A NOVEL FLASH BOOSTED MULTI-EFFECT DISTILLATION PROCESS

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Summary

Distillation is a kind of separation process which separates the component substances from a liquid mixture by selective evaporation and condensation phenomena. The outcomes of this process are both very low (sometimes nearly pure components) and high concentrate components. This procedure is very common in process industries such as desalination, refineries, food and beverage, and oil and gas.

In this project (thesis) a novel Flash Boosted Multi-Effect Distillation (FB-MED) process has been introduced. The FB-MED process is particularly effective when using low-grade sensible heat source, such as industrial waste heat or geothermal resources, with temperatures typically below 100°C. One of its main applications is in saline water distillation coupled with low grade sensible heat sources. Another application of this novel design is in the evaporation plant of alumina refineries to utilise their available low grade sensible heat resources for the aim of feed liquor evaporation to produce both freshwater and highly concentrated liquor.

Following the abovementioned applications, this process can also be customised for alumina refinery evaporation plants to utilise the available flashed vapour waste stream to reduce the live steam consumption, which is currently used as the heat source of the evaporation plant. This customised novel process, called Flash Boosted Thermal Vapour Compression Multi Effect Evaporation (FB-TVC-MEE), is introduced and explained.

This thesis includes 6 chapters as follow:

**Chapter 1** includes a brief explanation about the history of the desalination process, major types of desalination techniques and their pros and cons.

**Chapter 2** introduces low grade sensible heat driven distillation processes, particularly for saline water desalination applications. The conventional Multi Effect Distillation (MED), used as a benchmark process, and the Boosted MED (B-MED), which introduced a 22% improvement with respect to freshwater yield as compared to the conventional MED processes, have been explained.
Chapter 3 introduces the novel Flash Boosted MED process, which is the topic of this project. It includes a detailed explanation with mathematical simulations and thermo-economic analysis of this novelty.

Chapter 4 introduces the application of the FB-MED system in seawater desalination and compares its performance with the conventional MED and the Boosted MED (B-MED) processes. It shows up to a 43% and 20% increase in freshwater yield as compared with the conventional MED and B-MED processes, respectively. The net present value of FB-MED is 53% greater than the conventional MED's, showing the novel FB-MED process is both technically and economically superior to conventional MED and B-MED processes.

Chapter 5 introduces the application of Flash Boosted Multi-Effect Evaporation (FB-MEE) processes for alumina refinery applications. It shows that this process can improve both freshwater and re-concentration production rates by 35% as compared to the optimised conventional MEE process.

Chapter 6 introduces a novel Flash Boosted Thermal Vapour Compression Multi-Effect Evaporation (FB-TVC-MEE) process that utilises the available flashed vapour waste heat streams of the evaporation units to reduce live steam consumption for evaporation purposes in alumina refineries. This process can save 82% of the selected evaporation unit's available live steam. Its thermal performance increases by 98% against the optimised conventional TVC-MEE process.

This thesis has also resulted in one patent, four peer reviewed journal articles and two conference papers, as listed below.

1. Patent:

2. Published Articles:


3. Conference Papers:


# Table of Contents

**Summary** ............................................................................................................................ 3  
**Acknowledgement** ............................................................................................................ 17  
**Statement of candidate contribution** ................................................................................. 19  
**Nomenclature** .................................................................................................................. 25  
**Chapter 1 Introduction of Desalination** .............................................................................. 29  
  1.1 Introduction ....................................................................................................................... 30  
  1.2 A Brief History of Desalination ........................................................................................... 31  
  1.3 Desalination in Australia .................................................................................................... 33  
  1.4 Desalination Technologies ................................................................................................. 35  
     1.4.1 Processes with phase change ..................................................................................... 36  
     1.4.2 Processes without phase change ................................................................................ 44  
  1.5 Energy Consumption and Environmental Impacts ............................................................ 46  
**Chapter 2 Low Grade Sensible Heat Driven Distillation** ....................................................... 51  
  2.1 Low Grade Sensible Heat Sources ...................................................................................... 52  
  2.2 Conventional Multi Effect Distillation (MED) Process ....................................................... 54  
  2.3 Boosted Multi Effect Distillation (B-MED) Process ............................................................ 55  
  2.4 Process Simulation ............................................................................................................. 57  
**Chapter 3 Flash Boosted MED, the New Novel Design** ........................................................ 61  
  3.1 Introduction ....................................................................................................................... 62  
  3.2 Mathematical Simulation ................................................................................................... 63  
     3.2.1 The primary MED section ........................................................................................... 65  
     3.2.2 Flashing section ........................................................................................................... 78  
     3.2.3 Overall mass, salinity and energy balances ............................................................... 85  
     3.2.4 Solving procedure ....................................................................................................... 86  
  3.3 Pumping Power Analysis: ................................................................................................... 88  
     3.3.1 Saline water pump ...................................................................................................... 89  
     3.3.2 Brine recirculation pump ............................................................................................ 89  
     3.3.3 Distillate extraction and brine blowdown pumps ...................................................... 89  
     3.3.4 Heat source medium pump ......................................................................................... 90  
     3.3.5 Drain pump ............................................................................................................... 90  
     3.3.6 Make-up water pump ................................................................................................. 90  
     3.3.7 *NCG* extraction vacuum pump ................................................................................ 90
<table>
<thead>
<tr>
<th>Chapter 3.4 Thermo-Economic Analysis</th>
<th>91</th>
</tr>
</thead>
<tbody>
<tr>
<td>3.4.1 Capital cost</td>
<td>92</td>
</tr>
<tr>
<td>3.4.2 Operating costs</td>
<td>94</td>
</tr>
<tr>
<td>3.4.3 Cash flows and capital budgeting metrics</td>
<td>96</td>
</tr>
</tbody>
</table>

