719E-4S^2'
C= 1.2026E-2-5.3566E-4S+1.8906E-6S^2
D= 6.87774E-7+1.517E-6S-4.4268E-9S^2

**MODUL

STREAM #	Assumed X2'	Mb2'	Mv2'	Cp	Calc X2'
Mb2' (LOOP 1)	50	73.59888	8.18	3.944558	61.31271922
Mb2' (LOOP 2)	61.31271922	60.01926	21.76	3.893989	61.39000674
Mb2' (LOOP 3)	61.39000674	59.9437	21.83	3.893648	61.39052885
Mb2' (LOOP 4)	61.39052885	59.94319	21.83	3.893645	61.39053237
Mb2' (LOOP 5)	61.39053237	59.94318	21.83	3.893645	61.3905324

** STREAMS DATA & RESULTS

	STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B	E.B.
Cell #3	Cell #3		53.62				224.22	2,598.96		
In	Mf03	81.78	53.60	45.00		3.96			258.45	111,222.36
In	Md2	35.50	59.36				248.28			
In	Mb2	118.89	58.27	61.90	0.76	3.89				
In	Mv2	22.29	57.54				240.66	2,605.75		
Internal	Md3'	22.29	57.13				238.92	2,605.03		
Internal	Mb3'	59.72	55.00	61.62	0.69	3.89				
Internal	Mv3'	22.05	54.31					2,600.16		
Internal	Ld3	35.14	53.62				224.22			
Internal	Lb3	118.26	55.00	62.24	0.74	3.89				
Internal	Vd3	0.36	53.62					2,598.96		
Internal	Vb3	0.63	54.31					2,600.16		
Out	Md3*	57.42	54.95				229.78		258.45	111,195.96
Out	Mb3	177.98	55.00	62.03	0.74	3.89				
Out	Mv3*	9.93	54.30					2,600.14		
Out	Mev (tvc)in	13.12	54.30					2,600.14		
									Err %	0.02

U	3.690603926
A	6724
dlta loss	0.415

Q	T3	delta T	Qcal	Err
52729.382	55.00	2.12	52729.38178	-5.82E-11

Md3"
3.012460294

** BOILING POINT ELEVATION

STREAM #	X	T3 (K)	BPE	Tv3 (°C)	Hv3
Mb3	0.06202832	328.15	0.7411561		
Lb3	0.062235615	328.15	0.7439988		
Mb3' (LOOP 1)	0.052500726	328.15	0.6137275	54.39	2600.292355
Mb3' (LOOP 2)	0.058272095	328.15	0.690172	54.31	2600.15968
Mb3' (LOOP 3)	0.058309409	328.15	0.6906736	54.31	2600.158809
Mb3' (LOOP 4)	0.058309648	328.15	0.6906768	54.31	2600.158803
Mb3' (LOOP 5)	0.058309649	328.15	0.6906768	54.31	2600.158803

** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T3	A	B	C	D	Cp
Mf03	45.00	53.60	3933.7967	0.85175	-0.008250235	5.99885E-05	3.964987
Mb3	62.03	55.00	3843.4694	1.360254	-0.013925983	7.77526E-05	3.889094
Lb3	62.24	55.00	3842.4137	1.365633	-0.013988322	7.7953E-05	3.888179
Mb3' (LOOP 1)	50	55.00	3906.535	1.014725	-0.0100305	6.54708E-05	3.942896
Mb3' (LOOP 2)	61.54273898	55.00	3845.9464	1.347578	-0.01377932	7.72816E-05	3.891239
Mb3' (LOOP 3)	61.61736688	55.00	3845.5654	1.349533	-0.013801918	7.73541E-05	3.890909
Mb3' (LOOP 4)	61.61784366	55.00	3845.5629	1.349546	-0.013802062	7.73546E-05	3.890907
Mb3' (LOOP 5)	61.6178467	55.00	3845.5629	1.349546	-0.013802063	7.73546E-05	3.890907

Cp= (A+BT+CT^2+DT^3)E-3

WHERE

T=	TEMPERATURE C
S=	WATER SALINITY G/Kg
A=	4206.8-6.6197S+1.2288E-2S^2
B=	-1.1262+5.4178E-2S-2.2719E-4S^2'
C=	1.2026E-2-5.3566E-4S+1.8906E-6S^2
D=	6.87774E-7+1.517E-6S-4.4268E-9S^2

