

A THERMODYNAMIC APPROACH TO THE SYNTHESIS OF THERMAL DESALINATION PROCESSES

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Keywords : Cogeneration, Combined Cycle, Cold composite stream, Cycle alone, Dual Purpose Plant, E-C Subsystem, Hot composite stream, Power Generation Cycles

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Summary

The basic principles for thermodynamic synthesis are presented covering the principal reported works on the subject.

A specific introduction is made focusing on the theoretical background to analyse thermal desalination process synthesis. Several advantages are obtained using appropriate strategies in the synthesis of the dual purpose desalination process.

1. Introduction

Seawater desalination processes to provide drinking water are usually coupled to electricity generation plants in many arid regions of the earth. The growing demand for water is accompanied by a parallel rise in the demand for electricity, and both processes require the expenditure of primary energy. Various combinations of power-desalination systems are conceivable. If the desalination processes involve thermal methods such as Multiple Effect Evaporation (MEE) or Multiple Stage Flash Evaporation (MSF), then it is advisable, from the energy consumption viewpoint, that the exhaust steam of the

power-producing turbine is used as the heating medium in the desalination plant (El-Sayed and Silver 1980). Therefore, in such dual-purpose systems, there will exist a Thermal Energy Recovery System (TERS) coupled to Power Generating Cycles (PGCs) which represent the structure of the complex.

Several studies have been published, aiming at the optimization of the desalination complex, or else different structures have been suggested to improve the performance. *However, these are the result of the designer creativity and not a consequence of a systematic or methodological background.* This situation is even more noticeable if we consider the large amount of literature in relation to the synthesis of thermal exchange networks and heat recovery and power generation integrated systems.

The aim of thermodynamic approaches is to apply and adapt the theoretical knowledge of the synthesis theory, to thermal (single or dual purpose) desalination systems using back pressure turbines, extraction-condensation turbines and gas turbines. We will obtain a thermodynamically non-ideal optimal solution for each amount of investment determined by the items used (heat transfer areas, turbine efficiencies and capacities). *The set of solutions forms a basis for the designer selection of those alternatives that satisfy economical optimum and process functional criteria. This feature is important because the relationship between better energy usage and larger investment costs depends on geographical and contingent factors, whereas the thermodynamic relationships do not, thus having general validity.*

2. Maximum Energy Recovery in Desalination Systems

2.1. The Multiple Flash Distillation System (MSF)

Desalination processes separate water from the brine through evaporation. The product must be a liquid; therefore the produced vapor should be condensed. Thus, a thermal exchange system will be attached to the desaltor complex, to which the feed (F), distillate (D) and waste stream (B) are related. On the other hand, an external energy supply (Q_v) and cooling utilities (Q_0) will be needed, as well as the devices for the flow streams involved, and those accessories which allow the effective vapor separation from the brine, its condensation and later accumulation as a distillate. Assuming that the cost of accessories and the chambers are not relevant compared to the thermal exchange area, and the mechanic energy for fluid circulation is negligible compared to the thermal energy consumed in the main heater; then, the most important items for the system synthesis are the heat exchange area and the heat consumed. Thus, the problem is to find the network structure associated to the desaltor complex that uses the minimum amount of utilities (Q_v) and (Q_0) per unit of heat exchange area and produced distillate.

If we adopt the minimum allowable temperature difference between streams (Δt), the system utility demands and the total exchange area can be easily outlined in a (T-H) diagram. To draw the stream evolutions in this diagram, we must remember the composite stream concept. All the heat capacity flowrates (mass flowrate times specific heat) are summed in a unique term, for each temperature interval defined by inlet and outlet stream temperatures, both for all hot and all cold streams. So defined, both composite streams can be represented in the T-H plane, and they are able to shift freely

along axis H. Figure 1 shows the amount of heat recovered (Q_e) from the hot composite stream (HCS) to be transferred to the cold-composite stream (CCS), shifting them so that both composite-streams be overlapped. This shift is limited by the appearance of a situation where temperatures of both composite streams are equalized. This point is a bottleneck or pinch for their integration, since the maximum recovery (Q_e) is determined and therefore, the minimum utilities (Q_v) and (Q_0) also.

