SYNTHESIS AND DESIGN OF OPTIMAL THERMAL MEMBRANE DISTILLATION NETWORKS

A Thesis

by

MADHAV NYAPATHI SESHU

Submitted to the Office of Graduate Studies of
Texas A&M University
in partial fulfillment of the requirements for the degree of

MASTER OF SCIENCE

August 2005

Major Subject: Chemical Engineering

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Approved by:

Chair of Committee,	Mahmoud El-Halwagi
Committee Members,	Juergen Hahn
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Head of Department,	Kenneth Hall

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ABSTRACT

Synthesis and Design of Optimal Thermal Membrane Distillation Networks. (August 2005)

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Chair of Advisory Committee: Dr. Mahmoud El-Halwagi

Thermal membrane distillation is one of the novel separation methods in the process industry. It involves the simultaneous heat and mass transfer through a hydrophobic semipermeable membrane through the use of thermal energy to bring about the separation of a feed mixture into two streams- a permeate and a retentate stream. Traditionally, studies on this technology have focused on the performance of individual modules as a function of material of the membrane and also configuration of the membrane. However, an investigation into the performance of a network of these modules has not been conducted in the past. A hierarchical parametric programming technique for synthesis of an optimal network of these modules is presented. A global mass allocation representation involving sources and sinks was used to solve the problem and derive criteria for optimality in specific regions of the parametric space. Two case studies have been presented to illustrate the applicability of the presented methodology.

ACKNOWLEDGMENTS

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I would like to thank my colleagues at the Process Integration Laboratory for their assistance and their company in the graduate office for the past two years. I would like to thank the departmental staff for making things smooth for me. I am also thankful to the team at Matrix Process Integration for giving me an opportunity to work on the consulting project that inspired the development of this work and for providing the inspiration to develop my thesis. I am thankful to Dow Chemical Company, NASA and the Department of Energy for funding my graduate studies.

I would also like to thank my friends from Bangalore for all the good times and for the intellectual stimulation over the years. Finally, I would like to thank my parents for their love and support.

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I INTRODUCTION

Thermal membrane distillation (TMD) is emerging as one of the promising separation technologies in the chemical process industry. As the name indicates, TMD has similarities to both membrane separation and distillation. As noted by Smolders and Franken (1989), “Both TMD and conventional distillation rely on vapor-liquid equilibrium as a basis for separation, and both processes require that the latent heat of vaporization be supplied to achieve the characteristic phase change.” A typical TMD setup is shown in Fig.1.

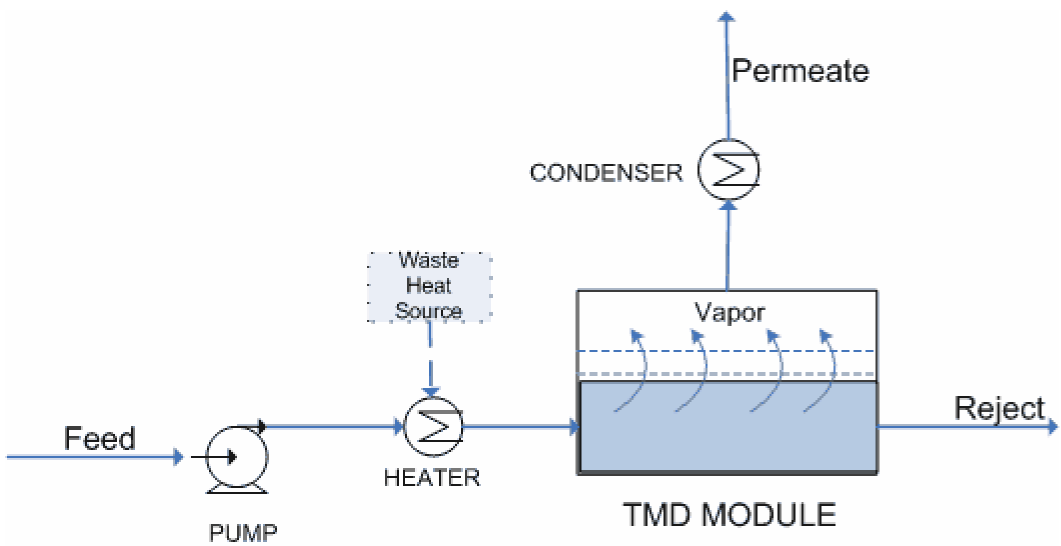

Fig.1. Typical membrane distillation unit

This thesis follows the style of Chemical Engineering Science

The projected benefits of the technology (Lawson and Lloyd, 1997; Weyl, 1967) are:

- 100 % theoretical rejection of ions, macro molecules, colloids, cells, and other non-volatiles from the solution
- Lower operating temperatures than conventional distillation
- Lower operating pressures than conventional pressure-driven membrane separation processes
- Reduced chemical interaction between membrane and process solutions
- Less demanding membrane mechanical property requirements
- Reduced vapor spaces compared to conventional distillation processes
- Lower capital, energy and space expenditure compared to distillation

Multi- stage operation of MD is amenable to recovering and reusing latent heat of vaporization many times, thus leading to reduced energy consumption. This can be implemented by deploying a strategy of using the permeate of one MD module to heat the feed of another and so on (Weyl, 1967).

1.1. REVIEW OF MEMBRANE DISTILLATION TECHNOLOGY

Lawson and Lloyd (1997) presented a comprehensive overview of membrane distillation technology and outlined the terminology and concepts associated with it, including transport phenomena, membrane properties and module design.

Sirkar (1997) presented an overview of membrane separation technologies containing details of membrane distillation technology and the scope for extension of membrane technologies to hybrid separation networks.

Several instances have been reported regarding the applications of membrane distillation as a novel separation technology. An investigation by Gryta and Karakulski (1999) about the applicability of membrane distillation for concentrated oil-water emulsions reported poor performance for solutions with high concentrations of oil.

An overview of Direct Contact Membrane Distillation (DCMD) reports the narrow range of applications of membrane distillation in the industry, such as desalination and water purification (Burgoyne and Vahdati, 2000; Cath et al., 2004). Gryta et al. (2001) and Karakulski et al. (2002) have studied the applicability of membrane distillation to water purification and potable water quality improvement.

Bouguecha and Dhahbi (2003) have studied the potential benefits of membrane distillation as a technology that can be driven by thermal energy at low enthalpy, such as geothermal energy, by using a hybrid air gap membrane distillation- fluidized bed crystallization assembly for desalination.

Tomaszewska (2000) has studied the application of membrane distillation in various applications such as in water and waste water treatment, in the food industry and in the concentration of sulfuric and hydrochloric acid solutions.

Vacuum membrane distillation has been studied experimentally for the concentration of fruit juices (Bandini and Sarti, 2002). Experimental studies on orange juice using a hydrophobic flat membrane and theoretical studies, accounting for heat and mass transfer, membrane morphology and juice solution properties have been carried out by Calabro et al. (1994).

Problems such as fouling of the membrane surface occurring during operation of membrane distillation modules in applications such as ultrapure water production, removal of bioreactor metabolites, concentration of acid and salt solutions and disposal of spent solutions have been discussed by Gryta (2001).

1.2. STUDIES ON MEMBRANE DISTILLATION

Rigorous models have been developed for generic membrane separation processes, with consideration of mass, momentum and energy balances and subsequent verification by experiments (Marriott and Sorensen, 2003). Experimental studies of thermal membrane distillation module performance as a function of reject temperature, salt concentration, cycle time and membrane material are available in literature (Peng et al., 2005). The effects of heat and mass transfer in membrane distillation have also been studied through experimental measurements (Schofield et al., 1987; Martinez-Diez and Vazquez-Gonzales, 2000)

Experiments have been carried out for the direct contact configuration of membrane modules, with consideration of fluid mechanics in the modules and the effect of membrane structure and properties on the permeate flux (Lagana et al.,2000; Fernandez-Pineda, 2002).

Theoretical studies have been conducted on sweeping gas membrane distillation systems using a Stefan-Maxwell based model to study vapor-liquid equilibria and heat and mass transfer relations (Rivier et al., 2002). This model has been used to predict the flux and selectivity under the given operating conditions. Experimental studies have also been conducted on gas permeation systems to determine characteristic parameters of the Knudsen and Poiseulle transport mechanisms (Fernandez-Pineda, 2002).

Models have been developed for plate and frame membrane distillation units based on mass and energy balances for hydrodynamic, temperature and concentration boundary layers.

An important feature in membrane distillation is the effect of module temperature on the permeate mass flux. It has been understood that membrane distillation is suited to applications with a high membrane temperature. Calculations have been presented in literature to estimate the performance of a flat plate unit in which thermal fluids in separate circuits area used to supply and remove the enthalpies of distillation and condensation (Foster et al., 2001). Recommendations have been consequently made to study the effects of pre-pressurizing of the membrane pores and control of dissolved gas concentrations in the feed and recycled permeate in order to prevent pore penetration and wetting (Agashichev and Sivakov, 1993).

Temperature polarization effects have been studied through experimental investigations of mass fluxes and evaporation efficiencies (Martinez-Diez, 2000), and also through the use of process models (Agashichev and Sivakov, 1993).

The effect of membrane material has been investigated in flat membrane modules to correlate the polymer content of the membrane casting solution and the membrane thickness with the magnitude of the flux (deZarate et al., 1995).Experimental studies have also been conducted to study the sensitivity of system performance to membrane material (Ohta et al., 1991).

The effect of concentration temperature and stirring rate on vapor flux in a PTFE membrane was studied experimentally was studied experimentally to yield insights on the thickness of thermal and concentration boundary layers (Sudoh et al., 1997).

This paper also studied the effects of thermal and concentration boundary layers on vapor permeation in the membrane distillation of an aqueous lithium bromide solution

Lawson and Lloyd (1996) carried out experiments on a lab scale on a DCMD apparatus without support for flat sheet membranes to measure the permeability parameter associated with molecular diffusion in membrane distillation. The experimental data was a good fit with the dusty gas transport model through porous media.

1.3. PAST STUDIES ON MEMBRANE NETWORK SYNTHESIS

The problem of synthesis of membrane networks has been solved in different cases in literature. Different solution techniques have been employed in the past to solve the problem. Uppaluri et al., (2004) have used a robust stochastic technique using a simulated annealing procedure for minimization of annualized cost of gas permeation networks. An optimal design strategy has been proposed by Qi and Henson (2000) for membrane networks separating multicomponent gas mixtures. The method is useful for screening of multi stage separation systems for multi component gas mixtures. Kookos (2002) presented an approach that could optimize the selection of membrane material along with the structure of the membrane network.

El-Halwagi (1992) proposed a technique for synthesis of reverse osmosis networks (RONs) for waste minimization. Srinivas et al., (1995) proposed a synthesis strategy for design of optimal pervaporation networks. In both cases, the procedure involved a structural representation of the process to embed all potential process streams. An optimization problem was formulated to minimize the total network cost. The objective was to minimize the total annualized network cost subject to technical, economic and environmental constraints. The solution of the optimization formulation included the optimal network structure and the specifications of the separation modules, booster pumps, energy recovery turbines, heaters and coolers.

Marriott and Sorensen (2003) introduced a novel design procedure for membrane systems incorporating detailed process models and a solution procedure involving the application of genetic algorithms. A pervaporation case study was used to validate the presented procedure.

A concept of special relevance to the TMDN synthesis problem is that of the heat induced separator, which is defined as an indirect contact unit, which employs an energy-separating agent to cause separation due to phase change. A systematic procedure for synthesis of Heat Induced Separation Networks (HISEN) was proposed by El-Halwagi et al., (1995).

A design procedure for Reverse Osmosis Networks (RON) subject to fouling was detailed by See et al., (2004). The study incorporates the data of fouling behavior into the design and operation of a two-stage network. The optimization formulation is non-linear due to the nature of fouling behavior, membrane behavior, network interactions and operating parameter constraints. The problem was solved using three optimization approaches:

- Comparison of individual designs;
- Deterministic gradient search methods and
- Simulated annealing based hybrid stochastic-deterministic approach.

1.4. EXTENSION TO MEMBRANE DISTILLATION-HYBRID NETWORKS

A generic procedure for the optimal design of membrane hybrid systems was presented by El-Halwagi (1993). The paper provides a framework for screening potential separation processes and synthesizing a minimum cost hybrid network of membrane modules and mass exchangers.

Crabtree et al., (1998) addressed the design of hybrid gas permeation membrane / condensation systems for pollution prevention using a Mixed Integer non Linear Program (MINLP). A short cut method was presented to solve the above problem and prove the merits of the hybrid system over the individual technologies.

A model for the analysis of a membrane distillation- crystallization hybrid system for concentration of sodium chloride solution was developed by Gryta (2002) and verified by experimental data. The effect of membrane wetting on long-term performance was also studied.

The performance of a fermentation process integrated with membrane distillation was investigated by Gryta (2002). It was found that the membrane distillation module aided the fermentation process due to removal of by-products. Also, the carbon-di-oxide from the fermentation process aided the transport of liquids through the boundary layer of the membrane. Experimental studies on hybrid membrane systems in desalination processes indicate the merits of integrated operations such as micro filtration, ultra filtration and nanofiltration modules in series and also Reverse Osmosis- Membrane Distillation systems (Drioli et al., 1999).