**MODUL

STREAM #	Assumed X3'	Mb3'	Mv3'	Cp	Calc X3'
Mb3' (LOOP 1)	50	73.59888	8.18	3.942896	61.54273898
Mb3' (LOOP 2)	61.54273898	59.79493	21.98	3.891239	61.61736688
Mb3' (LOOP 3)	61.61736688	59.72251	22.05	3.890909	61.61784366
Mb3' (LOOP 4)	61.61784366	59.72205	22.05	3.890907	61.6178467
Mb3' (LOOP 5)	61.6178467	59.72205	22.05	3.890907	61.61784672

Steam Jet Ejector Formula: A/B

Ra	Mms	Mev	Ps	Pv	Pms	Tv	PCF	TCF	Ms
0.798611	10.48	13.12278	23.56867	15.09475	236.1845	54.30	1.414269	0.972179	23.60278

** STREAMS DATA & RESULTS

		STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B.	E.B.
Cell #4		Cell #4		50.40				210.73	2,593.34		
	In	Mf04	54.90	49.95	45.00		3.96			545.56	157,876.10
		Md3	117.86	54.95				229.78			
		Mb3	355.96	55.00	62.03	0.74	3.89				
		Mv3	16.84	54.30				227.07	2,600.14		
	Internal	Md4'	16.84	53.62				224.22	2,598.96		
		Mb4'	38.19	51.07	64.68	0.67	3.88				
		Mv4'	16.70	50.40					2,593.34		
		Ld4	116.92	50.40				210.73			
		Lb4	353.58	51.07	62.45	0.73	3.89				
		Vd4	0.94	50.40					2,593.34		
		Vb4	2.38	50.40					2,593.34		
	Out	Md4*	133.76	50.88				212.75		545.56	158,097.02
		Mb4	391.77	51.07	62.66	0.73	3.88				
		Mv4*	20.03	50.40					2,593.34		
										Err %	-0.14

U	3.515141263
A	4453
ΔT loss Cold	0.68

Q	T4	delta T	Qcal	Err
39989.568	51.07	2.55	39989.56818	0

Md4"
6.878393374

** BOILING POINT ELEVATION

STREAM #	X	T4 (K)	BPE	Tv4 (°C)	Hv3
Mb4	0.062663648	324.22	0.7292878		
Lb4	0.062445786	324.22	0.7263765		
Mb4' (LOOP 1)	0.050532617	324.22	0.5720495	50.49	2593.504165
Mb4' (LOOP 2)	0.057826667	324.22	0.6654189	50.40	2593.340697
Mb4' (LOOP 3)	0.057872603	324.22	0.666018	50.40	2593.339648
Mb4' (LOOP 4)	0.057872887	324.22	0.6660217	50.40	2593.339642
Mb4' (LOOP 5)	0.057872889	324.22	0.6660217	50.40	2593.339642

1.291048

** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T4	A	B	C	D	Cp
Mf04	45.00	49.95	3933.7967	0.85175	-0.008250235	5.99885E-05	3.963232
Mb4	62.66	51.07	3840.2371	1.376677	-0.014116529	7.83657E-05	3.884162
Lb4	62.45	51.07	3841.3444	1.371066	-0.01405136	7.81558E-05	3.885124
Mb4' (LOOP 1)	50	51.07	3906.535	1.014725	-0.0100305	6.54708E-05	3.940914
Mb4' (LOOP 2)	64.58809995	51.07	3830.5071	1.425303	-0.014684392	8.0201E-05	3.875678
Mb4' (LOOP 3)	64.6799708	51.07	3830.0448	1.427582	-0.014711151	8.02878E-05	3.875274
Mb4' (LOOP 4)	64.68054031	51.07	3830.042	1.427597	-0.014711316	8.02883E-05	3.875272
Mb4' (LOOP 5)	64.68054384	51.07	3830.0419	1.427597	-0.014711317	8.02883E-05	3.875272

Cp= (A+BT+CT^2+DT^3)E-3

WHERE
T= TEMPERATURE C
S= WATER SALINITY G/Kg
A= 4206.8-6.6197S+1.2288E-2S^2
B= -1.1262+5.4178E-2S-2.2719E-4S^2'
C= 1.2026E-2-5.3566E-4S+1.8906E-6S^2
D= 6.87774E-7+1.517E-6S-4.4268E-9S^2