**Chapter 4 Application of the Flash Boosted MED Process in Seawater Desalination** 101

<table>
<thead>
<tr>
<th>4.1 Introduction</th>
<th>102</th>
</tr>
</thead>
<tbody>
<tr>
<td>4.2 Simulation results</td>
<td>102</td>
</tr>
<tr>
<td>4.3 Thermo-Economic analysis</td>
<td>112</td>
</tr>
<tr>
<td>4.3.1 Capital costs</td>
<td>113</td>
</tr>
<tr>
<td>4.3.2 Operating costs</td>
<td>114</td>
</tr>
<tr>
<td>4.3.3 Unit product cost</td>
<td>116</td>
</tr>
<tr>
<td>4.3.4 Net present value and incremental analysis</td>
<td>117</td>
</tr>
<tr>
<td>4.4 Conclusions</td>
<td>120</td>
</tr>
</tbody>
</table>

**Chapter 5 Application of the Flash Boosted MEE Process in Alumina Refinery Plant** 123

<table>
<thead>
<tr>
<th>5.1 Introduction</th>
<th>124</th>
</tr>
</thead>
<tbody>
<tr>
<td>5.2 Numerical Analysis and Validation</td>
<td>127</td>
</tr>
<tr>
<td>5.3 Results and Discussions</td>
<td>132</td>
</tr>
<tr>
<td>5.3.1 Production rate and waste heat performance ratio</td>
<td>132</td>
</tr>
<tr>
<td>5.3.2 Auxiliary pumping power</td>
<td>134</td>
</tr>
<tr>
<td>5.3.3 Capital cost analysis</td>
<td>136</td>
</tr>
<tr>
<td>5.4 Conclusion</td>
<td>138</td>
</tr>
</tbody>
</table>

**Chapter 6 A Novel Flash Boosted Thermal Vapour Compression MEE Process for Alumina Refineries** 141

<table>
<thead>
<tr>
<th>6.1. Introduction</th>
<th>142</th>
</tr>
</thead>
<tbody>
<tr>
<td>6.2. Process Description</td>
<td>145</td>
</tr>
<tr>
<td>6.3. Mathematical Simulation and Capital Cost Analysis</td>
<td>147</td>
</tr>
<tr>
<td>6.3.1 Technical simulation</td>
<td>147</td>
</tr>
<tr>
<td>6.3.2 Capital cost analysis</td>
<td>154</td>
</tr>
<tr>
<td>6.4. Results and Discussion</td>
<td>156</td>
</tr>
<tr>
<td>6.5. Conclusion</td>
<td>160</td>
</tr>
</tbody>
</table>

**Bibliography** 161

**Appendices** 171

- Appendix A. Enthalpies 173
- Appendix B. BPE and NEA 175
- Appendix C. Overall Heat Transfer Coefficient 177
List of Figures

Figure 1-1 Major Desalination Processes ................................................................. 35

Figure 1-2 Standard (Conventional) MSF schematic design [27]............................... 37

Figure 1-3 One of the 40 MSF units of Al-Jubail Desalination Plant – Kurt Kiefer [64] ......... 38

Figure 1-4 Standard (Conventional) MED schematic design for seawater application with horizontal falling film evaporator ................................................................. 39

Figure 1-5 A Multi Effect Plate Distiller plant [68] ...................................................... 41

Figure 1-6 Schematic design of a Thermal Vapour Compression Multi Effect Distillation (TVC-MED) process for seawater application with horizontal falling film evaporators .......................... 42

Figure 1-7 Schematic design of a Mechanical Vapour Compression Multi Effect Distillation (MVC-MED) process for seawater application with horizontal falling film evaporators .......... 42

Figure 1-8 Tobruk MED-TVC desalination plant in Libya, 40 ML/day capacity, Completion date: 2000 [72] ......................................................................................................................... 43

Figure 1-9 Osmosis and reverse osmosis flows .......................................................... 44

Figure 1-10 Schematic design of typical RO plant ....................................................... 46

Figure 1-11 Theoretical minimum energy consumption for desalting pure NaCl solution. The highlighted range is the typical recovery range for an SWRO plant. [32] ......................................... 48

Figure 2-1 Schematic design of a 4-effect conventional MED process coupled with a sensible heat source ............................................................................................................. 54

Figure 2-2 Schematic design of a 4-Effect preheated feed MED process coupled with a sensible heat source ........................................................................................................... 55

Figure 2-3 Schematic design of a Boosted MED (B-MED) system ................................. 56

Figure 2-4 The two serially connected Alfa Laval's single-effect rising film plate evaporators modules (JWP-16-C Series) [110] ..................................................................... 56

Figure 2-5 Schematic design of an optimised conventional MED system (typical quantities as per the simulation for 65°C inlet heat source temperature) [3] ........................................... 60

Figure 2-6 Schematic design of an optimised Boosted MED (B-MED) system (typical quantities as per the simulation for 65°C inlet heat source temperature) [3] ........................................... 60

Figure 3-1 Schematic design of the Flash Boosted MED (FB-MED) system, includes 4-effects primary MED boosted by 3 flashing chambers ......................................................... 63

Figure 3-2 Schematic design of the first effect of the primary MED section .................. 66
Figure 3-3 Temperature-energy profile for the first effect of the primary MED section of the FB-MED system ................................................................. 68

Figure 3-4 Schematic design of the effects of the primary MED section (Except the first effect) without vapour injection from the flashing chambers \( k \in \{2, \ldots, n\} \) .............................................. 70

Figure 3-5 Schematic design of the effects of the primary MED section (Except the first effect) with vapour injection from the flashing chambers \( k \in \{2, \ldots, n\}; i \in \{1, \ldots, j\} \) ......................................... 71

Figure 3-6 (A) Temperature-energy profile for the second to the last effect of the primary MED section of the FB-MED system without flashed feed stream (B) Temperature-energy profile for the second to the last effect of the primary MED section of the FB-MED system with flashed feed stream .......................................................................................... 73

Figure 3-7 Schematic design of the condenser with vapour injection ........................................ 76

Figure 3-8 Temperature-energy profile for the condenser of the FB-MED system .................... 77

Figure 3-9 Schematic design of the flashing section ................................................................ 79

Figure 3-10 Schematic design of the liquid-liquid heat exchanger ........................................... 80