A hybrid superstructure obtained from combining the superstructures of distillation columns and of gas permeation networks was used for structural and parametric optimization for a propylene-propane splitter and displayed the merits of system integration (Kookos, 2003).
Kovasin et al., (1986) employed a dynamic programming approach to synthesize an optimal membrane network of ultra filtration- diafiltration modules using dynamic programming.

For large-scale applications, multiple TMD modules are needed. These modules may be arranged in series, parallel, or combination. In order to enhance the permeation flux, the feed to each TMD may be heated to some optimal temperature. The vapor permeate has to be condensed. Therefore, a number of heaters and condensers may be needed. Because of the relatively low flux of TMD, it may be necessary to increase the feed flow rate to each stage through reject recycle. Consequently, it is important to synthesize optimal TMD networks (TMDN). A TMDN is composed of multiple TMD modules, pumps, heaters, condensers, mixers, and splitters. The following sections describe the problem of synthesizing a system of TMD modules and a systematic procedure for designing an optimal TMDN.

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II PROBLEM STATEMENT

The TMDN synthesis task can be stated as follows: Given a feed flow rate Q_F and feed concentration C_F, it is desired to synthesize a minimum cost system of TMD modules, booster pumps, heaters, and condensers that can separate the feed into two streams; a permeate and a retentate (reject). The cost of the system consists of two annualized cost components- the capital cost for equipment and operating cost for heating and cooling requirements.

The system is designed toward a specified product stream (permeate) which is required to meet two requirements:

1. The permeate flow rate should be no less than a given flow rate, i.e.,

$$Q_P \geq Q_P^{\min} \tag{1}$$

2. The concentration of a certain species in the permeate should not exceed a certain limit (e.g., environmental, salt content, product quality, etc.):

$$C_P \leq C_P^{\max} \tag{2}$$

The flow rate per module is typically bounded by manufacturer's constraints:

$$q_F^{\min} \leq q_F \leq q_F^{\max} \tag{3}$$

Fig. 2 shows a schematic representation of the TMDN problem.

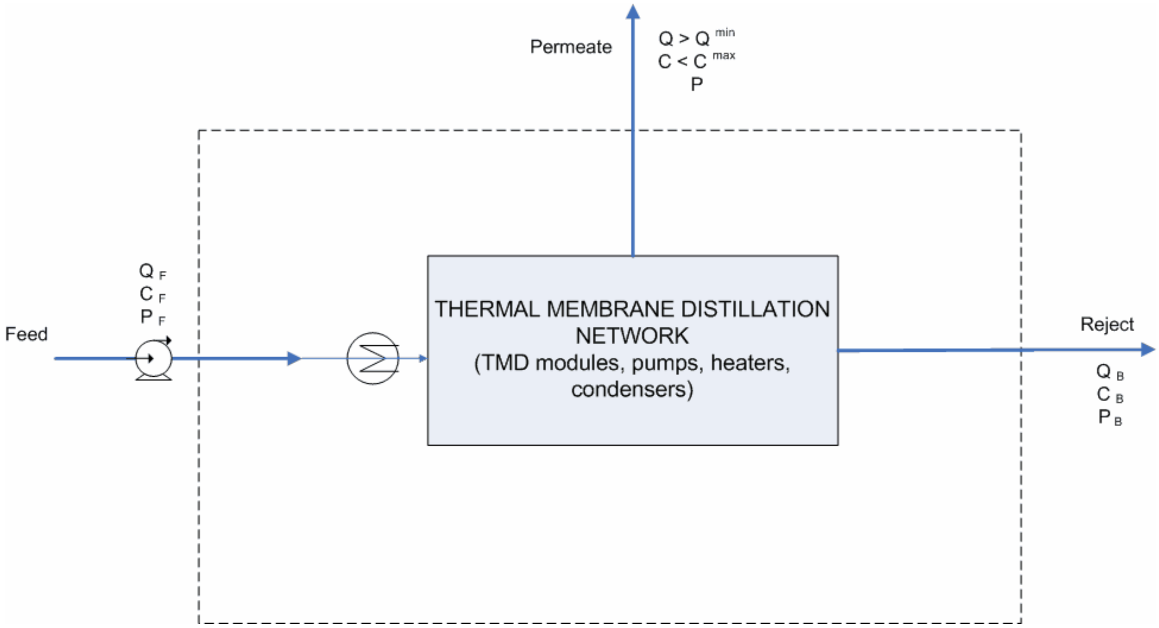

Fig. 2. Schematic representation of TMDN synthesis problem

2.1. DESIGN CHALLENGES

The TMDN synthesis problem is associated with the following design challenges:

- The number of modules to be used
- The total membrane area for the network
- The configuration of the modules (e.g., in series, parallel, etc.)
- The placement of heaters and condensers
- The optimal values of heating and cooling duties
- Energy integration be achieved between heaters and condensers
- The optimal values of operating variables for each module (temperature and concentration)
- Optimal allocation policy for recycle and bypass streams

2.2. POSSIBLE CONFIGURATIONS OF A MEMBRANE NETWORK

Some common configurations of membrane networks are the series, parallel and the tapered arrangements. Membranes are usually arranged in series, as shown in Fig. 3. , when a higher degree of separation is required than is possible in an individual module.

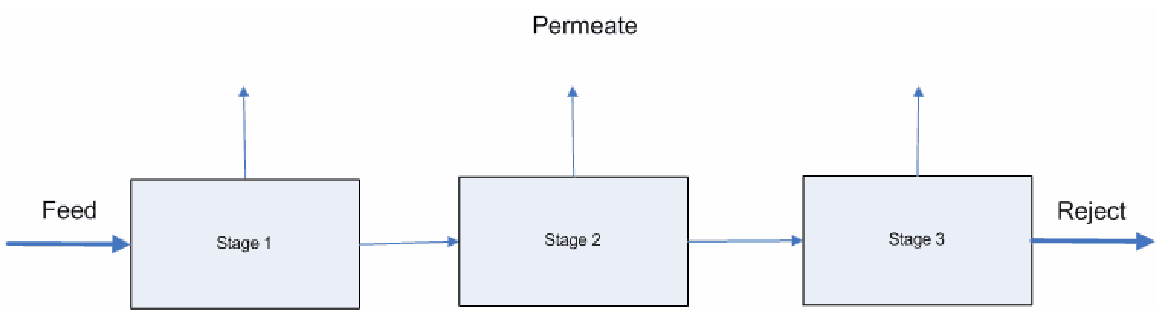

Fig. 3. Schematic representation of series membrane module configuration

Membranes modules are arranged in parallel when the volume of the feed to the network exceeds the capacity of an individual module. .A representation of a parallel membrane configuration is shown in Fig 4.

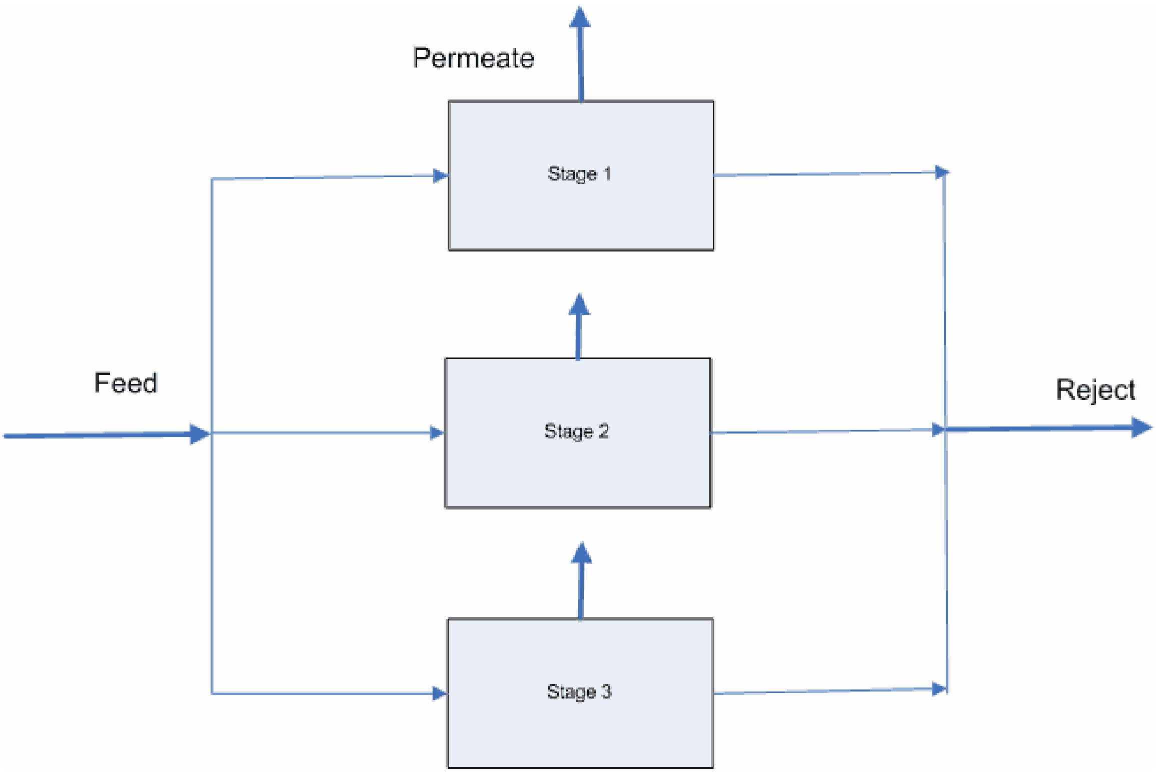

Fig. 4. Schematic representation of parallel membrane module configuration

The tapered arrangement of membranes, also called Christmas tree arrangement is a hybrid of the series and parallel configurations. It consists of a series of membrane stages. Each stage consists of a single membrane or a set of membranes in parallel, as shown in Fig. 5.

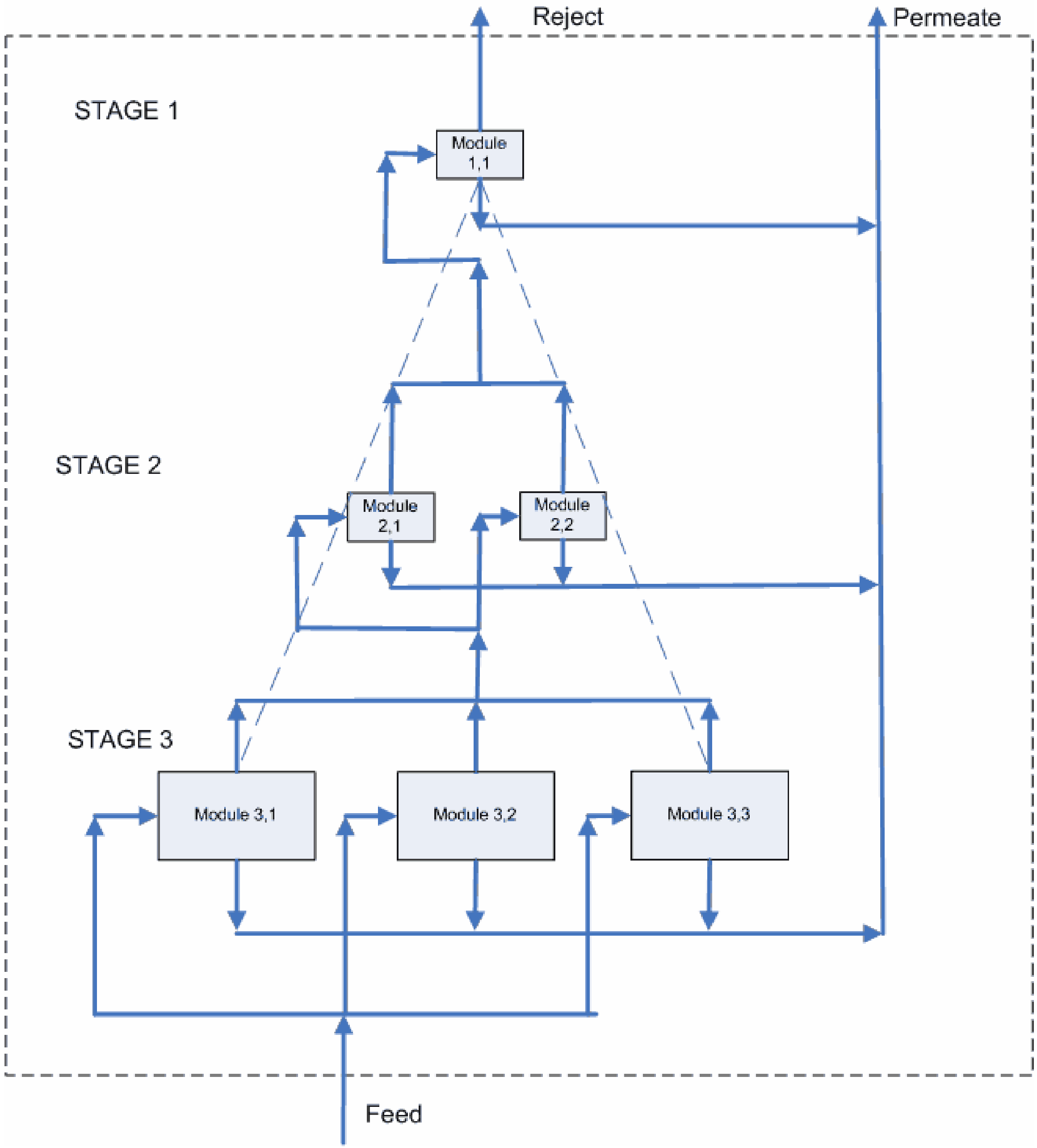

Fig. 5. Schematic representation of tapered membrane module configuration

III MODEL DEVELOPMENT-NETWORK SYNTHESIS

3.1. SOURCE-SINK REPRESENTATION OF TMDN

A source-sink approach (El-Halwagi et al., 1996) was employed to represent the network configuration. A source is a process stream that is rich in the constituent(s) that has to be removed in the separation task.