**MODUL

STREAM #	Assumed X4'	Mb4'	Mv4'	Cp	Calc X4'
Mb4' (LOOP 1)	50	49.40725	5.49	3.940914	64.58809995
Mb4' (LOOP 2)	64.58809995	38.24795	16.65	3.875678	64.6799708
Mb4' (LOOP 3)	64.6799708	38.19362	16.70	3.875274	64.68054031
Mb4' (LOOP 4)	64.68054031	38.19328	16.70	3.875272	64.68054384
Mb4' (LOOP 5)	64.68054384	38.19328	16.70	3.875272	64.68054386

** STREAMS DATA & RESULTS

		STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B.	E.B.
Cell #5		Cell #5		47.13				197.03	2,587.59		
	In	Mf05	54.90	42.41	45.00		3.96				
		Md4	140.63	50.88				212.75		600.45	150,940.23
		Mb4	391.77	51.07	62.66	0.73	3.88				
		Mv4	13.15	50.40				210.73	2,593.34		
	Internal	Md5'	13.15	49.72				207.88	2,592.15		
		Mb5'	42.25	47.72	58.46	0.59	3.90				
		Mv5'	12.64	47.13					2,587.59		
		Ld5	139.71	47.13				197.03			
		Lb5	389.78	47.72	62.98	0.72	3.88				
		Vd5	0.92	47.13					2,587.59		
		Vb5	1.99	47.13					2,587.59		
	Out	Md5	152.86	47.48				198.53			
		Mb5	432.04	47.72	62.54	0.71	3.88			600.45	150,651.99
		Mv5	15.56	47.13					2,587.59		
										Err %	0.19

U	3.515141263
A	4453
dlta loss	0.68

Q	T5	delta T	Qcal	Err
31346.675	47.72	2.00	31346.67471	0

** BOILING POINT ELEVATION

STREAM #	X	T5 (K)	BPE	Tv5 (°C)	Hv3
Mb5	0.062541751	320.87	0.7103659		
Lb5	0.062983719	320.87	0.7161361		
Mb5' (LOOP 1)	0.048858302	320.87	0.5380435	47.18	2587.680282
Mb5' (LOOP 2)	0.053068103	320.87	0.5897899	47.13	2587.589023
Mb5' (LOOP 3)	0.053090529	320.87	0.5900685	47.13	2587.588532
Mb5' (LOOP 4)	0.053090647	320.87	0.59007	47.13	2587.588529
Mb5' (LOOP 5)	0.053090648	320.87	0.59007	47.13	2587.588529

** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T5	A	B	C	D	Cp
Mf05	45.00	42.41	3933.7967	0.85175	-0.008250235	5.99885E-05	3.959654
Mb5	62.54	47.72	3840.8565	1.37354	-0.014080088	7.82483E-05	3.88284
Lb5	62.98	47.72	3838.6125	1.384881	-0.014211945	7.86732E-05	3.880883
Mb5' (LOOP 1)	50	47.72	3906.535	1.014725	-0.0100305	6.54708E-05	3.939229
Mb5' (LOOP 2)	58.41960193	47.72	3862.0169	1.263492	-0.01281471	7.42023E-05	3.901191
Mb5' (LOOP 3)	58.46445341	47.72	3861.7844	1.264731	-0.012828824	7.42471E-05	3.90099
Mb5' (LOOP 4)	58.46469006	47.72	3861.7831	1.264737	-0.012828898	7.42474E-05	3.900989
Mb5' (LOOP 5)	58.46469131	47.72	3861.7831	1.264737	-0.012828899	7.42474E-05	3.900989

Cp= (A+BT+CT^2+DT^3)E-3

WHERE

T=	TEMPERATURE C
S=	WATER SALINITY G/Kg
A=	4206.8-6.6197S+1.2288E-2S^2
B=	-1.1262+5.4178E-2S-2.2719E-4S^2'
C=	1.2026E-2-5.3566E-4S+1.8906E-6S^2
D=	6.87774E-7+1.517E-6S-4.4268E-9S^2

**MODUL

STREAM #	Assumed X5'	Mb5'	Mv5'	Cp	Calc X5'
Mb5' (LOOP 1)	50	49.40725	5.49	3.939229	58.41960193
Mb5' (LOOP 2)	58.41960193	42.28653	12.61	3.901191	58.46445341
Mb5' (LOOP 3)	58.46445341	42.25409	12.64	3.90099	58.46469006
Mb5' (LOOP 4)	58.46469006	42.25392	12.64	3.900989	58.46469131
Mb5' (LOOP 5)	58.46469131	42.25392	12.64	3.900989	58.46469131