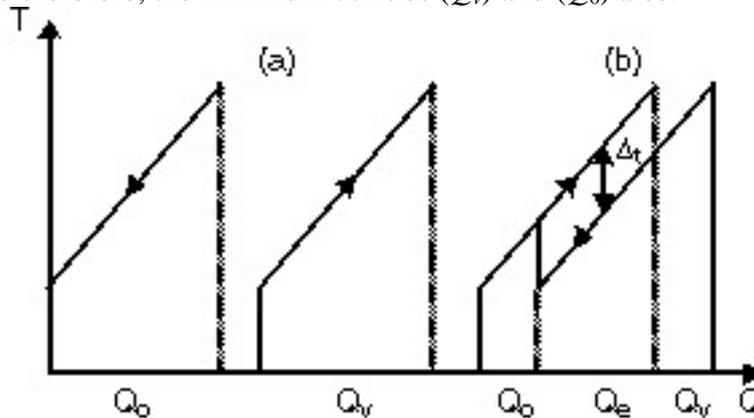


Figure 1. The shifting of Composite Streams determines the maximum Q_e and the minimum Q_v .

Linnhoff and Flowers (1978) suggest a method (TI method) for calculating the minimum utilities, pinch points and thermal requirements for each temperature interval. These are laid out in terms of their thermal levels so that a heat cascade is generated, where the energy flows from the top to the bottom levels. In each level, the amount of heat flow arises from the energy balance and each level is determined in terms of the inlet and outlet temperatures of all the streams of the system.

Unlike the classical problems, we must consider "flash evaporations", unknown temperatures related to each stage (and therefore to the streams involved), amount of distillate (D) that must be produced and finally, the close link existing between the thermal recovery area, the driving force (Δt) and the number of stages (N), which do not occur in classical network synthesis problems. To model the problem under such new conditions, we adopt the following hypothesis:

- Only the heat recovery area is analyzed
- Heat losses are neglected.
- Heat capacity (C_p), boiling point elevation (BPE), heat transfer coefficients (U) and latent heat of evaporation (λ_v) are constant values for each stream in the process.

Let us consider a cold stream F_0 (CS) which evolves taking heat from a hot stream $D+B$ (HS) as indicated in Figure 2, both with the same capacity flow. Unlike a conventional heat exchange system, here the simultaneous vapor and heat transfer implies an enthalpy transfer between both streams. In fact, such transport is done as a consequence of a differential cooling (or flashing) (dT) - the existence of a pressure gradient dP is assumed-. The generated vapor stream (dV) will transfer heat to the CS upon condensing and transferring thermal energy (dQ_e). This will cause a temperature

increase (dt) on the cold stream that by a simple energy balance is equal to the drop (dT) experimented by the HS. The condensed vapor (dV) generates a liquid stream (dD); a new stream whose evolution must be equal to the hot stream which originates it. Therefore, the HCS has a constant heat capacity flow, which is the result of the accumulation of distillate (D) and waste stream (B).

In the T-H plane, the HCS evolution will be a straight line, equivalent to a cooling operation with energy transferred under the form of sensitive heat. This line will be parallel to that representing the evolution of the cold stream, since both have the same heat capacity flow.

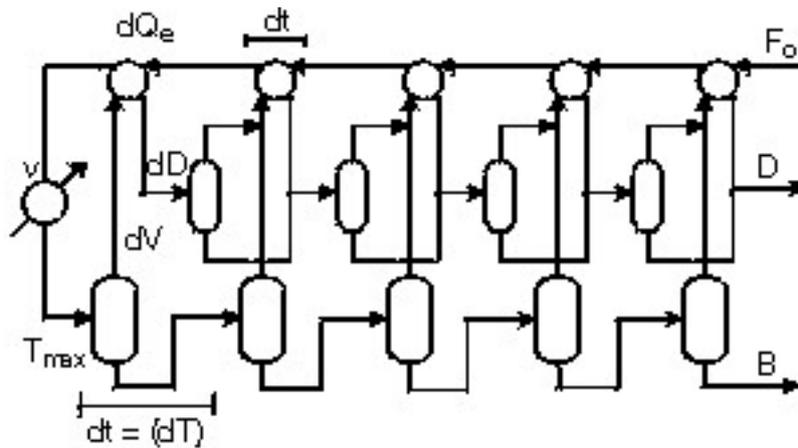


Figure 2. The MSF system and the heat and cold streams.