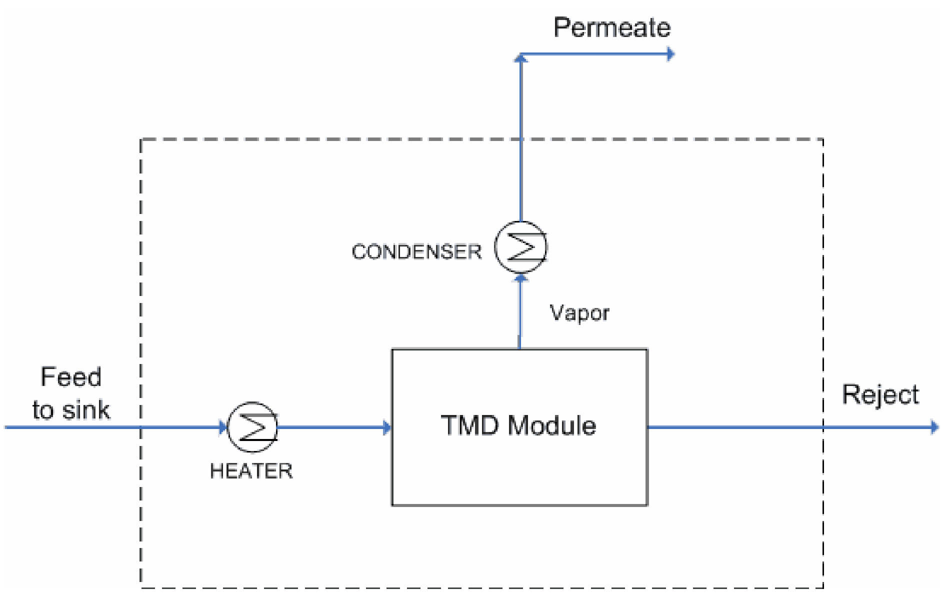

Fig. 6. Building block for a sink

The sink is a destination of a source and may be composed of a group of units that generate other sources in their own right. In the TMDN case, we propose a sink composed of a building block consisting of a heater, a TMD module and a permeate condenser. The size of each element is an optimization variable (including a zero size which indicates that the element does not exist). Each sink produces two sources (permeate and reject). Either one or both may be rerouted back to the network to be assigned to new sinks.

A network superstructure was formulated to account for all possible system configurations, as shown in Fig. 6.

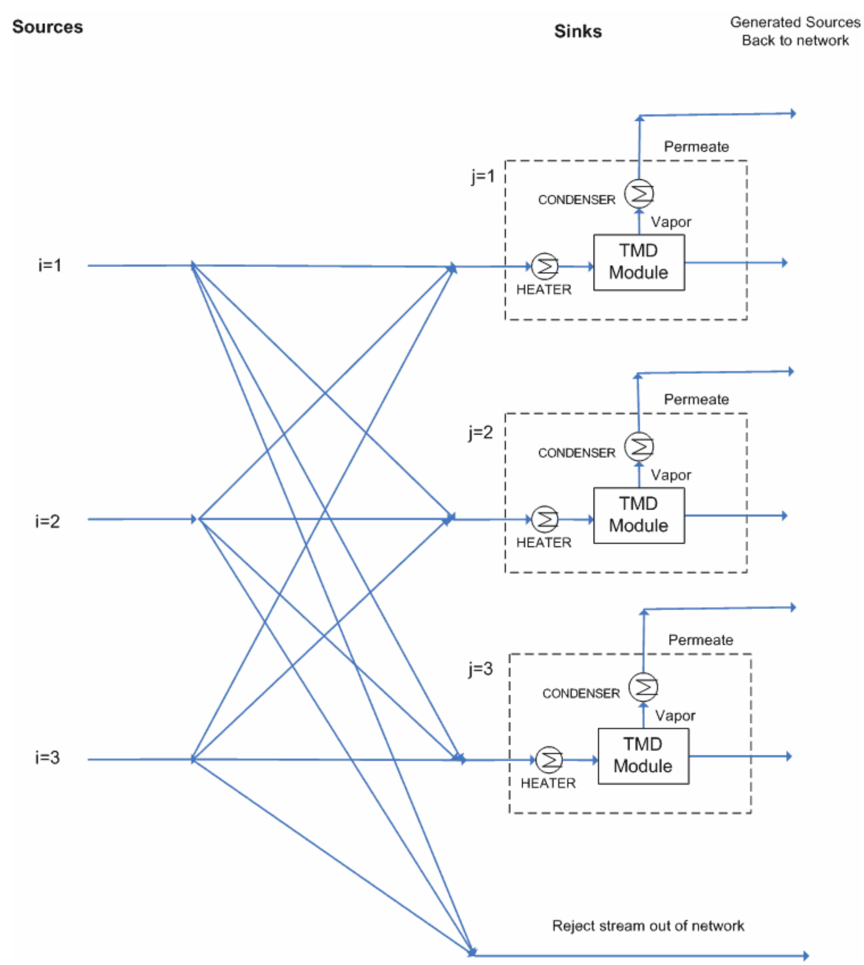

Fig. 7. Network superstructure

In the case of the TMD network problem, the fresh influent and the reject streams from individual modules are modeled are sources. The individual membrane distillation modules are represented as sinks. Fig. 7 illustrates the representation for the sink building block with its three elements: the heater, the TMD module, and the permeate condenser.

Each source may split into several fractions to be fed to the sinks. In turn, each sink produces two sources (permeate and reject) that may be recycled back to the network for further assignment to another sink. In the special cases when there is a sharp separation leading to a pure permeate, then the permeate stream leaves the network without the need for further processing.

The network synthesis problem can now be formulated as follows:

"Given a set of process streams(sources) R= { i | i= 1, $N_{SOURCES}$ } and a set of sinks(modules) S = { j | j= 1, N_{SINKS} }, it is desired to synthesize a network of thermal membrane distillation modules, heaters, condensers and booster pumps at a minimum annualized cost that can achieve the given separation task by preferential transport of selected constituents across the membranes".

The feed to the network is allocated to the various sinks, each of which has a reject outlet component. Permeate is withdrawn from individual modules and leaves the system without further mixing. The reject from a given module in the network is either recycled back to any module in the network (including the same) or bypassed and sent to the outlet of the network.

3.2. OPTIMIZATION FORMULATION

Minimize: Total Cost = (Membrane cost) + (heating cost) + (cooling cost) (4)

Membrane Cost= $Cost_{membrane}\left[\sum_{N}\frac{Permeate_i}{\phi_i}\right]$ (5)

Annualized Membrane Cost = $\left[\dfrac{Cost_{membrane}\left[\sum_{N}\dfrac{Permeate_i}{\phi_i}\right]}{Depreciation\_Period}\right]$ (6)

Heating Cost = $Cost_{heating}\left[\left(\sum_{i}\sum_{j}C_p.Flow_{i,j}.\Delta T_{i,j}\right)+\left(\sum_{j}C_p.Feed_j.\Delta T_{feed,j}\right)\right]$

(7)

Annualized heating cost= Heating Cost * annual operating time (8)

Application of above results in objective function yields (1) subject to the constraints in Section 3.3

Min= $\dfrac{Cost_{membrane}\left[\sum_{N}\dfrac{Permeate_i}{\phi_i(T_i)}\right]}{Dep\_Period}+Cost_{heat}C_p\left[\sum\left(Flow_{i,j}\Delta T_{i,j}\right)+\left(Feed_j\Delta T_{feed,j}\right)\right]$

(9)

3.3. CONSTRAINTS

Overall material balance yields over the network yields (7) and (8)

$$P_i-(\theta_i G_i) \ =0 \tag{10}$$

$$(1-\theta_j).G_j-\sum_{j}Flow_{i,j}=0 \tag{11}$$

Material balance for permeate based on aggregate and modular separation factors α and θ yields:

$$\alpha.Feed-\sum_{N}\theta_j.G_j=0 \tag{12}$$

Accounting for the routing of fresh feed in the network yields:

$$Feed - \left(\sum_{N} Feed_j \right) - Overall\_Bypass = 0 \tag{13}$$

Defining the gross feed to individual module in terms of recycle and bypass components yields:

$$G_j - \sum_{i} \sum_{j} Flow_{i,j} - Feed_j = 0 \tag{14}$$

Temperature dependence of flux at individual module j is modeled as

$$\phi_j(T_j) - k.\exp\left(A - \frac{B}{T_j + C} \right) = 0 \tag{15}$$

$$k = \left(\frac{\phi_{nom}}{P_{sat,nom}} \right) \tag{15-a}$$

Component material balance for the impurity for the network yields

$$z_{feed} Feed = z_{permeate} Permeate + \sum_{N} y_j reject_j \tag{16}$$

Component Material balance for individual module yields

$$z_j Gross_j = z_{feed} Feed_j + \sum_{i} y_j Flow_{i,j} \tag{17}$$

$$\forall i, j : Flow_{i,j} \geq 0 \tag{18}$$

$$\forall j : Feed_j \geq 0 \tag{19}$$

$$\forall j : reject_j \geq 0 \tag{20}$$

$$\forall j : G_j \geq 0 \tag{21}$$

$$\forall j : \theta_j \geq 0 \tag{22}$$

$$\forall j : T_j \geq 0 \tag{23}$$

The objective function for the TMD network has a cumulative character due to summation of the costs of the individual modules. The individual objective functions are structurally the same. Therefore, the minimum network cost is necessarily the sum of the individual minimum costs of modules.

$$MIN(Network\_Cost) = \sum_{N} MIN\left(Module\_Cost\right)_i \tag{24}$$

Similarly, the optimality condition for a module can be extended to the network. In the region of the parametric space corresponding to typical capital and operating costs, the optimum operating temperature is the minimum temperature for the TMDN. The caveat is that the induction is restricted to cases of repeating sink structure.

$$f^{-1}\left(MIN(Network\_Cost)\right) = f^{-1}\left(MIN(Module\_Cost)\right) \tag{25}$$

$$\Rightarrow T_{optimal,network} = T_{optimal,Module} \tag{26}$$

3.4. SOLUTION TECHNIQUE

The model was simulated for progressively increasing levels of network dimension in order to study the response of the objective function as shown in Fig. 8.

3.5. MODEL ASSUMPTIONS

1 Separation performance of the thermal membrane distillation modules is a function of temperature. This is due to the fact that the flux of permeate is a function of temperature