** STREAMS DATA & RESULTS

		STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B.	E.B.
Cell #6		Cell #6		43.49				181.82	2,581.15		
	In	Mf06	54.90	42.41	45.00		3.96				
		Md5	152.86	47.48				198.53		655.35	159,869.81
		Mb5	432.04	47.72	62.54	0.71	3.88				
		Mv5	15.56	47.13				197.03	2,587.59		
	Internal	Md6'	15.56	46.45				194.18	2,586.39		
		Mb6'	39.52	44.07	62.51	0.58	3.88				
		Mv6'	15.38	43.49					2,581.15		
		Ld6	151.79	43.49				181.82			
		Lb6	429.48	44.07	62.91	0.70	3.88				
		Vd6	1.06	43.49					2,581.15		
		Vb6	2.55	43.49					2,581.15		
	Out	Md6	167.35	43.95				183.75			
		Mb6	469.00	44.07	62.88	0.70	3.88			655.35	159,966.52
		Mv6	19.00	43.49					2,581.15		
										Err %	-0.06

U	3.515141263
A	4453
dlta loss	0.68

Q	T6	delta T	Qcal	Err
37219.833	44.07	2.38	37219.83262	0

** BOILING POINT ELEVATION

STREAM #	X	T6 (K)	BPE	Tv6 (°C)	Hv3
Mb6	0.062880074	317.22	0.6960166		
Lb6	0.062913766	317.22	0.6964451		
Mb6' (LOOP 1)	0.047034357	317.22	0.5024796	43.57	2581.284651
Mb6' (LOOP 2)	0.053258344	317.22	0.5766531	43.49	2581.152795
Mb6' (LOOP 3)	0.053291138	317.22	0.5770501	43.49	2581.152089
Mb6' (LOOP 4)	0.053291308	317.22	0.5770522	43.49	2581.152085
Mb6' (LOOP 5)	0.053291309	317.22	0.5770522	43.49	2581.152085

** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T6	A	B	C	D	Cp
Mf06	45.00	42.41	3933.7967	0.85175	-0.008250235	5.99885E-05	3.959654
Mb6	62.88	44.07	3839.1383	1.382229	-0.01418109	7.85737E-05	3.879236
Lb6	62.91	44.07	3838.9674	1.383092	-0.014191125	7.86061E-05	3.879086
Mb6' (LOOP 1)	50	44.07	3906.535	1.014725	-0.0100305	6.54708E-05	3.937376
Mb6' (LOOP 2)	62.44797236	44.07	3841.3333	1.371122	-0.014052015	7.81579E-05	3.881156
Mb6' (LOOP 3)	62.51356086	44.07	3840.9998	1.372814	-0.014071653	7.82212E-05	3.880865
Mb6' (LOOP 4)	62.51390136	44.07	3840.9981	1.372822	-0.014071755	7.82215E-05	3.880863
Mb6' (LOOP 5)	62.51390313	44.07	3840.9981	1.372822	-0.014071755	7.82215E-05	3.880863

Cp= (A+BT+CT^2+DT^3)E-3

WHERE

T= TEMPERATURE C

S= WATER SALINITY G/Kg

A= 4206.8-6.6197S+1.2288E-2S^2

B= -1.1262+5.4178E-2S-2.2719E-4S^2'

C= 1.2026E-2-5.3566E-4S+1.8906E-6S^2

D= 6.87774E-7+1.517E-6S-4.4268E-9S^2

**MODUL

STREAM #	Assumed X6'	Mb6'	Mv6'	Cp	Calc X6'
Mb6' (LOOP 1)	50	49.40725	5.49	3.937376	62.44797236
Mb6' (LOOP 2)	62.44797236	39.55873	15.34	3.881156	62.51356086
Mb6' (LOOP 3)	62.51356086	39.51722	15.38	3.880865	62.51390136
Mb6' (LOOP 4)	62.51390136	39.51701	15.38	3.880863	62.51390313
Mb6' (LOOP 5)	62.51390313	39.51701	15.38	3.880863	62.51390314