Part (a) of Figure 1 shows the maximum utility consumption. This quantity decreases when a stream overlapping is produced; or what is the same, the feasibility of thermal exchange among them. This "overlap" is limited by the value of minimum allowable Δt , since $\Delta t = 0$ is assumed, the thermal energy recovery will be maximum but the area will be infinite. Consequently, if a finite Δt is adopted, the structure that corresponds to the maximum overlap implies a minimum utility consumption Q_v and a maximum thermal recovery Q_e , which also implies a maximum production of distillate compatible with this evolution. For a real system, finite temperature drops must be introduced when cooling the HCS. This temperature drop (Δt_f) corresponds to pressure drop (ΔP_f). The value Δt_f is limited by Δt . On the other hand, the vapor produced must be condensed and consequently a new temperature difference must be introduced Δt_e . In this form, in our model the "original" driving force Δt is naturally divided into two components; the temperature drop associated with the pressure drop for the flash operations Δt_f - mass transfer driving force - and the temperature drop for condensation operations Δt_e - driving force for the heat transfer.

$$\Delta t = \Delta t_f + \Delta t_e + \text{BPE} \quad (1)$$

with $\Delta T = T_{max} - T_0$.

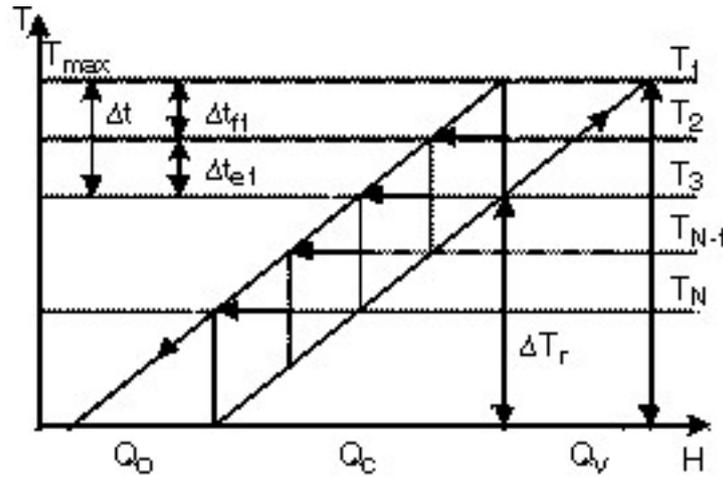


Figure 3. The MSF evolution in T-H diagram and its heat cascade associated with N pinch points. From *Chem. Engng. Sci.* **44** (2) 283-296 (1989).

Adopting a set of temperature differences $\Delta t_{f,n}$, we can draw the evolution indicated in Figure 3. From this, the following relationships among the fundamental variables of the system can be determined:

$$A = \sum_{n=1}^N A_n = \sum_{n=1}^N \frac{FCp}{U} \ln \left(1 + \frac{\Delta t_{f,n}}{\Delta t_{e,n}} \right) \quad (2)$$

$$D = \sum_{n=1}^N V_n = F_0(\rho_1 + (1 - \rho_1)\rho_2 + (1 - \rho_1)(1 - \rho_2)\rho_3 + \dots + (1 - \rho_1)\dots(1 - \rho_{N-1})\rho_N) \quad (3)$$

Here the evaporated fraction is defined in each stage as:

$$\rho_n = V_n / F_{N-1} = F_n (Cp / \lambda_v) \Delta t_{f,n}$$

$$F_0 = F, \quad F_N = B$$

From the necessary and sufficient conditions it can be shown that the assigning police (Scenna 1987a):

$$\Delta t_{f1} = \Delta t_{f2} = \dots = \Delta t_{fn} \quad (4)$$

maximizes production D and minimizes the area A . Then, the specific area (A^*) is minimized. Consequently, the specific area (A/D) in terms of such allocation considering equations 2 and 3 results:

$$A^* = \left(\frac{Cp}{U} \right) \left(\frac{\Delta t_r}{\Delta f} \right) \frac{\ln \left[\frac{(\Delta t - BPE)}{\Delta t_e} \right]^{\Delta T_r / \Delta t_f}}{1 - \left(\frac{1 - Cp \Delta t_f}{\Delta \lambda_v} \right)} \quad (5)$$

where:

$$\Delta T_r = \Delta T - \Delta t, N = \frac{\Delta t_r}{\Delta t_f}$$

From eq. (2) and (4):

$$D = F [1 - (1 - \rho)^N]$$

From eq. (3) and (4), and

$$A = \left(\frac{F C_p}{U} \right) \ln \left(1 + \frac{\Delta t_{f,n}}{\Delta t_{e,n}} \right)^N$$

On the other hand, once Δt was adopted, the specific energy consumption

$$(Q_v^*)$$

is determined:

$$Q_v^* = \frac{Q_v}{D} = \frac{C_p \Delta t}{\left[1 - \left(1 - \frac{C_p}{\lambda_v} \Delta t_f \right) \right]^{\Delta T_r / \Delta t_f}} \quad (6)$$

Then, once parameters Δt and N have been determined, minimum utility

$$(Q_v^*)$$

and specific area are achieved by finding out the thermal exchange operations involving the maximum overlap between composite streams and the allocation policy shown in equation 4. The relations obtained are not greatly affected if the waste stream recirculation is adopted.