2 Temperature at a module is independent of that at other modules.

3 Specific heat of the mixture does not vary with temperature.

4 Antoine's equation is used to describe the relation between the vapor pressure of the mixture and its temperature.

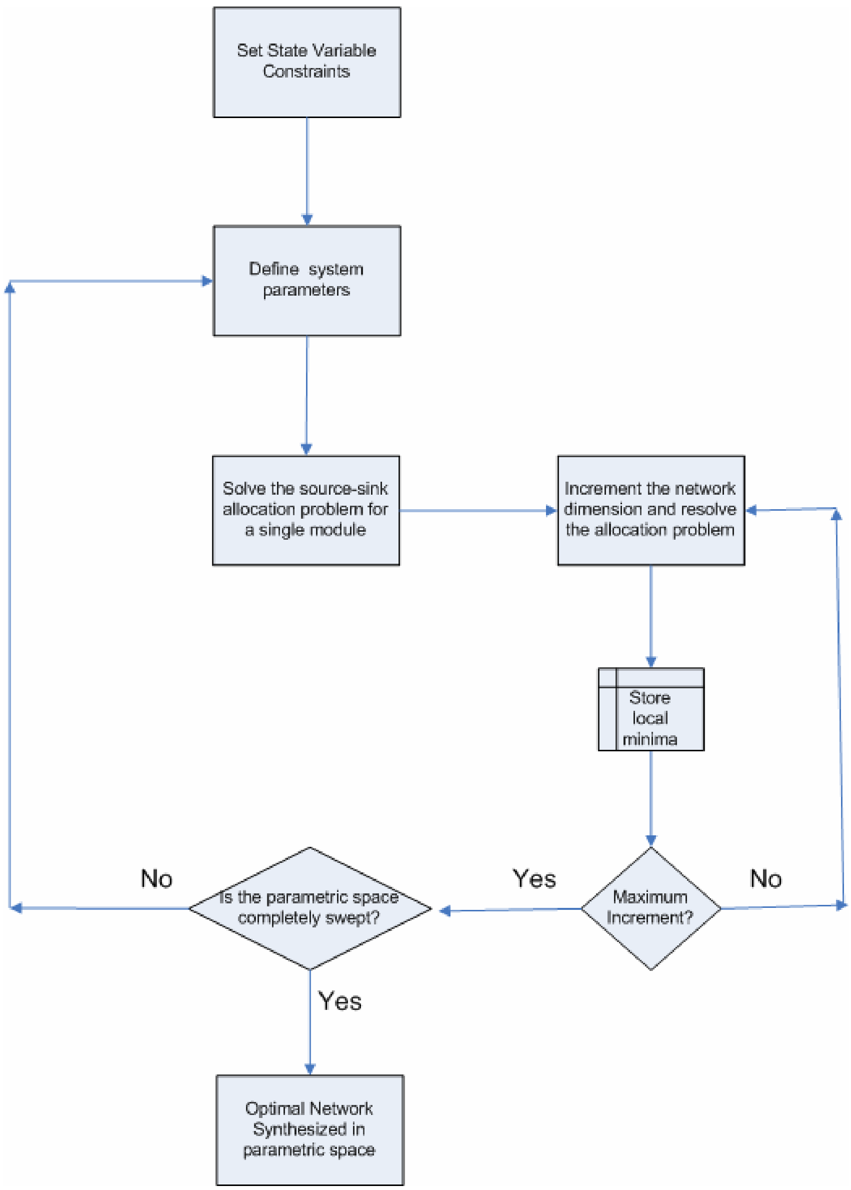

Fig. 8. Hierarchical technique for optimal network synthesis

IV CASE STUDY I: DESALINATION USING TMD TECHNOLOGY

The model described in Section 3 was used to synthesize a TMD network for a desalination case study. The parameters for the system are tabulated in Table 1. The three representative cases are tabulated in Table 2 at different ratios of cost parameters which correspond to different regions of the parametric space

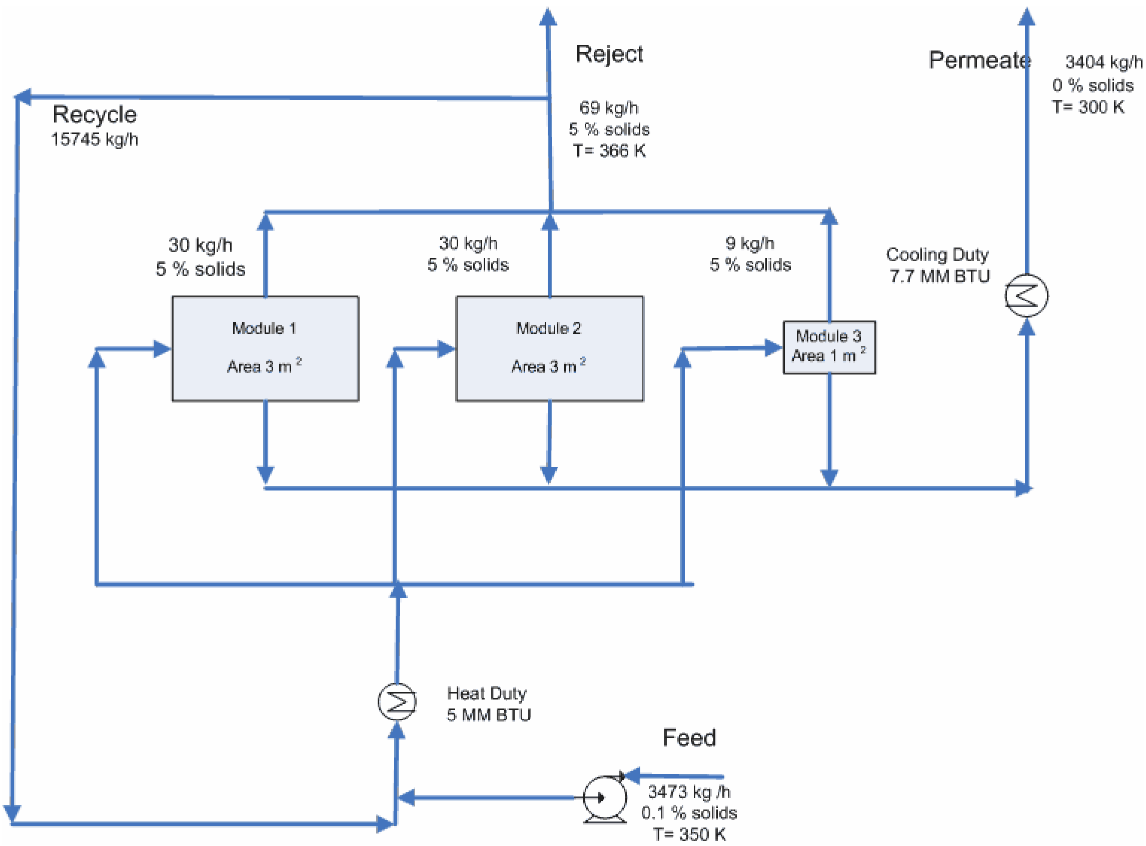

Fig. 9. Synthesized flow scheme for network after optimization

The model was solved using Hyper LINGO version 8.0.The resulting flow scheme synthesized is depicted in Fig. 9.

4.1. LOCAL OPTIMA IN PARAMETRIC SUB-SPACE 1

This is the typical case, in which there is competition between membrane and heating costs.

.

It was observed that the membrane cost is the dominant component of the total cost in the region where the ratio of cost of membrane (per m $^{2)}$ exceeds the cost of heating (per MM BTU) is 0.045.This critical point marks the shift in system behavior. The plane described by this parameter ratio separates two regions with different optimality criteria. The region described by a parameter ratio above the critical ratio (where membrane cost is dominant) has a local optimum corresponding to the maximum value of temperature. The region with a parameter ratio below the critical point consists of two regions.

The first is an intermediate region, where the optimal temperature is between the feed temperature and the maximum allowable operating temperature. This is shown in Figure 2.The optimal temperature is obtained by solving

$$\frac{d}{dT}\left(Total\_Cost\right)=0 \qquad (23)$$

The second is a region, described by a higher parametric ratio than the first, where the membrane cost dominates to a degree that the optimal operating temperature is the maximum allowable temperature. This can be described as a “membrane cost controlling” region. The realistic value of the cost ratio, based on 2005 industry standards is 15.This lies in the membrane cost controlling region. Thus, the optimal value of temperature for of the network is the maximum allowable temperature for the system in the region described by typical values of capital and operating costs. The sensitivity of the objective function to the state variable is shown in Fig. 10.

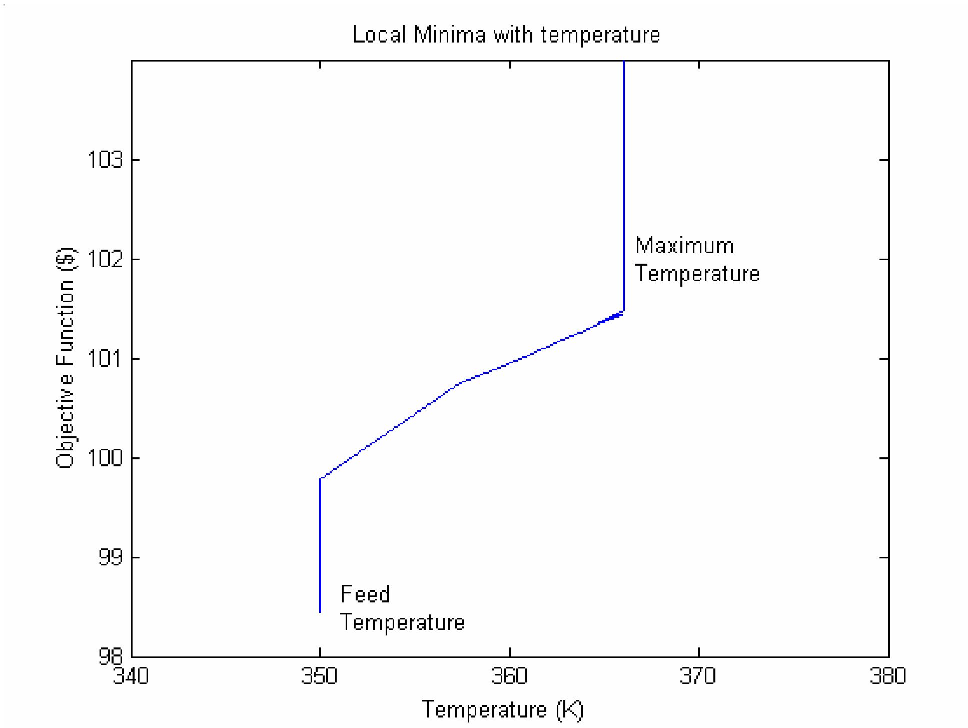

Fig. 10. Objective function with temperature

4.2. LOCAL OPTIMA IN PARAMETRIC SUB-SPACE 2

This is an extreme case which is useful in determining the behavior of the system at an extremity of the parametric space. It is observed that the optimal operating temperature for the system in this parametric region is the temperature of the feed for a region in which the heating cost was the realistic values used in Section 4.1. The objective function was therefore significantly lower than in sub space 1.It should be noted, however, that there exist regions in this sub- space in which the optimal temperature can lie in the intermediate region. This phenomenon corresponds to scenarios described by values of heating cost much lower than the realistic values. The value of this optimal temperature can be obtained by solving (24) and verifying second order criteria for minima.

$$\frac{d}{dT}(Heating\_Cost)=0 \tag{24}$$

4.3. LOCAL OPTIMA IN PARAMETRIC SUB-SPACE 3

This is another extreme case in which the system behavior can be studied another extremity of the parametric space. This is a simple case in which the structure of the objective function dictates that the optimal configuration would lie exclusively in the "membrane cost controlling" region

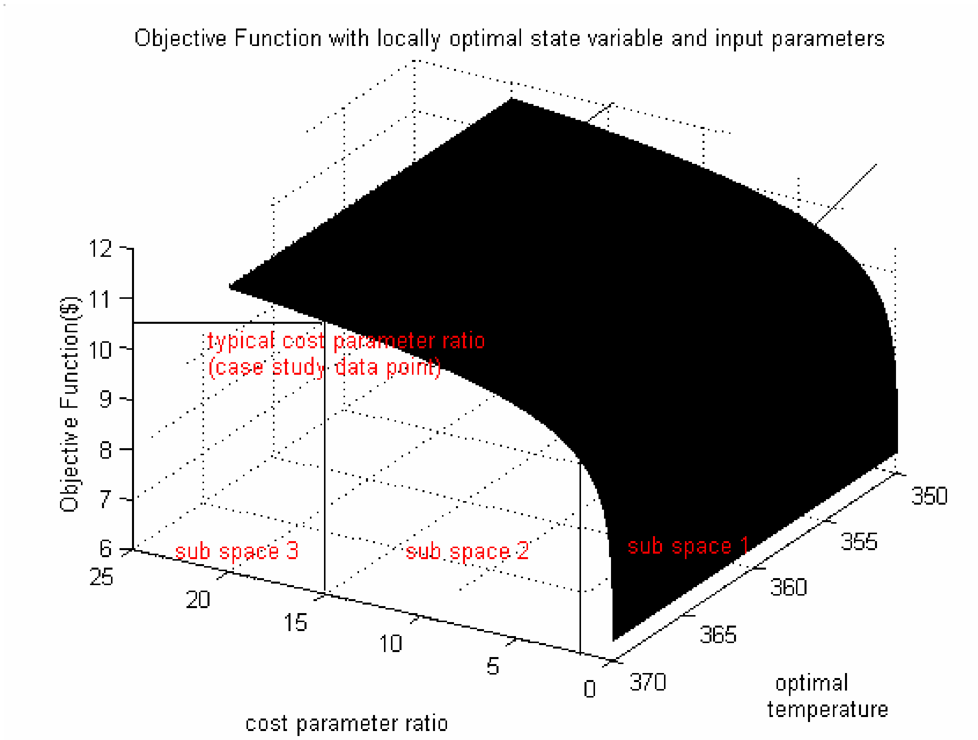

Fig. 11. Objective Function with locally optimal state variable in parametric space

The optimal temperature would therefore be the maximum allowable temperature for the system. A surface plot of parameter ratio, temperature and objective function is shown in Fig. 11.

4.4. GENERAL OBSERVATIONS

The following observations were made based on the experiments described above.

- The total flow rate of recycled reject is constant with respect to system configuration.

- The objective function in all three cases was the same irrespective of flow rate allocation constraints on sources.
- Inter module temperature gradients do not exist in any of the above cases.

V CASE STUDY II: SYRUP CONCENTRATION USING TMD

TMD technology can be employed in the dextrose syrup manufacturing process for the partial concentration of dextrose syrup, shown in Fig. 12. Studies on application of spirally wound modules of hollow cellulose acetate membranes for removing water from a solution of maple sap have been conducted by Underwood and Willits (1969). Laboratory tests indicate that water can be removed from the solution up to the extent of 55 %.This translates to syrup of 11 % concentration. Also, the permeate was found to contain only trace quantities of sugars, which is consistent with the sharp split assumption of the TMD model developed.