** STREAMS DATA & RESULTS

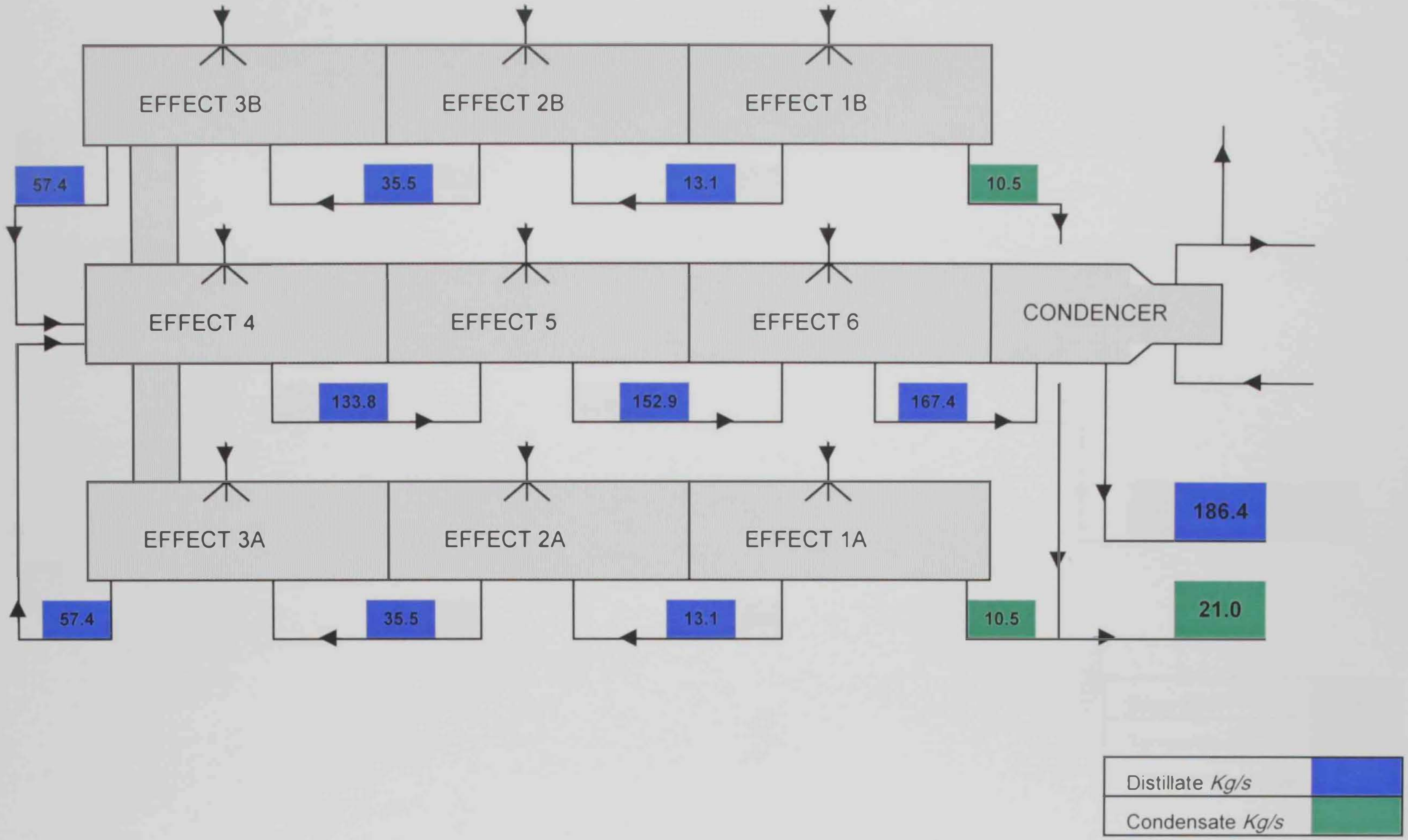
		STREAM #	FLOW	T	SALINITY	BPE	CP	h	H	M.B.	E.B.
Condenser		F.Condenser		43.49				181.82	2,581.15		
	In	Mf+Mcw	1,195.32	32.70	45.00		3.95			1,381.67	234375.7
		Md6	167.35	43.95				183.75			
		Mv6	19.00	43.49					2,581.15		
	Internal	Vdc'	0.13	43.49					2,581.15		
		Ldc	167.22	43.49				181.82			
		Mdc	19.13	43.49				181.82			
	Out	Mcw	539.97	42.41	45.00		3.96			1,381.67	234591.1
		Mf	655.35	42.41	45.00		3.96				
		Distillate	186.35	43.49				181.82	2,581.15		
										Err %	-0.09