Figure 3-11 Schematic design of the flashing chamber \( i \in \{1, \ldots, j\} \) ........................................ 81

Figure 3-12 Schematic design of the recycling process ............................................................. 83

Figure 3-13 Schematic design of the deaerator ......................................................................... 85

Figure 3-14 Schematic design of the FB-MED process ............................................................ 86

Figure 3-15 The relevant flow chart of the FB-MED process simulation .................................... 88

Figure 3-16 MED and MSF plant capital costs as a function of daily production capacity [3], [55]. Dashed lines have been added to bound the region of high density upper limit of data points. ........................................................................................................... 93

Figure 4-1 Schematic design of an optimised conventional MED system (typical quantities as per the simulation for 65°C inlet heat source temperature) [3] ........................................................................ 103

Figure 4-2 Schematic design of an optimised Boosted MED (B-MED) system (typical quantities as per the simulation for 65°C inlet heat source temperature) [3] .................................................. 104

Figure 4-3 Schematic design of an optimised Flash Boosted MED (FB-MED) system, (typical quantities as per the simulation for 65°C inlet heat source temperature) [3] .............................. 105

Figure 4-4 Freshwater production rate comparison among the three systems for an assortment of heat source temperatures. The percentages on top of both the boosted and flash boosted MED columns represent the increment of freshwater production over and above the optimised conventional MED process ............................................................................................. 107

Figure 4-5 Waste heat performance comparison among the three systems and the percentage of increment of boosted and flash boosted MED's performance ratios compare to the optimised conventional MED for an assortment of heat source temperatures .................................................................................. 108
Figure 4-6  Temperature profile for the 1st effect of the conventional MED, the booster and the 1st effect of the B-MED and both the 1st effect and the liquid-liquid heat exchanger of the FB-MED system for 75°C heat source inlet temperature configuration .................................................. 110

Figure 4-7  Energy released of each effect of all three systems for 75°C heat source inlet temperature configuration ........................................................................................................ 111

Figure 4-8  $UA$-value of each effect of all three systems for 75°C heat source inlet temperature configuration ............................................................................................................................. 111

Figure 4-9  Specific pumping power (kWh/m$^3$) vs heat source inlet temperature ($^\circ$C) of the optimised conventional MED, B-MED and the FB-MED processes (The values are indicative of the percentage ratio as compared to the conventional MED alternative for each specific heat source temperature) ................................................................................................................. 112

Figure 4-10  A comparison of the normalised capital costs of optimised MED, B-MED and FB-MED plants for each considered heat source temperature. The values are indicative of the percentage decrease in the specific capital cost of the B-MED or FB-MED, as compared to the conventional MED alternative for each specific heat source temperature. [3] ............................................................................................................. 114

Figure 4-11  Breakdown of annual operating costs vs. heat source fluid inlet temperature of the optimised conventional MED, B-MED and FB-MED processes. The percentage increase in total annual operating cost over the MED alternative is noted for each B-MED and FB-MED for each heat source temperature. [3] ............................................................................................................................. 116

Figure 4-12  Specific annual operational costs of optimised MED, B-MED and FB-MED plants for each considered heat source temperature. The values above each column are indicative of the percentage difference in specific operating cost of the B-MED, or FB-MED, to the conventional MED alternative for each specific heat source temperature. [3] ............................................................................................................. 116

Figure 4-13  $UPC$ of the optimised MED, B-MED and FB-MED plants for each considered heat source temperature. The values above each column are indicative of the percentage difference in $UPC$ of the B-MED, or FB-MED, to the conventional MED alternative for each specific heat source temperature. [3] ............................................................................................................................. 117

Figure 4-14  Net Present Value ($NPV$) vs. interest rate plots for the optimised conventional MED, B-MED and FB-MED plant designs for each considered heat source temperature. [3] .............. 118

Figure 4-15  Net Present Value ($NPV$) vs. interest rate plots for the difference between FB-MED and B-MED for all the relevant heat source temperatures. It also highlights the incremental internal rate of return, $\Delta IRR$. [3].............................................................................................. 119

Figure 4-16  $NPV$ vs. number of installed flashing chambers for all considered temperatures of the heat source fluid. Note that the plant designs utilising zero flashing chambers are conventional MED plants, whilst all remaining plant designs are FB-MED. The percentage values above some data points indicate the percentage increase in $NPV$ for that FB-MED design compared with the conventional MED alternative (0 number of flashing chamber) for that particular heat source temperature. [3]................................................................. 119

Figure 4-17  Net present value of the optimised MED, B-MED and FB-MED plants for each considered heat source temperature. The values above each column are indicative of the
greatest percentage increase in $NPV$ of the B-MED and FB-MED over the conventional MED alternative for a specific heat source temperature. [3]

**Figure 5-1** The two serially connected Alfa Laval's single-effect rising film plate evaporators modules (JWP-16-C Series) [110]

**Figure 5-2** Schematic design of a Multi Effect Evaporation (MEE) unit in alumina refineries

**Figure 5-3** Schematic design of a Multi Stage Flash (MSF) evaporation unit in alumina refineries

**Figure 5-4** Schematic design of a Flash Boosted Multi Effect Evaporation (FB-MEE) unit (parallel feed) (Three falling film effects and two flash vessels) coupled with an 85°C low grade heat source [98]

**Figure 5-5** Schematic of the temperature-energy profiles of the conventional MEE process for 85°C heat source inlet temperature

**Figure 5-6** Schematic of the temperature-energy profiles of the FB-MEE process for 85°C heat source inlet temperature

**Figure 5-7** Schematic of a conventional MEE process coupled to an 85°C low grade heat source

**Figure 5-8** Freshwater and re-concentrated process liquor production rates (Left) and waste heat performance ratio for both FB-MEE and conventional MEE systems (Right) over an assortment of heat source temperatures

**Figure 5-9** Specific pumping power (normalised by the total freshwater production rate) breakdown for both conventional MEE and FB-MEE

**Figure 5-10** Capital cost and specific capital cost ($/(m^3/day)) ratios

**Figure 6-1** The two serially connected Alfa Laval's single-effect rising film plate evaporator modules (JWP-16-C Series) [110]