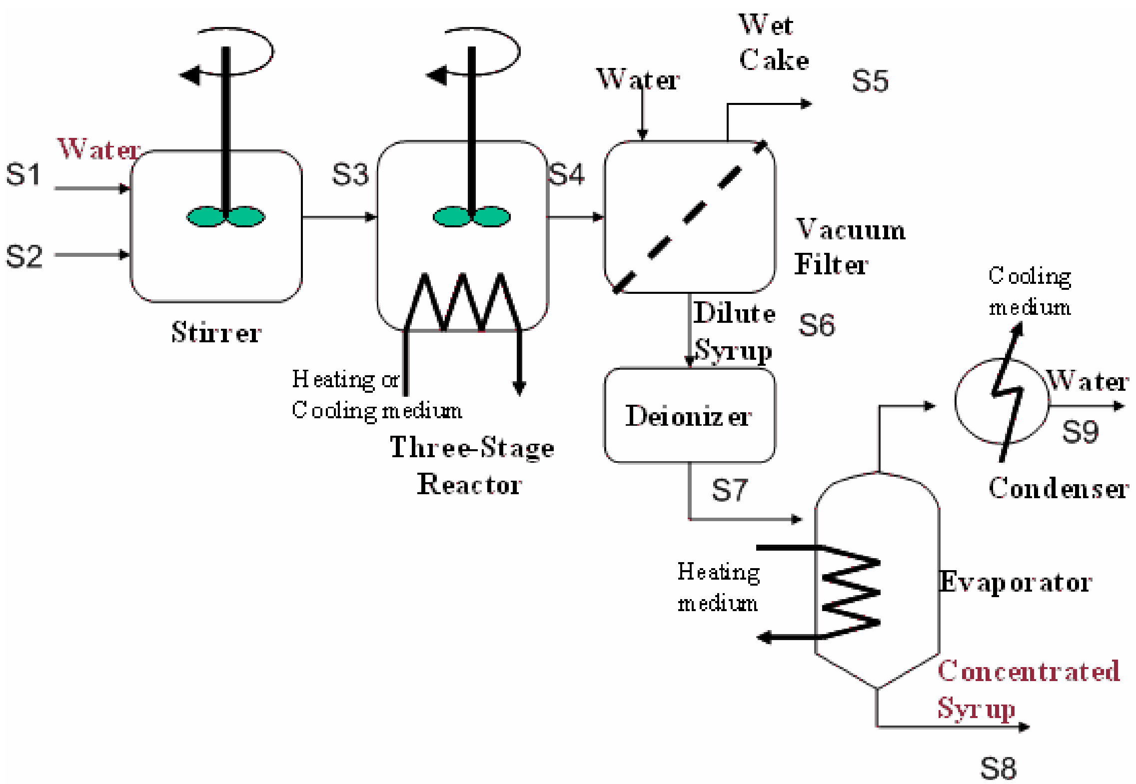

Fig. 12. Dextrose syrup production process (Silayo et al., 2003)

Since the specifications of commercial sugar syrup dictate a dextrose concentration of about 66 %, conventional evaporation is used for the rest of the concentration task. This hybrid separation scheme translates to significant savings on energy costs. The flow sheet for the syrup production process is shown in Fig 12. The objective of the case study is to consider the use of TMDN for syrup concentration in conjunction with the evaporator-condenser scheme. A TMDN was synthesized with stream S7 of the system as the feed to the network. The hybrid separation network for concentration of syrup is shown in Fig. 13.

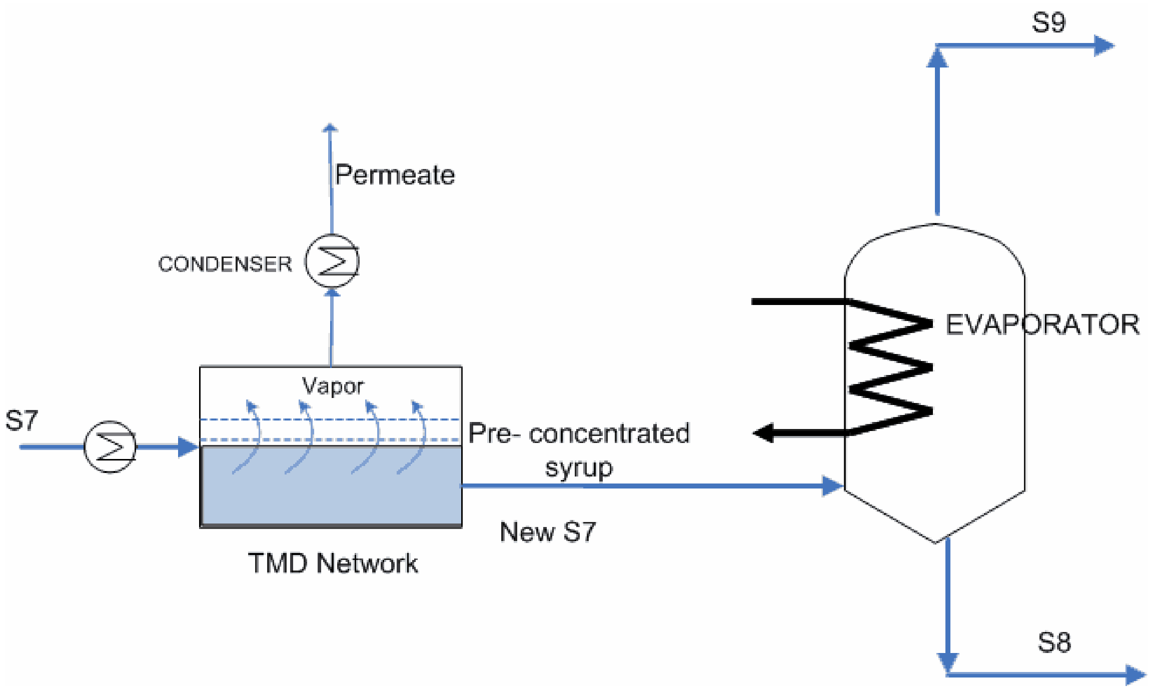

Fig. 13. Hybrid TMD- Evaporation system for concentration of dextrose syrup

A detailed flow sheet of the TMD network synthesized for the separation task is shown in Fig. 14.The reject from the network is the pre-concentrated syrup, which is the new feed to the evaporator. The evaporator heat duty is reduced by over 50 % due to the TMD network. Similarly, a TMDN can be coupled with a traditional separation device to reduce overall energy intensity.

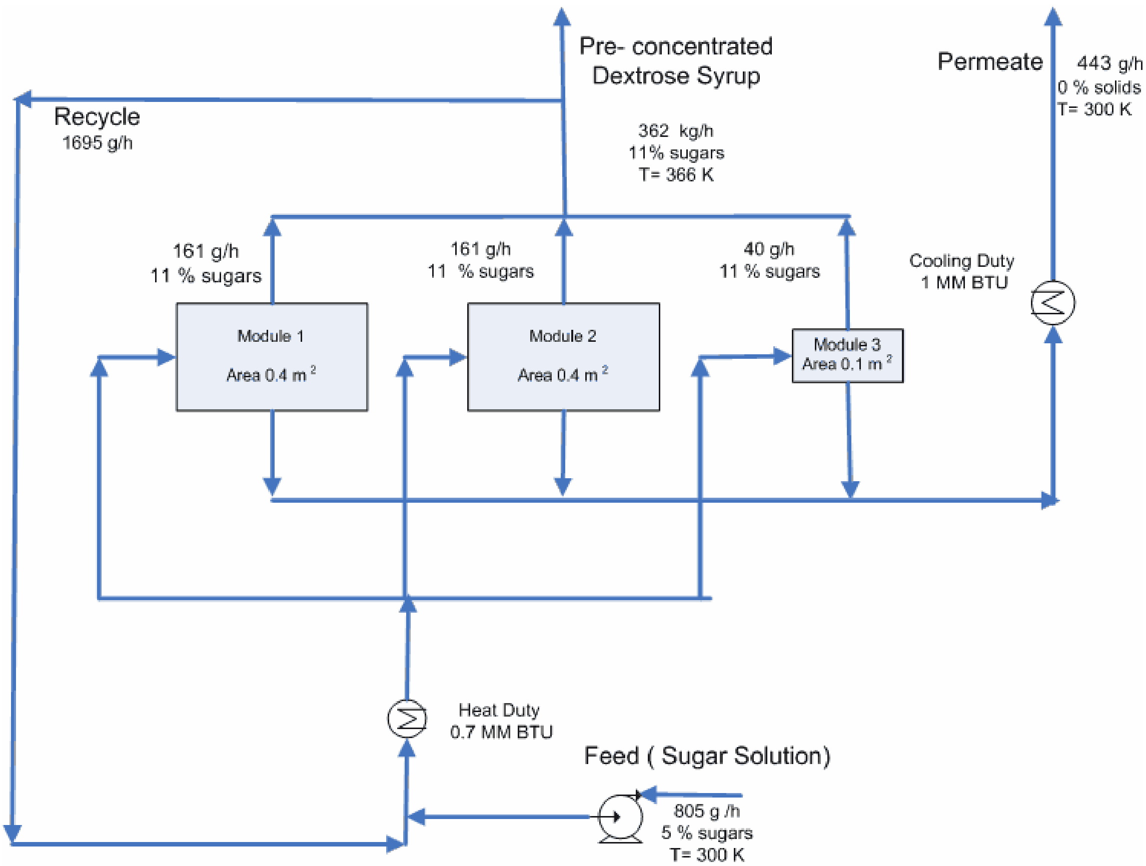

Fig. 14. Synthesized flow sheet for dextrose concentration case study

VI SUMMARY AND CONCLUSIONS

A method has been presented to solve the TMD network synthesis problem. Practical considerations may force the need to use other technologies in conjunction with the TMD network in a hybrid network to achieve the overall separation task. The results of the experiments indicate the fact that a single stage of modules can perform the separation task of a network. In case of limitations on module area, a parallel configuration of modules can be deployed to accomplish the separation task. Also, for values of cost parameters in the case studies , the operating region lies in the parametric sub space in which the optimal operating temperature corresponds to the maximum allowable temperature.

NOMENCLATURE

C_p	specific heat of water , KJ/ kg K
Dep_Time	depreciation period, y
$Feed_j$	mass flow rate of feed stream bypassed to membrane j , kg/h
$Flow_{i,j}$.	mass flow rate from source i to sink j, kg/h
G_j	gross mass flow rate to sink j , kg/h
Overall Bypass	mass flow rate bypassed from feed to the reject, kg/h
$Permeate_j$	mass flow rate of permeate emanating from sink j, kg/h
$N_{SOURCES}$	number of sources in the network superstructure
N_{SINKS}	number of sinks in the superstructure

Greek Letters

ϕ_i	flux of permeate through membrane in module j, kg/ m2 h
θ_j	fractional separation at module j
$\Delta T_{i,j}$	temperature difference between source I and sink j, K
$\Delta T_{feed,j}$	temperature difference between the feed and the module j, K

Subscripts

i,j	from source i to sink j
feed	feed stream descriptor
feed,j	from feed to sink j
nom	nominal value of variable