This originates the known "rejection" and thermal "recovery" sections in the MSF systems. If the same Δt_f is adopted in both zones, there exists a simple relationship among parameter Δt , N and those belonging to the rejection zone (Scenna 1987a). On the other hand, the global area is not very sensitive to the flow variation (with respect to the feed flow) used for the cooling stream in the rejection zone. Consequently, no meaningful mistake is made when using the previously obtained relationships.

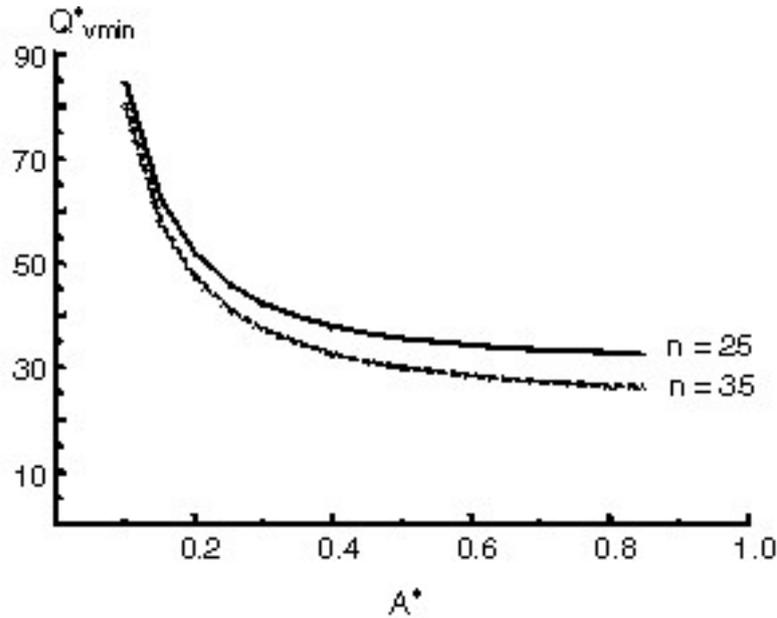


Figure 4. The minimum specific heat consumption - with stage number as parameter - vs. the the specific area in a MSF system. From *Chem. Engng. Sci.* 44(2) 283-296 (1989).

Figure 4 depicts the optimal values of Q_{vmin}^* , vs. A^* for a given value of N . In this curve, each point represents an optimal solution from the thermodynamics point of view, but it is also possible to determine economic optimal points. In fact, if the cost per unit of product (C) is expressed as:

$$C = C_a A^* + C_H Q_v^* \quad (7)$$

It is easy to show that the slope $(dQ_{vmin}^*/dA^*) = -\eta C_a/C_H$ establishes the value of the area such that any further increase of dA^* increases the cost variation. Then, given the unitary cost of the heat exchange area (C_a), the specific cost of the energy (C_c), and the factor η (a function of the amortisation and interest rates); the optimal values

$$(N_{min}, Q_{vmin}^*, A_{min}^*)$$

can be determined. Obviously, these are approximate values because we have assumed that the cost associated to each stage at the optimum is negligible. In this form, starting from general curves, with a few data we can find out all the optimal parameters for the preliminary design of a MSF system. Moreover, these values so obtained agree with others reported in the literature using mathematical programming (Coleman 1971).

It can be concluded that by using this model, the determination of the optimum parameters for the preliminary design of MSF systems is possible, extending the concepts of the synthesis of heat exchange networks as was proposed by Linnhoff and Flowers (1978). Nevertheless, if quantitative relationships are to be derived for MEE systems, we must extend the model "adding" a heat duty in the upper thermal level to originate a vapor stream, giving rise to an evaporation cascade through the stages. Such

a situation generates an increase in distillate production, as well as a variation of the heat exchange area. However, the heat cascade does not modify its thermal levels or the location of the N pinch points. Consequently, the structural relationships obtained for the MSF systems are also valid for MEE systems, since they were obtained from the analysis of such cascade. In the next section, we will derive optimal solutions for the MEE systems and its relation with the optimal solutions for MSF systems, using the previous optimal principles.

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