**Figure 6-2** Schematic design of a Multi Stage Flash (MSF) evaporation unit in alumina refineries

**Figure 6-3** Schematic design of a Multi Effect Evaporation (MEE) unit in alumina refineries

**Figure 6-4** Schematic design of the conventional TVC-MEE process coupled with the available waste stream (as the benchmark) of the considered alumina refinery plant. Other than plant specifications, all figures are based on our simulation

**Figure 6-5** Schematic design of the FB-TVC-MEE process coupled with the available waste stream of the considered alumina refinery plant. Other than plant specifications, all figures are based on our simulations

**Figure 6-6** Schematic of the temperature-energy profiles for the first effect, second effect and the condenser of the conventional TVC-MEE
**Figure 6-7** Schematic of the temperature-energy profiles for the first effect, second effect and the condenser of the FB-TVC-MEE. All figures are based on our simulation. .......................... 153

**Figure 6-8** Thermo-compressor entrainment chart [176] ........................................................ 153

**Figure 6-9** Live steam saving and percentage recovery of flashed vapour waste stream for TVC-MEE and FB-TVC-MEE processes, benchmarked against extant MSF process. ......................... 157

**Figure 6-10** Optimisation trajectory as the process evolves from an optimised TVC-MEE to an optimised FB-TVC-MEE. ............................................................................................................ 158

**Figure 6-11** Gain Output Ratio (GOR) of the current MSF, TVC-MEE and FB-TVC-MEE processes .................................................................................................................................................. 159

**Figure 6-12** Specific pumping powers for the two proposed evaporation processes .......... 160

**Figure B 1** Boiling Point Elevation (BPE) of aqueous solutions [1]................................................. 175
List of Tables

**Table 2-1** Assumptions of thermodynamic simulations [3] ........................................................ 59

**Table 2-2** Main flowrates for optimised MED and B-MED over the considered heat source temperature range [3] ........................................................................................................... 59

**Table 3-1** The thermodynamic assumed parameters for the solving procedure.................. 87

**Table 4-1** Assumptions of thermodynamic simulations [3] ...................................................... 102

**Table 4-2** Production rates and power consumption of the Conventional, Boosted and Flash Boosted MED processes [3] .................................................................................................................. 106

**Table 4-3** Main flowrates for MED, B-MED and FB-MED over the considered heat source temperature range [3] .............................................................................................................. 107

**Table 4-4** Assumptions of our economic analyses [3] .............................................................. 113

**Table 4-5** Thermo-economic simulation results for MED, B-MED and FB-MED over the considered heat source temperature range [3] ........................................................................ 113

**Table 5-1** Assumptions of thermodynamic simulations [98] ................................................ 130

**Table 5-2** Area coefficients consistent with the degree of superheat in each effect for 85°C heat source inlet temperature processes ......................................................................................... 137

**Table 6-1** Assumed specifications ............................................................................................. 145

**Table 6-2** Assumptions and boundary conditions of our thermodynamic simulation .......... 148

**Table 6-3** Available live steam, steam consumptions and recovered flashed vapour for the considered processes .................................................................................................................. 156

**Table 6-4** Pump specifications .................................................................................................. 159

**Table 6-5** Degree of superheat in each effect .......................................................................... 159

**Table C 1** Heat transfer coefficients in kW/(m²·K) for MEE-TVC and FB-MEE-TVC processes. 178
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Finally, a special thanks to my brother, Milad Lagevardi, for editing my thesis.
Statement of candidate contribution

I certify that, except where specific reference is made in the text to the work of others, the contents of this thesis are original and have not been submitted to any other university.

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1. This thesis does not contain work that I have published, nor work under review for publication.
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My publications are listed as below:

1. Patent: (My contribution is 80%)

2. First Paper: (My contribution is 80%)

   Chapters 3 and 4 include some contents of this article.

3. Second Paper: (My contribution is 80%)

Chapters 2, 3 and 4 include some contents of this article.

4. Third Paper: (My contribution is 80%)


Chapter 5 is identical to this article.

5. Fourth Paper: (My contribution is 80%)


Chapter 6 is identical to this article.

6. First Conference Paper: (My contribution is 80%)


Chapter 5 is identical to this conference paper.

7. Second Conference Paper: (My contribution is 80%)


Chapters 2, 3 and 4 includes some contents of this article.
In accordance with regulations of the University of Western Australia, this thesis, which is somewhere between the old style and the series of papers style, includes six chapters. As mentioned at the beginning of each chapter, Chapters 3 and 4 include my published papers in Desalination Journal and my conference paper in IDA 2015 World Congress on Desalination and Water Reuse, San Diego, California, USA, August 30-September 4, 2015. Chapter 5 is identical to my conference paper in 10th International Alumina Quality (AQW) Workshop, Perth, Australia, 19th-23rd April 2015, 327-336 and my published paper in Hydrometallurgy Journal. Chapter 6 is identical to my published article in Applied Thermal Engineering Journal.

My publications and contributions are listed as following:

1. **Patent:**


The core idea of this patent is mine, however without assistance of the co-authors and the Office of Innovation and Industry (OII), I could not develop it. At the beginning of my PhD, I brought about this idea and simulated the process and proved that this idea really works. Prof. Chua assisted me in all steps to develop and prepare this idea as a patent. I have also used the valued comments of both co-authors which could bring about a substantial development to this work. In the final year of my PhD, I brought about another novel process which is used in mineral refineries (see Chapter 6). Again I have a special thanks to Prof. Chua for his efforts and valued comments regarding to the second novelty. My contribution to this patent, considering the second novelty, is 80%.

2. **First Paper:**


This is my first paper introducing the application of the first novelty in seawater desalination. I prepared the simulation (including both technical and thermo-economic) and results, and wrote the first draft of the manuscript, including all graphs and tables. Prof. Chua polished the article and his valued comments could help me to have a precious article. I also considered the other co-authors’ valued comments which could further develop my work. My contribution to this work is 80%.