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APPENDIX A

LINGO PROGRAMS WITH OUTPUT FOR TMDN SYNTHESIS

Typical Case

```
! THERMAL MEMBRANE DISTILLATION NETWORK SYNTHESIS
BINARY AQUEOUS MIXTURE ,LINEARIZED FLUX TEMPERATURE
RELATION , SHARP SPLIT OPERATION

The number of modules(sinks)  is N
the number of sinks and sources is  N.
Subscripts:  i is for sources  and  j is for sinks ;

Sets:

        sources / 1..3/  ;
        sinks/ 1..3 / : normfact,f,z_in,gross_input,
gross_output,reject_bypass,temp,flux,permeate,theta,area,deltat_feed,ysink ;
        placed(sources, sinks) : w,deltat;

Endsets

!Objective Function;

        Min= total_cost;

        total_cost= heating_cost +  membrane_cost+ cooling_cost ;

        heating_cost=@ABS(cost_heat * recycleheat);
        !excess_heat=@IF(recycleheat #GT# 0, 0, @ABS(recycleheat) );
        membrane_cost= (cost_area* totalarea);
        cooling_cost=cost_cool*permeate_total*2260;
        ! pumping_cost=

! Parametric Data;

        ! feed massflow rate in kg/h;
        feed= 3473;
        temp_feed=350;

        ! Feed Concentration;
        zfeed= .001;

        ! Feed Cp in kJ/ kg C;
        cp=4.186;
```

```
!overall recovery is alpha;
alpha=.98;
permeate_total=alpha*feed;

! defining the nominal flux(lb/ ft^2 h ) and temperature(F);
temp_nom=124;
flux_nom=80;
psat_nom=97;

!Operating Cost: cost  of heating in $  per KJ ;
cost_heat= .000006;

ratio=cost_area/(cost_heat*1000000);

!Operating Cost: cost  of cooling in $  per lb ;
cost_cool=.000009;

! Fixed Cost: Cost of membrane heating in $  per metre ^2 of membrane ;
cost_area= 90;
!cost_area=0;

! slope of linearized temperature- flux curve  ( lb/ h F);
m= 3643;

!Nominal recovery at module;
theta_nom= 0.06;
s=329;

!Antoines Constants ;
a=18.3;
b=3816.44;
!c=-46.11;

! temperature drop across the module as afraction;
tempdrop=0.2;

! Generic  Model Equations;

! defining temperature dependency of flux;
@for(sinks(j): normfact(j)=@exp(a- (b/(temp(j)-46.11 ) ) )/psat_nom );
@for(sinks(j): flux(j)=flux_nom*normfact(j) );

! defining temperature dependancy of modular recovery;
@for(sinks(j): theta(j)= theta_nom*temp(j)/temp_nom );
```

```
! Accounting for output of module "j";
@FOR (sinks(j) : gross_output(j)=reject_bypass(j)+
@SUM( sources(i) :  w(i,j) ) );

! Accounting for input to module "j";
@FOR (  sinks (j) : gross_input(j)= f(j) + @sum(sources(i)  : w(i,j) ) );

! Overall Mass balance over module  " j ";
@FOR(sinks(j): permeate(j)=gross_input(j)-gross_output(j));
@FOR(sinks(j): permeate(j)= gross_input(j)*theta(j) );

! Component Mass Balance for module "j";
@FOR( sinks(j) :gross_input(j)*z_in(j)= (  (f(j)*zfeed) +
 @SUM(sources(i): ( w(i,j)*ysink(j)  )) ) );

! Overall Component Mass Balance ;
(zfeed*feed_total) =( zpermeate*permeate_total)+ (
@SUM(sinks(j):reject_bypass(j)*ysink(j)) );

! Sharp Split Mass balance to relate input and output
 concentrations to a module ;
@FOR( sinks(j) :ysink(j) = z_in(j)*gross_input(j)/gross_output(j) );

! Accounting for feed bypass;
feed= @SUM( sinks( j): f(j)) ;

! Accounting for reject bypass;
reject_output =  @SUM(sinks (j) : reject_bypass(j) ) ;

! Accounting for permeate;
permeate_total= (feed)-( reject_output) ;

! Double checking calculations ;
check=@SUM( sinks (j): permeate (j) )-permeate_total ;

! calculating total recycled reject;
@FOR(sources(i): @for(sinks(j): deltat(i,j)=temp(i)-temp(j) )  );
@for(sinks(j): deltat_feed(j)=temp(j)-temp_feed ) ;
recycleheat/cp= @sum(sinks(j): f(j)*deltat_feed(j) )+ @sum(sources(i):
@sum(sinks(j)  : w(i,j)*(deltat(i,j)+(temp(j)*tempdrop) ) ) );
recycled=@sum(sources(i):@sum(sinks(j) :w(i,j) ) );

! calculating area ;
@for( sinks(j):area(j)=permeate(j)/flux(j)) ;
totalarea=@sum(sinks(j):area(j));
```

```
@for(sinks(j):z_in(j)>0.00001 );
@for(sinks(j):z_in(j)<1 );

@for(sinks(j):temp(j)<366);
@for(sinks(j):temp(j)>200);

@for(sinks(j):ysink(j)<=.45);
@for(sinks(j):ysink(j)>=0);

z_reject_overall=@sum(sinks(j):reject_bypass(j)*ysink(j) );
```

```
  Local optimal solution found at iteration:         11559
Objective value:                                   735.7898


                       Variable           Value        Reduced Cost
                     TOTAL_COST        735.7898            0.000000
                   HEATING_COST        30.34367            0.000000
                  MEMBRANE_COST        636.2181            0.000000
                   COOLING_COST        69.22800            0.000000
                      COST_HEAT       0.6000000E-05        0.000000
                    RECYCLEHEAT        5057278.            0.000000
                      COST_AREA        90.00000            0.000000
                      TOTALAREA        7.069090            0.000000
                      COST_COOL       0.9000000E-05        0.000000
                 PERMEATE_TOTAL        3403.540            0.000000
                           FEED        3473.000            0.000000
                      TEMP_FEED        350.0000            0.000000
                          ZFEED       0.1000000E-02        0.000000
                             CP        4.186000            0.000000
                          ALPHA       0.9800000            0.000000
                       TEMP_NOM        124.0000            0.000000
                       FLUX_NOM        80.00000            0.000000
                       PSAT_NOM        97.00000            0.000000
                          RATIO        15.00000            0.000000
                              M        3643.000            0.000000
                      THETA_NOM       0.6000000E-01        0.000000
                              S        329.0000            0.000000
                              A        18.30000            0.000000
                              B        3816.440            0.000000
                     TEMPDROP         0.2000000           0.000000
                   FEED_TOTAL          3473.000           0.000000
                    ZPERMEATE          0.000000           0.000000
                REJECT_OUTPUT           69.46000           0.000000
                        CHECK          0.000000          0.1954711
                     RECYCLED          15745.53           0.000000
            Z_REJECT_OVERALL            3.473000           0.000000
                 NORMFACT( 1)          6.018349           0.000000
                 NORMFACT( 2)          6.018349           0.000000
                 NORMFACT( 3)          6.018349           0.000000
                        F( 1)          3472.352           0.000000
                        F( 2)          0.000000           0.000000
                        F( 3)         0.6475901           0.000000
```

Z_IN(1)	0.4114296E-01	0.000000
Z_IN(2)	0.1000000E-04	0.000000
Z_IN(3)	0.5775355E-01	0.000000
GROSS_INPUT(1)	19214.93	0.000000
GROSS_INPUT(2)	0.000000	0.000000
GROSS_INPUT(3)	3.604599	0.000000
GROSS_OUTPUT(1)	15812.03	0.000000
GROSS_OUTPUT(2)	0.2317647E-08	0.000000
GROSS_OUTPUT(3)	2.966236	0.000000
REJECT_BYPASS(1)	69.45077	0.000000
REJECT_BYPASS(2)	0.000000	0.000000
REJECT_BYPASS(3)	0.9227207E-02	0.000000
TEMP(1)	366.0000	0.000000
TEMP(2)	366.0000	0.000000
TEMP(3)	366.0000	0.000000
FLUX(1)	481.4679	0.000000
FLUX(2)	481.4679	0.000000
FLUX(3)	481.4679	0.000000
PERMEATE(1)	3402.902	0.000000
PERMEATE(2)	0.000000	0.000000
PERMEATE(3)	0.6383629	-.8032749E-08
THETA(1)	0.1770968	0.000000
THETA(2)	0.1770968	0.000000
THETA(3)	0.1770968	0.000000
AREA(1)	7.067764	0.000000
AREA(2)	0.000000	0.000000
AREA(3)	0.1325868E-02	0.000000
DELTAT_FEED(1)	16.00000	0.000000
DELTAT_FEED(2)	16.00000	0.000000
DELTAT_FEED(3)	16.00000	0.000000
YSINK(1)	0.4999732E-01	0.000000
YSINK(2)	0.9950333E-05	0.000000
YSINK(3)	0.7018268E-01	0.000000
W(1, 1)	2.007507	-0.3474789E-08
W(1, 2)	0.000000	0.000000
W(1, 3)	2.957009	0.000000
W(2, 1)	15728.48	0.000000
W(2, 2)	0.000000	0.000000
W(2, 3)	0.000000	0.000000
W(3, 1)	12.08411	0.000000
W(3, 2)	0.000000	0.000000
W(3, 3)	0.000000	0.000000
DELTAT(1, 1)	0.000000	0.000000
DELTAT(1, 2)	0.000000	0.000000
DELTAT(1, 3)	0.000000	0.000000
DELTAT(2, 1)	0.000000	24.04921
DELTAT(2, 2)	0.000000	0.000000
DELTAT(2, 3)	0.000000	0.000000
DELTAT(3, 1)	0.000000	0.000000
DELTAT(3, 2)	0.000000	23.65417
DELTAT(3, 3)	0.000000	0.000000

Row	Slack or Surplus	Dual Price
1	735.7898	-1.000000
2	0.000000	-1.000000
3	0.000000	-1.000000
4	0.000000	-1.000000

5	0.000000	-1.000000
6	0.000000	-0.2118600
7	0.000000	0.8722787E-01
8	0.000000	0.000000
9	0.000000	-7.248423
10	0.000000	-755.8972
11	0.000000	-0.2176496
12	0.000000	-0.2849438
13	0.000000	7.952723
14	0.000000	-6.558930
15	0.000000	-5057278.
16	0.000000	0.000000
17	0.000000	-7692000.
18	0.000000	-7.069090
19	0.000000	0.000000
20	0.000000	588.8850
21	0.000000	0.000000
22	0.000000	636.2955
23	0.000000	-1.988864
24	0.000000	-144.7401
25	0.000000	105.6932
26	0.000000	0.000000
27	0.000000	0.1982738E-01
28	0.000000	1.321165
29	0.000000	0.000000
30	0.000000	0.2478422E-03
31	0.000000	199.4755
32	0.000000	0.000000
33	0.000000	0.3742032E-01
34	0.000000	0.1838491E-02
35	-0.2317647E-08	0.1838491E-02
36	0.000000	0.1838490E-02
37	0.000000	0.000000
38	0.000000	0.000000
39	0.000000	0.000000
40	0.000000	-0.1838491E-02
41	-0.2816427E-08	-0.1838491E-02
42	0.000000	-0.1838490E-02
43	0.000000	0.1038128E-01
44	0.000000	0.1038128E-01
45	0.000000	0.1038127E-01
46	-0.2842714E-06	0.000000
47	0.000000	0.000000
48	0.000000	0.000000
49	0.000000	0.000000
50	0.000000	0.000000
51	-0.9950333E-05	0.000000
52	0.000000	0.000000
53	0.000000	0.4018560E-03
54	0.000000	-0.1838491E-02
55	0.000000	0.1838491E-02
56	0.000000	0.1954711
57	0.000000	-0.5040782E-04
58	0.000000	0.000000
59	0.000000	-0.7426674E-04
60	0.000000	23.65417
61	0.000000	0.000000

62	0.000000	0.000000
63	0.000000	-0.3034995E-03
64	0.000000	23.65417
65	0.000000	0.000000
66	0.000000	-0.8721160E-01
67	0.000000	0.000000
68	0.000000	-0.1626485E-04
69	-0.1494773E-06	-0.2511600E-04
70	-0.2316877E-08	0.000000
71	0.000000	-90.00000
72	0.000000	-90.00000
73	0.000000	-90.00000
74	0.000000	-90.00000
75	0.4113296E-01	0.000000
76	0.000000	0.000000
77	0.5774355E-01	0.000000
78	0.9588570	0.000000
79	0.9999900	0.000000
80	0.9422464	0.000000
81	0.000000	0.000000
82	0.000000	0.000000
83	0.000000	23.65838
84	166.0000	0.000000
85	166.0000	0.000000
86	166.0000	0.000000
87	0.4000027	0.000000
88	0.4499900	0.000000
89	0.3798173	0.000000
90	0.4999732E-01	0.000000
91	0.9950333E-05	0.000000
92	0.7018268E-01	0.000000
93	0.000000	0.000000