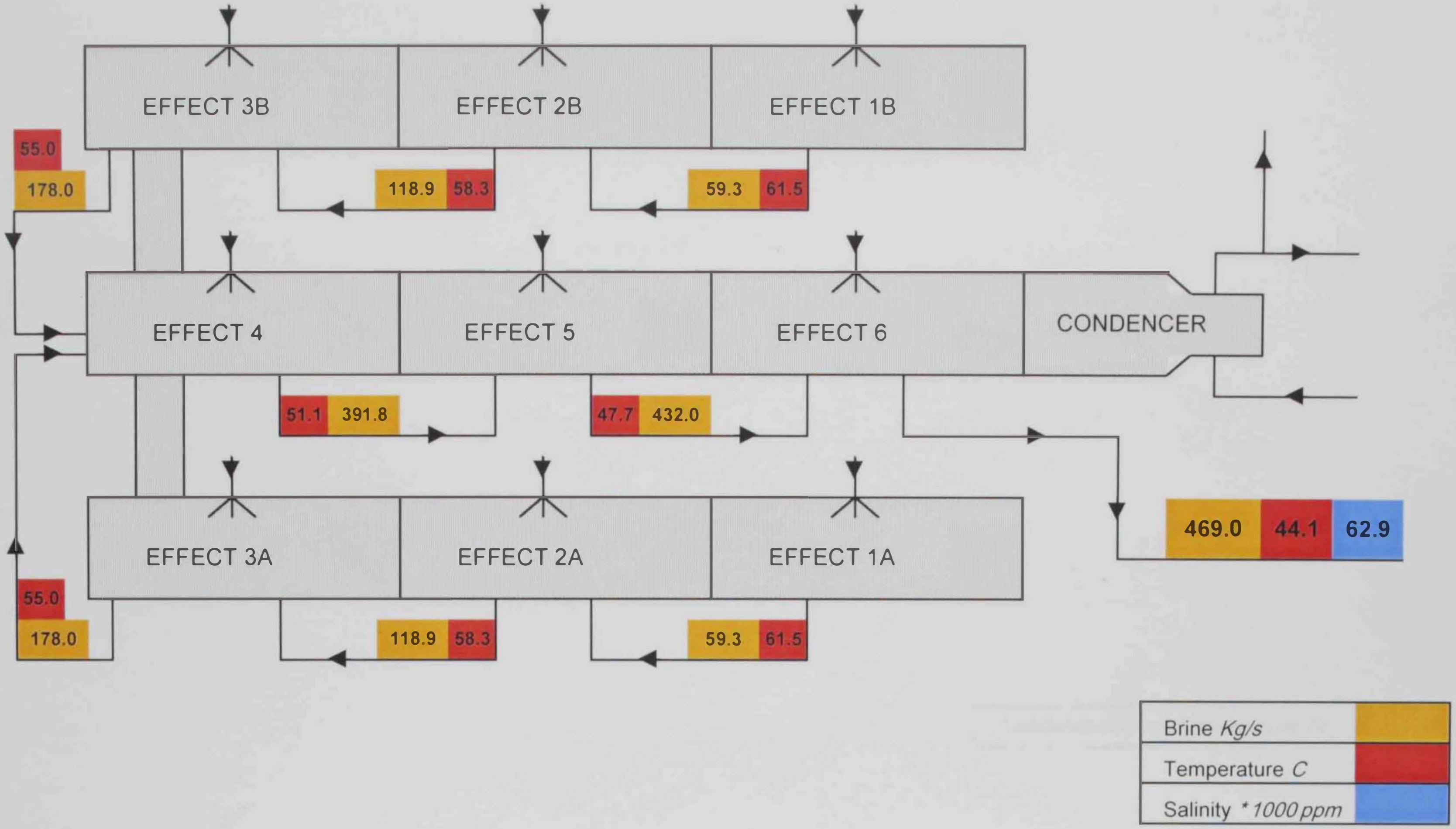
U	3.779248493
A	2874

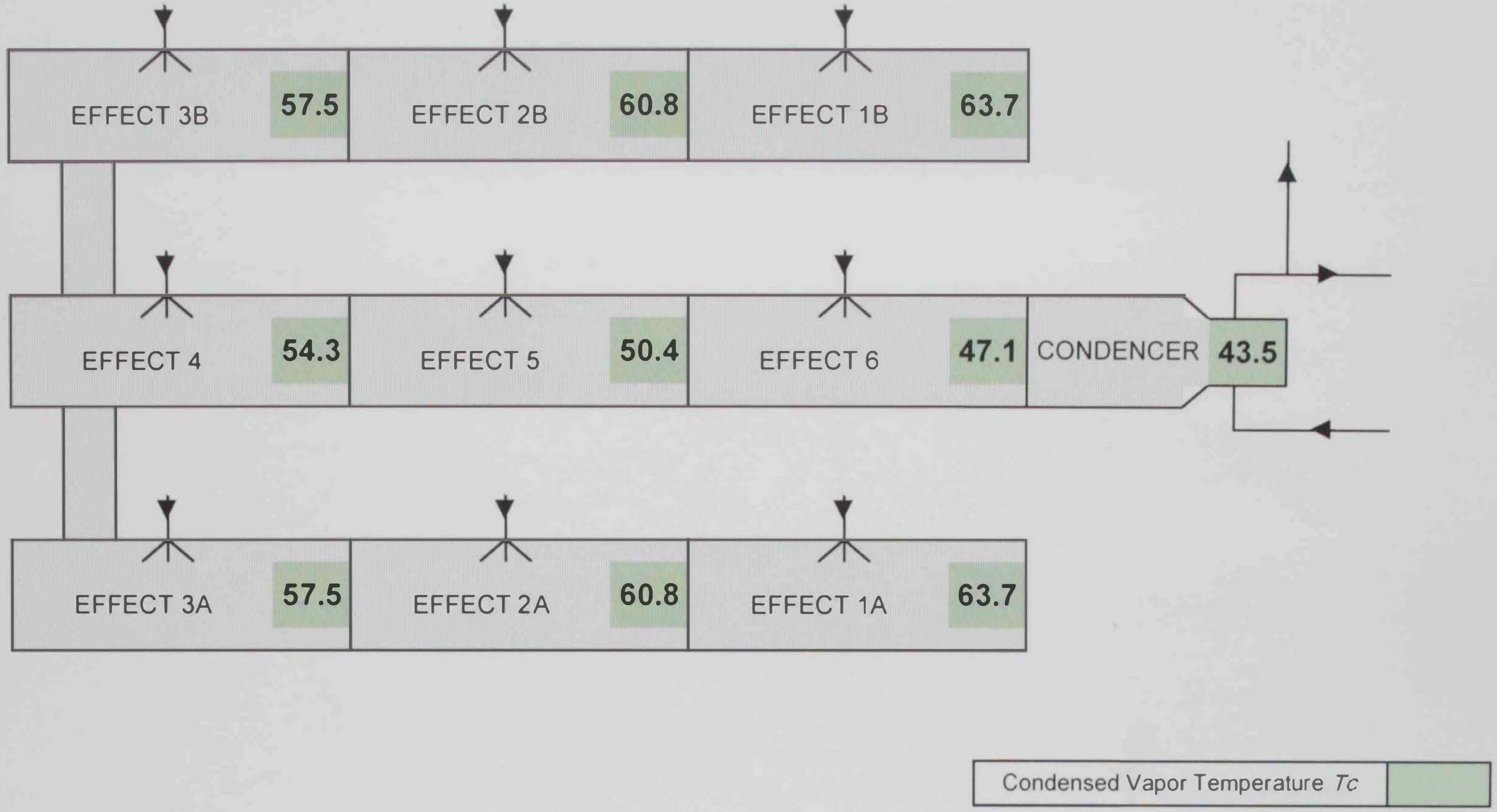
Q	Tvc	delta T lm	Qcal	Err
45906.995	43.49	4.23	45909.7279	-2.733122614