3. **Second Paper:**

This is my second paper comparing the first novelty with the conventional and previous novel methods with respect to thermo-economic issues, proving the economic potential of our novelty over other relevant methods. I prepared my analytical thermo-economic estimation method, simulation and results. I also prepared all relevant graphs and tables. Mr. May prepared the first draft. Prof. Chua polished the manuscript. I also considered his valued comments regarding to the content of this article. My contribution to this paper is 80%.

4. Third Paper:


This is related to the application of the first novelty into the alumina refinery’s evaporation plant. I prepared all simulations and results, and wrote the manuscript including all graphs and tables. Prof. Chua polished the article. I also applied his valued comments which could substantially develop the content of this article. The last three co-authors supplied the initial data of the considered alumina refinery evaporation process. My contribution to this paper is 80%.

5. Fourth Paper:


This is related to the second novelty which is customised for alumina refineries. This article is based on a real world application in one of the largest alumina refineries in Western Australia. I did all simulations, prepared the results and wrote the first draft of the manuscript. Prof. Chua polished the manuscript, and his valued comments could help me to add a very precious graph to this article. He also helped me to amend some essential graphs. The last three co-authors supplied the initial data of the abovementioned alumina refinery evaporation process. My contribution to this paper is 80%.

6. First Conference Paper:


This conference paper is related to the application of our first novelty into the alumina refinery’s evaporation plant. Prof. Chua was the presenter. My contribution to this conference paper is 80%.
7. Second Conference Paper:


This conference paper is related to the application of our first novelty into the seawater desalination process. Prof. Chua is the presenter. My contribution to this conference paper is 80%.
<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>$A$</td>
<td>Evaporator area (m$^2$)</td>
</tr>
<tr>
<td>$AEC$</td>
<td>Annual Electrical Cost (US$/yr)</td>
</tr>
<tr>
<td>$a$</td>
<td>Constant</td>
</tr>
<tr>
<td>$BPE$</td>
<td>Boiling Point Elevation (°C)</td>
</tr>
<tr>
<td>$CC$</td>
<td>Capital Cost (US$)</td>
</tr>
<tr>
<td>$CMP$</td>
<td>Compression Ratio</td>
</tr>
<tr>
<td>$CRF$</td>
<td>Capital Recovery Factor</td>
</tr>
<tr>
<td>$D_t$</td>
<td>Total process condensate production rate (m$^3$/day)</td>
</tr>
<tr>
<td>$ENT$</td>
<td>Entrainment Ratio</td>
</tr>
<tr>
<td>$EUP$</td>
<td>Electricity Unit Price (US$/kWh)</td>
</tr>
<tr>
<td>$EXP$</td>
<td>Expansion Ratio</td>
</tr>
<tr>
<td>$F$</td>
<td>Percentage of desalination plant capital cost covered by loan (%)</td>
</tr>
<tr>
<td>$f$</td>
<td>Plant availability</td>
</tr>
<tr>
<td>$g$</td>
<td>Acceleration of gravity (m/s$^2$)</td>
</tr>
<tr>
<td>$GOR$</td>
<td>Gain Output Ratio</td>
</tr>
<tr>
<td>$\Delta H$</td>
<td>Heat Transfer Rate (MW)</td>
</tr>
<tr>
<td>$\Delta h_{\text{avail.}}$</td>
<td>Maximum exploitable energy of the heat source (kJ/kg)</td>
</tr>
<tr>
<td>$\Delta h_{\text{ref}}$</td>
<td>Specific reference enthalpy of the distillate (kJ/kg)</td>
</tr>
<tr>
<td>$H$</td>
<td>Brine level (m)</td>
</tr>
<tr>
<td>$h$</td>
<td>Enthalpy (kJ/kg)</td>
</tr>
<tr>
<td>$h_f$</td>
<td>Enthalpy of saturated liquid (kJ/kg)</td>
</tr>
<tr>
<td>$h_{fg}$</td>
<td>Enthalpy of evaporation/condensation (latent heat) (kJ/kg)</td>
</tr>
<tr>
<td>$h_g$</td>
<td>Enthalpy of steam (kJ/kg)</td>
</tr>
<tr>
<td>$\Delta IRR$</td>
<td>Incremental Rate of Return (%), the value of interest rate which yields an $NPV$ equal to zero</td>
</tr>
<tr>
<td>$Income_1$</td>
<td>Income of the desalination plant in the first year</td>
</tr>
<tr>
<td>$IRR$</td>
<td>Internal Rate of Return</td>
</tr>
<tr>
<td>$i$</td>
<td>Annual interest rate</td>
</tr>
<tr>
<td>$k$</td>
<td>Thermal Conductivity (kW/(m·K))</td>
</tr>
<tr>
<td>$L$</td>
<td>Flash Stage Length, (m)</td>
</tr>
<tr>
<td>$M_{D,\text{total}}$</td>
<td>Total process condensate mass flow rate (kg/s)</td>
</tr>
<tr>
<td>$M_{HC,\text{total}}$</td>
<td>Total high condensate liquor mass flow rate (kg/s)</td>
</tr>
<tr>
<td>$MARR$</td>
<td>Minimum Acceptable Rate of Return</td>
</tr>
<tr>
<td>$m$</td>
<td>Mass flow rate (kg/s)</td>
</tr>
<tr>
<td>$NCG$</td>
<td>Non-Condensable Gas</td>
</tr>
<tr>
<td>$NEA$</td>
<td>Non-Equilibrium Allowance (°C)</td>
</tr>
<tr>
<td>$NEA_{10}$</td>
<td>$NEA$ for a 10-ft flashing stage (°C)</td>
</tr>
<tr>
<td>$NPV$</td>
<td>Net Present Value ($)</td>
</tr>
</tbody>
</table>
\( n \)  
Desalination Plant Operating Life

\( OPEX \)  
Operating Expenditure ($/year)

\( \Delta P \)  
Pressure Difference (kPa)

\( P \)  
Pressure (bar)

\( Pr \)  
Prandtl number

\( PR \)  
Performance ratio

\( PR_{WH} \)  
Waste heat performance ratio

\( p \)  
Specific pumping power (kWh/m\(^3\))