Zero Heating Cost

```
! THERMAL MEMBRANE DISTILLATION NETWORK SYNTHESIS
BINARY AQUEOUS MIXTURE ,LINEARIZED FLUX TEMPERATURE
RELATION , SHARP SPLIT OPERATION

The number of modules(sinks)  is N
the number of sinks and sources is  N.
Subscripts:  i is for sources  and  j is for sinks ;

Sets:

sources / 1..3/  ;
        sinks/ 1..3 / : normfact,f,z_in,gross_input,
gross_output,reject_bypass,temp,flux,permeate,theta,area,
deltat_feed,ysink ;
placed(sources, sinks) : w,deltat;
```

```
Endsets

!Objective Function;

Min= total_cost;

total_cost= heating_cost +  membrane_cost+ cooling_cost ;

heating_cost=@ABS(cost_heat * recycleheat);
!excess_heat=@IF(recycleheat #GT# 0, 0, @ABS(recycleheat) );
membrane_cost= (cost_area* totalarea);
cooling_cost=cost_cool*permeate_total*2260;
! pumping_cost=

! Parametric Data;

! feed massflow rate in kg/h;
feed= 3473;
temp_feed=350;

! Feed Concentration;
zfeed= .001;

! Feed Cp in kJ/ kg C;
cp=4.186;

!overall recovery is alpha;
alpha=.98;
permeate_total=alpha*feed;

! defining the nominal flux(lb/ ft^2 h ) and temperature(F);
temp_nom=124;
flux_nom=80;
psat_nom=97;

!Operating Cost: cost  of heating in $  per KJ ;
!cost_heat= .000006;
 cost_heat=0;
!ratio=cost_area/(cost_heat*1000000);

!Operating Cost: cost  of cooling in $  per lb ;
cost_cool=.000009;

! Fixed Cost: Cost of membrane heating in $  per metre ^2 of membrane ;
cost_area= 90;
!cost_area=0;
```

```
! slope of linearized temperature- flux curve  ( lb/ h F);
m= 3643;

!Nominal recovery at module;
theta_nom= 0.06;
s=329;

!Antoines Constants ;
a=18.3;
b=3816.44;
!c=-46.11;

! temperature drop across the module as afraction;
tempdrop=0.2;

! Generic  Model Equations;

! defining temperature dependency of flux;
@for(sinks(j): normfact(j)=@exp(a- (b/(temp(j)-46.11 ) ) )/psat_nom );
@for(sinks(j): flux(j)=flux_nom*normfact(j) );

! defining temperature dependancy of modular recovery;
@for(sinks(j): theta(j)= theta_nom*temp(j)/temp_nom );

! Accounting for output of module "j";
@FOR (sinks(j) : gross_output(j)=reject_bypass(j)+
@SUM( sources(i) :  w(i,j) ) );

! Accounting for input to module "j";
@FOR (  sinks (j) : gross_input(j)= f(j) + @sum(sources(i)  : w(i,j) ) );

! Overall Mass balance over module  " j ";
@FOR(sinks(j): permeate(j)=gross_input(j)-gross_output(j));
@FOR(sinks(j): permeate(j)= gross_input(j)*theta(j) );

! Component Mass Balance for module "j";
          @FOR( sinks(j) :gross_input(j)*z_in(j)= (  (f(j)*zfeed) +
@SUM(sources(i): ( w(i,j)*ysink(j)  )) ) );

! Overall Component Mass Balance ;
          (zfeed*feed_total) =( zpermeate*permeate_total)+ (
@SUM(sinks(j):reject_bypass(j)*ysink(j)) );

! Sharp Split Mass balance to relate input and output
concentrations to a module ;
```

```
@FOR( sinks(j) :ysink(j) = z_in(j)*gross_input(j)/gross_output(j) );

! Accounting for feed bypass;
feed= @SUM( sinks( j): f(j)) ;

! Accounting for reject bypass;
reject_output =  @SUM(sinks (j) : reject_bypass(j) ) ;

! Accounting for permeate;
permeate_total= (feed)-( reject_output) ;

! Double checking calculations ;
check=@SUM( sinks (j): permeate (j) )-permeate_total ;

! calculating total recycled reject;
@FOR(sources(i): @for(sinks(j): deltat(i,j)=temp(i)-temp(j) )  );
@for(sinks(j): deltat_feed(j)=temp(j)-temp_feed ) ;
recycleheat/cp= @sum(sinks(j): f(j)*deltat_feed(j) )+ @sum(sources(i):
@sum(sinks(j)  : w(i,j)*(deltat(i,j)+(temp(j)*tempdrop) ) ) );
recycled=@sum(sources(i):@sum(sinks(j) :w(i,j) ) );


! calculating area ;
@for( sinks(j):area(j)=permeate(j)/flux(j)) ;
totalarea=@sum(sinks(j):area(j));

@for(sinks(j):z_in(j)>0.00001 );
@for(sinks(j):z_in(j)<1 );

@for(sinks(j):temp(j)<366);
@for(sinks(j):temp(j)>200);

@for(sinks(j):ysink(j)<=.45);
@for(sinks(j):ysink(j)>=0);

z_reject_overall=@sum(sinks(j):reject_bypass(j)*ysink(j) );
```

```
Local optimal solution found at iteration:          251
Objective value:                                705.4461

            Variable           Value        Reduced Cost
          TOTAL_COST        705.4461            0.000000
        HEATING_COST        0.000000            0.000000
       MEMBRANE_COST        636.2181            0.000000
        COOLING_COST        69.22800            0.000000
```

COST_HEAT	0.000000	0.000000
RECYCLEHEAT	5057278.	0.000000
COST_AREA	90.00000	0.000000
TOTALAREA	7.069090	0.000000
COST_COOL	0.9000000E-05	0.000000
PERMEATE_TOTAL	3403.540	0.000000
FEED	3473.000	0.000000
TEMP_FEED	350.0000	0.000000
ZFEED	0.1000000E-02	0.000000
CP	4.186000	0.000000
ALPHA	0.9800000	0.000000
TEMP_NOM	124.0000	0.000000
FLUX_NOM	80.00000	0.000000
PSAT_NOM	97.00000	0.000000
M	3643.000	0.000000
THETA_NOM	0.6000000E-01	0.000000
S	329.0000	0.000000
A	18.30000	0.000000
B	3816.440	0.000000
TEMPDROP	0.2000000	0.000000
FEED_TOTAL	3473.480	0.000000
ZPERMEATE	0.000000	0.000000
REJECT_OUTPUT	69.46000	0.000000
CHECK	0.000000	0.1869284
RECYCLED	15745.53	0.000000
Z_REJECT_OVERALL	3.473480	0.000000
NORMFACT(1)	6.018349	0.000000
NORMFACT(2)	6.018349	0.000000
NORMFACT(3)	6.018349	0.000000
F(1)	3470.868	0.000000
F(2)	2.028243	0.000000
F(3)	0.1033304	0.000000
Z_IN(1)	0.4117563E-01	0.000000
Z_IN(2)	0.2322221E-01	0.000000
Z_IN(3)	0.6758778E-02	0.000000
GROSS_INPUT(1)	19206.97	0.000000
GROSS_INPUT(2)	11.04690	0.000000
GROSS_INPUT(3)	0.5124295	0.000000
GROSS_OUTPUT(1)	15805.48	0.000000
GROSS_OUTPUT(2)	9.090527	0.000000
GROSS_OUTPUT(3)	0.4216799	0.000000
REJECT_BYPASS(1)	69.37555	0.000000
REJECT_BYPASS(2)	0.7187290E-01	0.000000
REJECT_BYPASS(3)	0.1258081E-01	0.000000
TEMP(1)	366.0000	0.000000
TEMP(2)	366.0000	0.000000
TEMP(3)	366.0000	0.000000
FLUX(1)	481.4679	0.000000
FLUX(2)	481.4679	0.000000
FLUX(3)	481.4679	0.000000
PERMEATE(1)	3401.493	0.000000
PERMEATE(2)	1.956370	0.000000
PERMEATE(3)	0.9074961E-01	0.000000
THETA(1)	0.1770968	0.000000
THETA(2)	0.1770968	0.000000
THETA(3)	0.1770968	0.000000
AREA(1)	7.064838	0.000000

```
           AREA( 2)       0.4063344E-02        0.000000
           AREA( 3)       0.1884853E-03        0.000000
    DELTAT_FEED( 1)        16.00000            0.000000
    DELTAT_FEED( 2)        16.00000            0.000000
    DELTAT_FEED( 3)        16.00000            0.000000
          YSINK( 1)       0.5003706E-01        0.000000
          YSINK( 2)       0.2821985E-01        0.000000
          YSINK( 3)       0.8213333E-02        0.000000
           W( 1, 1)        2.007510            0.000000
           W( 1, 2)        4.328458            0.000000
           W( 1, 3)        0.000000            0.000000
           W( 2, 1)        15721.93            0.000000
           W( 2, 2)        3.943288            0.000000
           W( 2, 3)        0.000000            0.000000
           W( 3, 1)        12.16806            0.000000
           W( 3, 2)       0.7469081            0.000000
           W( 3, 3)       0.4090991            0.000000
      DELTAT( 1, 1)        0.000000            0.000000
      DELTAT( 1, 2)        0.000000            0.000000
      DELTAT( 1, 3)        0.000000            0.000000
      DELTAT( 2, 1)        0.000000            0.000000
      DELTAT( 2, 2)        0.000000            0.000000
      DELTAT( 2, 3)        0.000000            0.000000
      DELTAT( 3, 1)        0.000000            23.71389
      DELTAT( 3, 2)        0.000000           0.1363905E-01
      DELTAT( 3, 3)        0.000000            0.000000

                Row    Slack or Surplus      Dual Price
                  1        705.4461           -1.000000
                  2        0.000000           -1.000000
                  3        0.000000           -1.000000
                  4        0.000000           -1.000000
                  5        0.000000           -1.000000
                  6        0.000000          -0.2031230
                  7        0.000000            0.000000
                  8        0.000000            0.000000
                  9        0.000000            0.000000
                 10        0.000000           -719.8430
                 11        0.000000          -0.2072684
                 12        0.000000            0.000000
                 13        0.000000            7.952723
                 14        0.000000           -6.558930
                 15        0.000000           -5057278.
                 16        0.000000           -7692000.
                 17        0.000000           -7.069090
                 18        0.000000            0.000000
                 19        0.000000            0.000000
                 20        0.000000            0.000000
                 21        0.000000            636.2955
                 22        0.000000           -1.988864
                 23        0.000000            0.000000
                 24        0.000000            105.6494
                 25        0.000000           0.6076431E-01
                 26        0.000000           0.2818658E-02
                 27        0.000000            1.320618
                 28        0.000000           0.7595539E-03
                 29        0.000000           0.3523322E-04
```

30	0.000000	0.000000
31	0.000000	0.000000
32	0.000000	0.000000
33	0.000000	0.000000
34	0.000000	0.000000
35	0.000000	0.000000
36	0.000000	0.000000
37	0.000000	0.000000
38	0.000000	0.000000
39	0.000000	0.000000
40	0.000000	0.000000
41	0.000000	0.000000
42	0.000000	0.000000
43	0.000000	0.000000
44	0.000000	0.000000
45	-0.1835378E-06	0.000000
46	0.000000	0.000000
47	0.000000	0.000000
48	0.000000	0.000000
49	-0.3034834E-07	0.000000
50	0.000000	0.000000
51	0.000000	0.000000
52	0.000000	0.000000
53	0.000000	0.000000
54	0.000000	0.000000
55	0.000000	0.1869284
56	0.000000	0.000000
57	0.000000	0.000000
58	0.000000	0.000000
59	0.000000	0.000000
60	0.000000	0.000000
61	0.000000	0.000000
62	0.000000	23.71389
63	0.000000	0.1363905E-01
64	0.000000	0.000000
65	0.000000	0.000000
66	0.000000	0.000000
67	0.000000	0.000000
68	-0.2055895E-06	0.000000
69	0.000000	0.000000
70	0.000000	-90.00000
71	0.000000	-90.00000
72	0.000000	-90.00000
73	0.000000	-90.00000
74	0.4116563E-01	0.000000
75	0.2321221E-01	0.000000
76	0.6748778E-02	0.000000
77	0.9588244	0.000000
78	0.9767778	0.000000
79	0.9932412	0.000000
80	0.000000	0.000000
81	0.000000	0.000000
82	0.000000	23.72817
83	166.0000	0.000000
84	166.0000	0.000000
85	166.0000	0.000000
86	0.3999629	0.000000

```
87        0.4217801          0.000000
88        0.4417867          0.000000
89        0.5003706E-01      0.000000
90        0.2821985E-01      0.000000
91        0.8213333E-02      0.000000
92         0.000000          0.000000
```