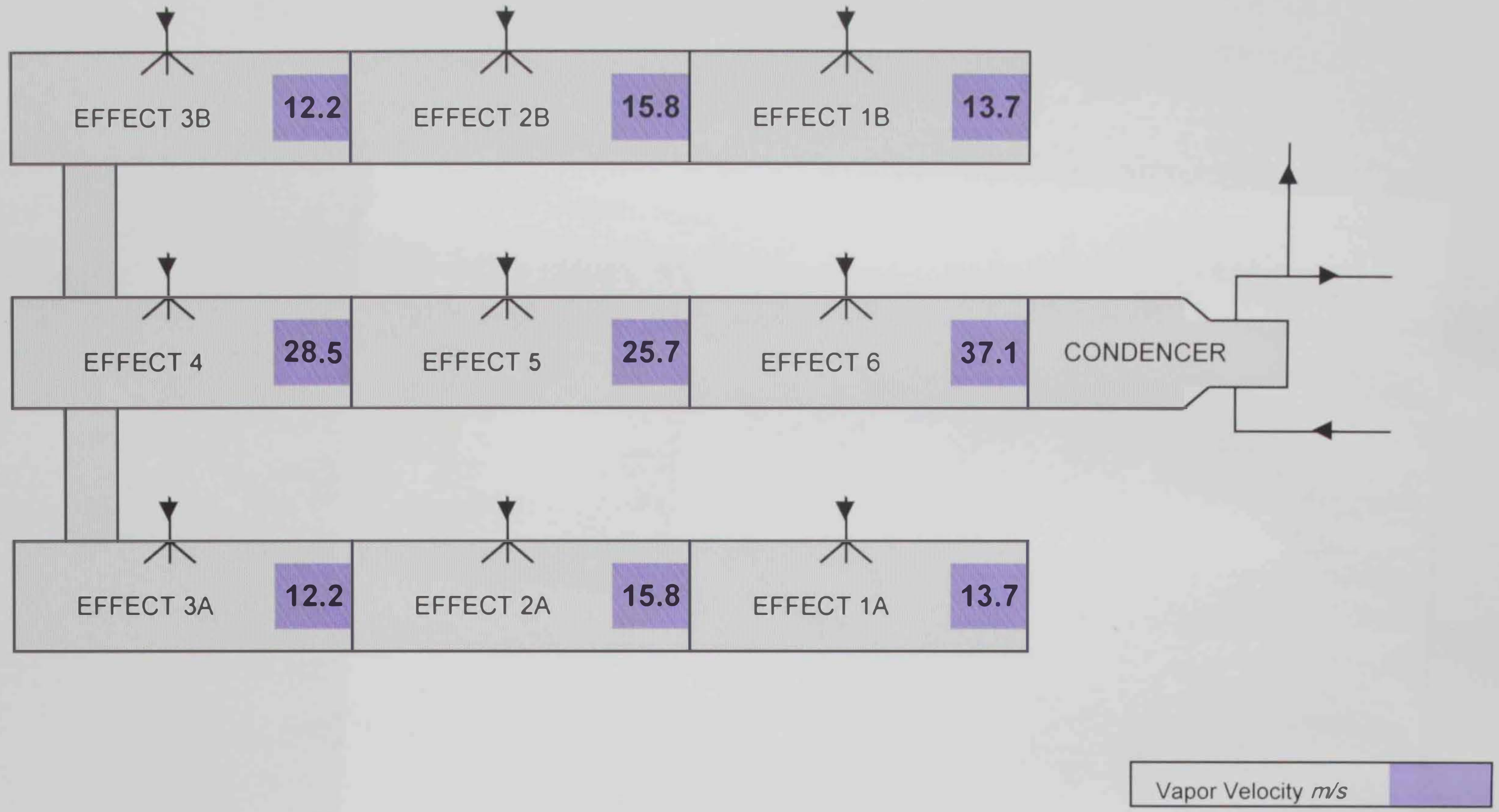
** SPECIFIC HEAT OF WATER AT CONSTANT PRESSURE

STREAM #	X	T	A	B	C	D	Cp
Mf+Mcw in	45.00	32.70	3933.7967	0.85175	-0.008250235	5.99885E-05	3.954924584
Mf & Mcw out	45.00	42.41	3933.7967	0.85175	-0.008250235	5.99885E-05	3.959654247


EFFECT 3B
EFFECT 2B
EFFECT 1B
57.4
35.5
13.1
10.5
EFFECT 4
EFFECT 5
EFFECT 6
CONDENCER
133.8
152.9
167.4
186.4
EFFECT 3A
EFFECT 2A
EFFECT 1A
21.0
57.4
35.5
13.1
10.5
Distillate Kg/s
Condensate Kg/s


EFFECT 3B
EFFECT 2B
EFFECT 1B
55.0
178.0
118.9
58.3
59.3
61.5
EFFECT 4
EFFECT 5
EFFECT 6
CONDENCER
51.1
391.8
47.7
432.0
469.0
44.1
62.9
EFFECT 3A
EFFECT 2A
EFFECT 1A
55.0
178.0
118.9
58.3
59.3
61.5
Brine Kg/s
Temperature C
Salinity *1000ppm


EFFECT 3B
57.5
EFFECT 2B
60.8
EFFECT 1B
63.7
EFFECT 4
54.3
EFFECT 5
50.4
EFFECT 6
47.1
CONDENCER
43.5
EFFECT 3A
57.5
EFFECT 2A
60.8
EFFECT 1A
63.7
Condensed Vapor Temperature *Tc*


EFFECT 3B
12.2
EFFECT 2B
15.8
EFFECT 1B
13.7
EFFECT 4
28.5
EFFECT 5
25.7
EFFECT 6
37.1
CONDENCER
EFFECT 3A
12.2
EFFECT 2A
15.8
EFFECT 1A
13.7
Vapor Velocity m/s