\( \dot{Q} \)  
Heat transfer rate (kW)

\( R \)  
Feed to vapour ratio

\( RF \)  
Recovery Factor

\( R^2 \)  
R-squared value

\( R_F \)  
Fouling resistance (m\(^2\).K/kW)

\( Re \)  
Reynolds number

\( r \)  
Evaporator tube radius, (m)

\( S \)  
Supplementary fee on loan as a percentage of the loan amount

\( SCC \)  
Specific Capital Cost ($/(m\(^3\)/day))

\( \Delta T \)  
Temperature difference (°C)

\( T \)  
Temperature (°C)

\( TBT \)  
Top Brine Temperature (°C)

\( TCC \)  
Total Capital Cost ($)

\( U \)  
Overall Heat Transfer Coefficient (kW/(m\(^2\).K))

\( UA \)  
\( UA \) Value (kW/K)

\( UPC \)  
Unit Product Cost ($/m\(^3\))

\( V \)  
Volumetric flowrate (m\(^3\)/s)

\( W \)  
Flash Stage Width (m)

\( WMP \)  
The market price of water per cubic meter ($/m\(^3\))

\( w \)  
Specific work (kWh/m\(^3\))

\( X \)  
Concentration (wt%)

\( x \)  
Steam quality

\( yr \)  
Year of plant operation

\( \alpha \)  
Film heat transfer coefficient (kW/(m\(^2\).K))

\( \alpha^+ \)  
Dimensionless heat transfer coefficient (kW/(m\(^2\) K))

\( \beta \)  
Local evaporator coefficient factor

\( \eta \)  
Overall pump efficiency

\( \theta \)  
Flag distinguishing FB-TVC-MEE from TVC-MEE

\( \lambda \)  
Constant

\( \mu \)  
Viscosity (Pa.s)

\( \rho \)  
Density (kg/m\(^3\))

\( \sigma \)  
Constant

\( \nu \)  
Kinematic Viscosity (m\(^2\)/s)

\( \phi \)  
Area coefficient

\( Y \)  
Overall area coefficient
Cost function

**Subscripts**

1. first effect
2. second effect
3. third effect

- **B**
  - highly concentrated outlet or Brine
- **B**
  - relevant boiling point
- **B**
  - Boosted MED process
- **C**
  - coolant (cooling water supply)
- **c_hex**
  - cold side of the liquid-liquid heat exchanger
- **calc.**
  - calculated
- **cond.**
  - condenser
- **D**
  - distillate
- **DR**
  - drain ratio
- **D_{Booster}**
  - total generated vapour from the booster effect (m³/day)
- **D_{FC}**
  - total flashed vapour production rate from flash chambers (m³/day)
- **D_{FV}**
  - total flashed vapour production rate from flash vessels (m³/day)
- **D_{inj}**
  - injected vapour flowrate (m³/day)
- **D_t**
  - total process condensate production rate (m³/day)
- **d**
  - discharge steam
- **e**
  - effects
- **F**
  - feed
- **FB**
  - Flash Boosted MED process
- **FBMEE**
  - Flash Boosted MEE process
- **FB**
  - Flash Boosted TVC-MEE process
- **FC**
  - Flashing Chamber
- **HC**
  - highly concentrated process liquor
- **HS**
  - heat source (for effects), hot side (for the condenser)
- **h_hex**
  - hot side of the liquid-liquid heat exchanger
- **hex**
  - liquid-liquid heat exchanger
- **i**
  - flashing chamber number
- **in**
  - inlet
- **inj**
  - injected
- **inn**
  - inner surface
- **irr**
  - irreversible
- **j**
  - total number of flashing stages
- **k**
  - MEE effect’s number
- **k**
  - MEE effect’s number which has been injected by flashed vapour
- **L**
  - liquid
- **lm**
  - log mean
- **MED**
  - Multi Effect Distillation process
- **MK**
  - make-Up
- **MSF**
  - Multi Stage Flash process
- **m**
  - motive steam
- **n**
  - total number of MED effects
- **o**
  - outer surface
- **out**
  - outlet
- **P**
  - effect’s pressure
- **P_1**
  - first effect’s pressure
- **P_2**
  - second effect’s pressure
- **P_d**
  - discharge pressure
- **P_i**
  - iᵗʰ flashing chamber’s pressure
- **P_k**
  - kᵗʰ effect’s pressure
- **P_n**
  - nᵗʰ effect’s pressure
- **p**
  - process condensate
pinch point
recycled brine
reversible
suction steam
saturated steam
stage
seawater
sensible-latent heat transfer
sensible-sensible heat transfer
total
vapour
injected vapour (flashed vapour from flashing chambers)
saturated vapour
pure water
zone 1 of the temperature-energy profile
zone 2 of the temperature-energy profile
zone 3 of the temperature-energy profile
Chapter 1

Introduction of Desalination

Desalination is the process of removing dissolved salts and minerals from saline water (such as seawater, brackish water and etc.). The main goal of the desalination process is to produce freshwater for the purpose of drinking, irrigation or industrial usage. However, the other outcome of this process is brine (sometimes called highly concentrated feed) which is mostly rejected to the environment in the seawater desalination application. In some industrial applications, such as alumina refineries, both of the high concentrate outlets of the evaporation unit (as an equivalent of thermal distillation process) and the freshwater production are valuable and reusable in the process (see Chapters 5 and 6).

This chapter includes a brief explanation about the history of desalination processes, the world’s water shortage and the importance of the desalination process to overcome this problem. Major types of desalination techniques and their pros and cons, the energy consumption and the environmental impacts of desalination processes have also been explained in this chapter.