Zero Membrane Cost

```
! THERMAL MEMBRANE DISTILLATION NETWORK SYNTHESIS
BINARY AQUEOUS MIXTURE ,LINEARIZED FLUX TEMPERATURE
RELATION , SHARP SPLIT OPERATION

The number of modules(sinks)  is N
the number of sinks and sources is  N.
Subscripts:  i is for sources  and  j is for sinks ;

Sets:

sources / 1..3/  ;
sinks/ 1..3 / : normfact,f,z_in,gross_input,
gross_output,reject_bypass,temp,flux,permeate,theta,area,deltat_feed,ysink ;
placed(sources, sinks) : w,deltat;

Endsets

!Objective Function;

Min= total_cost;

total_cost= heating_cost +  membrane_cost+ cooling_cost ;

heating_cost=@ABS(cost_heat * recycleheat);
!excess_heat=@IF(recycleheat #GT# 0, 0, @ABS(recycleheat) );
membrane_cost= (cost_area* totalarea);
cooling_cost=cost_cool*permeate_total*2260;
! pumping_cost=

! Parametric Data;

! feed massflow rate in kg/h;
feed= 3473;
temp_feed=350;

! Feed Concentration;
zfeed= .001;

! Feed Cp in kJ/ kg C;
```

```
cp=4.186;

!overall recovery is alpha;
alpha=.98;
permeate_total=alpha*feed;

! defining the nominal flux(lb/ ft^2 h ) and temperature(F);
temp_nom=124;
flux_nom=80;
psat_nom=97;

!Operating Cost: cost  of heating in $  per KJ ;
cost_heat= .000006;

ratio=cost_area/(cost_heat*1000000);

!Operating Cost: cost  of cooling in $  per lb ;
cost_cool=.000009;

! Fixed Cost: Cost of membrane heating in $  per metre ^2 of membrane ;
!cost_area= 90;
cost_area=0;

! slope of linearized temperature- flux curve  ( lb/ h F);
m= 3643;

!Nominal recovery at module;
theta_nom= 0.06;
s=329;

!Antoines Constants ;
a=18.3;
b=3816.44;
!c=-46.11;

! temperature drop across the module as afraction;
tempdrop=0.2;

! Generic  Model Equations;

! defining temperature dependency of flux;
@for(sinks(j): normfact(j)=@exp(a- (b/(temp(j)-46.11 ) ) )/psat_nom );
@for(sinks(j): flux(j)=flux_nom*normfact(j) );

! defining temperature dependancy of modular recovery;
@for(sinks(j): theta(j)= theta_nom*temp(j)/temp_nom );
```

```
! Accounting for output of module "j";
@FOR (sinks(j) : gross_output(j)=reject_bypass(j)+
@SUM( sources(i) :  w(i,j) ) );


! Accounting for input to module "j";
@FOR (  sinks (j) : gross_input(j)= f(j) + @sum(sources(i)  : w(i,j) ) );

! Overall Mass balance over module  " j ";
@FOR(sinks(j): permeate(j)=gross_input(j)-gross_output(j));
@FOR(sinks(j): permeate(j)= gross_input(j)*theta(j) );

! Component Mass Balance for module "j";
@FOR( sinks(j) :gross_input(j)*z_in(j)= (  (f(j)*zfeed) +
@SUM(sources(i): ( w(i,j)*ysink(j)  )) ) );

! Overall Component Mass Balance ;
(zfeed*feed_total) =( zpermeate*permeate_total)+ (
@SUM(sinks(j):reject_bypass(j)*ysink(j)) );

! Sharp Split Mass balance to relate input and output concentrations
 to a module ;
@FOR( sinks(j) :ysink(j) = z_in(j)*gross_input(j)/gross_output(j) );

! Accounting for feed bypass;
feed= @SUM( sinks( j): f(j)) ;

! Accounting for reject bypass;
reject_output =  @SUM(sinks (j) : reject_bypass(j) ) ;

! Accounting for permeate;
permeate_total= (feed)-( reject_output) ;

! Double checking calculations ;
check=@SUM( sinks (j): permeate (j) )-permeate_total ;

! calculating total recycled reject;
@FOR(sources(i): @for(sinks(j): deltat(i,j)=temp(i)-temp(j) )  );
@for(sinks(j): deltat_feed(j)=temp(j)-temp_feed ) ;
recycleheat/cp= @sum(sinks(j): f(j)*deltat_feed(j) )+ @sum(sources(i):
@sum(sinks(j)  : w(i,j)*(deltat(i,j)+(temp(j)*tempdrop) ) ) );
recycled=@sum(sources(i):@sum(sinks(j) :w(i,j) ) );

! calculating area ;
@for( sinks(j):area(j)=permeate(j)/flux(j)) ;
```

```
totalarea=@sum(sinks(j):area(j));

@for(sinks(j):z_in(j)>0.00001 );
@for(sinks(j):z_in(j)<1 );

@for(sinks(j):temp(j)<366);
@for(sinks(j):temp(j)>200);

@for(sinks(j):ysink(j)<=.45);
@for(sinks(j):ysink(j)>=0);

z_reject_overall=@sum(sinks(j):reject_bypass(j)*ysink(j) );
```

```
Local optimal solution found at iteration:           536
Objective value:                                98.45515
```

Variable	Value	Reduced Cost
TOTAL_COST	98.45515	0.000000
HEATING_COST	29.22715	0.000000
MEMBRANE_COST	0.000000	0.000000
COOLING_COST	69.22800	0.000000
COST_HEAT	0.6000000E-05	0.000000
RECYCLEHEAT	4871192.	0.000000
COST_AREA	0.000000	0.000000
TOTALAREA	13.24843	0.000000
COST_COOL	0.9000000E-05	0.000000
PERMEATE_TOTAL	3403.540	0.000000
FEED	3473.000	0.000000
TEMP_FEED	350.0000	0.000000
ZFEED	0.1000000E-02	0.000000
CP	4.186000	0.000000
ALPHA	0.9800000	0.000000
TEMP_NOM	124.0000	0.000000
FLUX_NOM	80.00000	0.000000
PSAT_NOM	97.00000	0.000000
RATIO	0.000000	0.000000
M	3643.000	0.000000
THETA_NOM	0.6000000E-01	0.000000
S	329.0000	0.000000
A	18.30000	0.000000
B	3816.440	0.000000
TEMPDROP	0.2000000	0.000000
FEED_TOTAL	3473.609	0.000000
ZPERMEATE	0.000000	0.000000
REJECT_OUTPUT	69.46000	0.000000
CHECK	0.000000	0.8623160E-02
RECYCLED	16624.09	0.000000
Z_REJECT_OVERALL	3.473609	0.000000
NORMFACT(1)	3.211268	0.000000
NORMFACT(2)	3.211268	0.000000
NORMFACT(3)	3.211268	0.000000
F(1)	0.000000	0.000000

```
               F( 2)        1763.862          0.000000
               F( 3)        1709.138          0.000000
            Z_IN( 1)       0.1000000E-04      0.000000
            Z_IN( 2)       0.3737903          0.000000
            Z_IN( 3)       0.2166894E-01      0.000000
     GROSS_INPUT( 1)       0.000000           0.000000
     GROSS_INPUT( 2)       10392.04           0.000000
     GROSS_INPUT( 3)       9705.052           0.000000
    GROSS_OUTPUT( 1)      0.5000000E-07       0.000000
    GROSS_OUTPUT( 2)       8632.099           0.000000
    GROSS_OUTPUT( 3)       8061.454           0.000000
   REJECT_BYPASS( 1)       0.000000           0.1038128E-01
   REJECT_BYPASS( 2)       3.919694           0.000000
   REJECT_BYPASS( 3)       65.54031           0.000000
            TEMP( 1)       350.0000           0.000000
            TEMP( 2)       350.0000           0.000000
            TEMP( 3)       350.0000           0.000000
            FLUX( 1)       256.9014           0.000000
            FLUX( 2)       256.9014           0.000000
            FLUX( 3)       256.9014           0.000000
        PERMEATE( 1)       0.000000           0.000000
        PERMEATE( 2)       1759.943           0.000000
        PERMEATE( 3)       1643.597           0.000000
           THETA( 1)      0.1693548           0.000000
           THETA( 2)      0.1693548           0.000000
           THETA( 3)      0.1693548           0.000000
            AREA( 1)       0.000000           0.000000
            AREA( 2)       6.850653           0.000000
            AREA( 3)       6.397775           0.2038581E-08
     DELTAT_FEED( 1)       0.000000           0.000000
     DELTAT_FEED( 2)       0.000000           0.2685565E-01
     DELTAT_FEED( 3)       0.000000           0.4292676E-01
           YSINK( 1)      0.4500000           0.000000
           YSINK( 2)      0.4500000           0.000000
           YSINK( 3)      0.2608695E-01       0.000000
            W( 1, 1)       0.000000           0.000000
            W( 1, 2)       2342.824           0.000000
            W( 1, 3)       2372.451           0.000000
            W( 2, 1)       0.000000           0.000000
            W( 2, 2)       3314.843           0.000000
            W( 2, 3)       3251.009           0.000000
            W( 3, 1)       0.000000           0.000000
            W( 3, 2)       2970.513           0.000000
            W( 3, 3)       2372.454           0.000000
       DELTAT( 1, 1)       0.000000           0.000000
       DELTAT( 1, 2)       0.000000           0.5025695E-01
       DELTAT( 1, 3)       0.000000           0.6817182E-01
       DELTAT( 2, 1)       0.000000           0.000000
       DELTAT( 2, 2)       0.000000           0.000000
       DELTAT( 2, 3)       0.000000           0.8165236E-01
       DELTAT( 3, 1)       0.000000           0.000000
       DELTAT( 3, 2)       0.000000           0.7460743E-01
       DELTAT( 3, 3)       0.000000           0.000000

                 Row    Slack or Surplus      Dual Price
                   1       98.45515           -1.000000
                 2       0.000000         -1.000000
```

3	0.000000	-1.000000
4	0.000000	-1.000000
5	0.000000	-1.000000
6	0.000000	-0.2834526E-01
7	0.000000	0.1744557E-01
8	0.000000	0.000000
9	0.000000	-6.981712
10	0.000000	-106.6852
11	0.000000	-0.3071847E-01
12	0.000000	-0.2849438
13	0.000000	0.000000
14	0.000000	0.000000
15	0.000000	-4871192.
16	0.000000	0.000000
17	0.000000	-7692000.
18	0.000000	-13.24843
19	0.000000	0.000000
20	0.000000	588.8850
21	0.000000	0.000000
22	0.000000	0.000000
23	0.000000	0.000000
24	0.000000	-146.1358
25	0.000000	0.000000
26	0.000000	0.000000
27	0.000000	0.000000
28	0.000000	0.000000
29	0.000000	0.000000
30	0.000000	0.000000
31	0.000000	0.000000
32	0.000000	107.8827
33	0.000000	100.7509
34	-0.5000000E-07	-0.8623160E-02
35	0.000000	0.1758120E-02
36	0.000000	0.1758120E-02
37	0.000000	-0.2312079E-04
38	0.000000	0.000000
39	0.000000	0.000000
40	-0.5000000E-07	0.8623160E-02
41	0.000000	-0.1758120E-02
42	0.000000	-0.1758120E-02
43	0.000000	0.000000
44	0.000000	0.1038128E-01
45	0.000000	0.1038128E-01
46	0.000000	0.2312079E-01
47	0.000000	0.000000
48	0.000000	0.000000
49	-0.2250000E-07	0.000000
50	-0.4500000	-0.4323025E-04
51	0.000000	0.000000
52	-0.7556886E-07	0.000000
53	0.000000	0.000000
54	-0.5000000E-07	-0.1758120E-02
55	0.000000	0.1758840E-02
56	-0.4999993E-07	0.8623160E-02
57	0.000000	0.000000
58	0.000000	-0.8585340E-02
59	0.000000	0.8585340E-02

60	0.000000	0.000000
61	0.000000	-0.8325558E-01
62	0.000000	0.000000
63	0.000000	0.000000
64	0.000000	0.000000
65	0.000000	-0.5958648E-01
66	0.000000	0.000000
67	0.000000	-0.1744557E-01
68	0.000000	0.000000
69	0.000000	-0.2511600E-04
70	0.000000	0.000000
71	0.000000	0.000000
72	0.000000	0.000000
73	0.000000	0.000000
74	0.000000	0.000000
75	0.000000	0.000000
76	0.3737803	0.000000
77	0.2165894E-01	0.000000
78	0.9999900	0.000000
79	0.6262097	0.000000
80	0.9783311	0.000000
81	16.00000	0.000000
82	16.00000	0.000000
83	16.00000	0.000000
84	150.0000	0.000000
85	150.0000	0.000000
86	150.0000	0.000000
87	0.000000	0.4323025E-04
88	0.000000	0.000000
89	0.4239130	0.000000
90	0.4500000	0.000000
91	0.4500000	0.000000
92	0.2608695E-01	0.000000
93	-0.2250000E-07	0.000000

APPENDIX B

DATA FOR DESALINATION CASE STUDY

Table 1

Desalination Case Study- Parametric Data

Parameters	Units	Value
Membrane Cost	$/ m^2	90
Heating Cost	$/kJ	0.000006
Permeate	lb/h	3403.58
Alpha	-	0.98
Cp	kJ/kg K	4.186
Fractional Module Temperature Drop	-	0.2
Nominal Temperature	K	324
Nominal Vapor Pressure	Pa	12418.88
Nominal separation factor	-	0.06
Nominal Flux	kg/m^2 h	80
A	-	23.238
B	K	3841
C	K	-46

Table 2

Desalination Case Study- Derived Data

Temp	K	350
Psat(T)		40260.72
Temp Scale Factor (T)	-	3.241896
Flux(T)	kg/ m ^2 h	259.3517
theta(T)	-	0.194514

Table 3

Desalination Case Study-Sensitivity of Objective Function

Case #	Description	Membrane Cost ($/m^2)	Heating Cost ($/MM BTU)	Area (m^2)	Module Temperature (K)	Objective Function ($)
1	General case	90	6	7.06	366 (max. temp)	73500
2	Only heating cost	0	6	13.24	350 (feed temp.)	9845
3	Only membrane cost	90	0	7.06	366 (max. temp)	70544

Table 4

Data for Dextrose Concentration Case Study

Parameters	**Units**	**Value**
Membrane Cost	$/ m^2	90
Heating Cost	$/kJ	0.000006
Permeate	lb/h	442.75
Cp	kJ/kg K	4.186
Fractional Module Temperature Drop	-	0.2
Nominal Temperature	K	324
Nominal Vapor Pressure	Pa	12418.88
Nominal separation factor	-	0.06
Nominal Flux	kg/m^2 h	80
A	-	23.238
B	K	3841
C	K	-46

VITA

Madhav Nyapathi Seshu was born on March 16th 1982 in Bangalore, India. He received Bachelor of Engineering degree in Chemical Engineering from RV College of Engineering, Bangalore, India in July 2003. He joined the graduate program at the Department of Chemical Engineering in Texas A&M University in August 2003 and completed his Masters degree in August 2005. While in Bangalore, Madhav worked on process design and modeling projects at the Indian Institute of Science, Bangalore and John Brown India. While at Texas A&M, he was involved in process systems related projects with NASA, the Department of Energy, Matrix Process Integration and PepsiCo.

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