Highlights:
- Introduce desalination and its importance in human life
- Review the history of desalination
- Summarizes the major desalination techniques and their pros and cons
- Energy consumption of desalination processes
- CO₂ emission and other environmental impacts of desalination processes

Publication:
A minor portion of this chapter has been extracted from the introduction sections of my below publications [2]–[4]:

Chapter 1. Introduction of Desalination

1.1 Introduction

Nearly 71% of the surface of the earth \( (510 \times 10^6 \text{ km}^2) \) is covered by the oceans and the remaining 29% is occupied by the lands [5]. There is certainly bountiful water available on earth, but only 3% of which is drinkable and 97% is saltwater [6], but nearly 70% of this available freshwater is frozen in glaciers, while the remaining 30% is in underground hard-to-reach aquifers, for which approximately 0.25% flows in rivers and lakes for direct use [7]. Therefore, traditional sources of available freshwater such as underground aquifers and surface water constitute a limited quantity worldwide. Furthermore, depletion of these sources are increasingly at an alarming rate [8].

Water scarcity is the mismatch of the demand and availability of freshwater in a particular location. It has become a worldwide issue with the pollution of existing water supplies, increasing population and industry activity, uneven freshwater to population distributions and changing rainfall patterns mean that many regions containing populated centres are becoming less capable of meeting the water supply requirements of the residing populations [7], [9]–[11]. Water-stressed countries currently encompass one-third of the world’s population and it is predicted to reach two-thirds by 2025 [12]. Moreover residential and industrial water shortage, agriculture also is being affected directly by water shortage. Farmers increasingly have to compete for water with urban residents and industries and it puts the global food security at risk [13].

Methods of attenuating such water supply issues include wastewater treatment and reuse, desalination, as well as water conservation schemes. Some eighty countries face severe water shortages [14], whilst some countries such as Kuwait, the United Arab Emirates and Saudi Arabia currently depend almost entirely on desalination for supply of water [8].

As a result of all above mentioned, seawater desalination has become an essential option to augment freshwater resources, especially in developing countries and many arid zones. As an example, the Gulf Cooperation Council (GCC) countries (in the Middle East region) produce around 39% of the worlds desalinated water production [15], [16].
Chapter 1. Introduction of Desalination

Desalination process is being increasingly adopted over traditional water supply methods because of the cost per unit volume water produced has come down for desalination while it has risen for traditional methods [8], [11], [17]. In 2011, approximately 150 countries worldwide used around 15,988 desalination plants (these include online, under construction and contracted) to produce desalinated waters [18]. The total global capacity of all online plants was 70.8 Mm$^3$/day in 2011 [19], which had a 10% increase in comparison with that in 2010. Also, 632 new plants have been added from mid-2011 to Aug-2012, which has increased the installed capacity to 74.8 Mm$^3$/day [19]. This data indicates the potential of the desalination market in both aspects of freshwater production rate and energy consumption.

1.2 A Brief History of Desalination

Word origin and history for desalination is dated back to 1943, and as a verb, ‘desalt’ was recorded in 1909 [20]; but the concept of desalination is much older, with references to it being found in ancient writings [21]. Historically, salt has been held as a precious commodity. It was found that the first goal of desalination was not related to producing fresh water, but rather to extract and use the salt from salty water by means of natural evaporation [6].

It is difficult to find the first time when humans desalinated salty water for producing freshwater, but Aristotle (384-322 BC) is one of the earliest recorded scientists who explained the desalination process. His understanding was based on experiments about when salt water turns into vapour, the condensed vapour did not carry on any salt [22]. In that time, the needs for producing fresh water for sailors was necessary in long distance trips. Ancient drawings show sailors boiled seawater and suspended large sponges from the mouth of a brass vessel to absorb what is evaporated [7]. Therefore, it can be said that the interest in desalination goes back as far as 4th century B.C.

The first published works for desalination are related to Muslim alchemists in 1551 who lived in the Middle East region [23]. Mouchot reported in 1869 that in the 15th century, Muslim alchemists used polished Damascus concave mirrors to focus solar rays onto glass vessels containing seawater to produce fresh water [24], [25].
Advanced technologies, that mimic natural processes such as evaporation-condensation or osmosis in order to obtaining fresh water from seawater, have been developed only in modern decades. Basic desalination processes were used on naval ships in the 17\textsuperscript{th} to 19\textsuperscript{th} centuries. For example, in 1790, the US Secretary of State, Thomas Jefferson, received an offer to sell the government a scheme to convert salt water to fresh water [6]. The first desalination units were eventually built for ships which were used to provide fresh boiler water, removing the need to travel with cargo holds full of water [21]. Years later, a British patent was granted in 1852 [26] and then in 1872 the first solar still was designed by a Swedish engineer, Carlos Wilson, and constructed in Chile [23]. In 1912, a 75 m\textsuperscript{3}/day desalination plant was installed in Egypt [27]. The island of Curacao in the Netherlands Antilles was the first location to make a major commitment to desalination in 1928, followed by a major seawater desalination plant built in Saudi Arabia in 1938 [6], [28]. During that period (1929-1937), the total desalinated water capacity increased due to the start of oil industry [27].

In the 1940s, during World War II, research on desalination was conducted to find proper ways to meet military requirements for fresh water in regions that soldiers were facing drinking water shortages [6]. For example, Telkes [29] developed a plastic still inflated with air for desalination, which was used by the US Air Force and US Navy during the World War II. After World War II, the United States and other countries continued their work on desalination. The U.S. Congress passed the Saline Water Conversion Act (PL 82-448) in 1952, which created and funded the Office of Saline Water within the Department of the Interior's Bureau of Reclamation [6].

In the 1960s, desalination science entered into a new and modern era; it was a special time for commercialisation because of the dramatic growth of population and water shortages experienced worldwide. New methods of desalination were considered based on fossil resources, because many oil-rich countries in the MENA region (Middle East and North Africa) had been faced with water shortages and therefore preferred to dedicate a part of their natural energy resources (oil and gas) to their local water desalination production instead of exportation [15]. Recently, desalinated water has become a commodity for many countries and desalination plants are not limited to the MENA region alone [30].
The first generation of desalination plants were commissioned in Shuwaikh, Kuwait and in Guernsey, Channel Island in 1960 [27]. By the late 1960s, desalination plants that could produce up to 8,000 m$^3$/day were beginning to be installed in various parts of the world and mostly worked based on thermal process which was expensive as they required a lot of energy [31]. Nevertheless, they were good enough for Middle Eastern oil-rich countries. In the 1970s, membrane processes were being used extensively and commercialised, however, the process was not yet scaled up for large production rate [27], [31]. Finally, in the 1980s, desalination became a fully commercial enterprise and this continues today [31]. According to the recent International Desalination Association (IDA) report [19], the global contracted capacity of desalination plants by the end of 2012 was around 80.5 Mm$^3$/day fresh water and it is expected to hit 104 Mm$^3$/day by 2016 [32]. The main worldwide use of desalinated water is for municipal and industrial purposes. In 2008, 70% of the globally usage was related to municipalities and 21% for industries [30].

1.3 Desalination in Australia

The Australian history of desalting goes back to the wood-fired stills of the Coolgardie goldfields over 100 years ago [33]. The main reason for providing potable water from the sea or brackish groundwater in Australia was related to mining, defence, communications transport, tourism activities and remote communities especially for those located in arid zones. A report on 2002 mentioned only a limited number (less than 10) of small desalination plants supplied public water, primarily because of the higher cost of desalinated water as compared to conventional water supplies [34]. However, due to the scarcity of conventional sources of freshwater and the growth of population in major cities over time, desalination is expected to become more popular in Australia. A new analysis shows desalination provides a more viable and cost effective water supply alternative as compared to large rain-dependent dams over a long term planning [35].

Australia's population is expected to grow to nearly 36 million people in 2050 [36]. This population growth needs growth in public infrastructure, including energy and water supply. A report showed that under this projection, it is estimated that
the total urban water use is to increase by approximately 42% from 1,505 Mm$^3$/yr in 2009 to 2,136 Mm$^3$/yr in 2026 [37]. Other key factors such as climate change, population density, economic growth and cost of water can also influence water usage in Australia and highlight the importance of the desalination industry's role. A recent analysis has estimated that the required energy for water provision will be around 21 Peta Joules (PJ), which in 2030 means a 200% increment as compared with that in 2006, only if 40% of the total water usage were provided by desalination [38]. In 2009, the capacity of desalination plants was estimated to be 294 ML/day in Australia. At the time, it was expected to increase to 2195 ML/day by 2013 [33], a 7.5-times increment over 4 years.

Various desalting processes are used in Australia, being installed in the 1980's and 1990's, the most common technology being Reverse Osmosis (see section 1.4.2) [39]. Between mid-2011 and mid-2012, the capacity of new plants contracted in Australia was about 62,089 m$^3$/day (all for Reverse Osmosis (RO) technology); however 2012 was a big year for Australian desalination industry, because two significant worldwide RO plants in Wonthaggi (444,000 m$^3$/day) and Adelaide (300,000 m$^3$/day) came online, standing second and third place in size, respectively, after the Magtaa desalination plant with the capacity of 500,000 m$^3$/day in Algeria [19]. A report in 2009 [33] has estimated the capacity of present, proposed and under construction desalination plants will be around 2,195 ML/day (13.4% in operation, 42.1% proposed and 44.5% under construction) in 2013, of which 60% of operated plants were in Western Australia (WA) state and Victoria (VIC) state, which at 0.2% had the lowest share.

The RO process is the most widely used process, as all large-scale desalination plants in Australia utilise this technology [40]. This is due to the available underground brackish water and the oceans water which surrounds Australia, making suitable feeds for the RO process [15]. Major RO desalination plants in Australia include Kwinana (144 ML/day, 2006), Bunningup (153 ML/day, 2011) and Karratha (140 ML/day, 2011) in Western Australia (WA), Olympic Dam (120 ML/day, 2011) and Adelaide (300 ML/day, 2011) in South Australia (SA), Wonthaggi (444 ML/day, 2011) in Victoria and Kurnell (250 ML/day, 2010) in New South Wales (NSW) and Tugun (133 ML/day, 2008) in Queensland [19], [30].
1.4 Desalination Technologies

In general, all applicable desalination processes can be divided into two main categories based on the phase change of saline feed water.

- **Desalination with phase change**: This category includes all heat driven processes where freshwater is produced by evaporation and condensation phenomena.

- **Desalination without phase change**: In this category, separation is achieved by passing saline water through membranes without involving phase change and RO (Reverse Osmosis) is the iconic example.

![Figure 1-1 Major Desalination Processes](image)

Figure 1-1 shows the most dominant methods for seawater desalination. Selecting the proper method for saline water desalination is dependent on many factors, such as location of the plant, plant capacity, plant life, initial costs, interest rate, required freshwater quality, type of energy sources, plant load factor, and water price cost. In industrial applications, for which the main aim is to extract freshwater from the process of liquid streams, it should be determined whether the selected process can handle the liquid or not. For instance, in mining industries and refineries such as alumina refineries, thermal desalination processes such as MED and MSF are more suitable than RO processes, which cannot handle the process streams. As an example, in the evaporation units of an alumina refinery plant (see Chapters 5 and 6), RO technologies cannot extract the freshwater from the process liquor (which can be considered as equivalent of...
Chapter 1. Introduction of Desalination

20wt% caustic soda solution); therefore the only option is to use thermal technologies.

1.4.1 Processes with phase change

The principle examples for processes with phase separation are MSF (Multi-Stage Flash), MED or MEE (Multi-Effect Distillation/Evaporation) and VCD (Vapour Compression Distillation), including TVC (Thermal Vapour Compression) and MVC (Mechanical Vapour Compression). For more information regarding techniques other than MED, MSF and VCDs which are explained in this section, the following references are suggested:

[8], [23], [27], [31], [34], [41]–[52]

These methods mimic the natural method of desalination using evaporation-condensation phenomena for producing fresh water from saline waters (or industrial process liquid streams). Due to the greater availability and lower cost of fossil fuels in the Middle East, the use of thermal desalination processes such as Multi Stage Flash (MSF) is localised primarily within these regions [53]. In 1996, 10,000 tons/yr of oil was required to produce 1,000 m$^3$/day freshwater [54]. In August 2012, the share of MSF and MED technologies of the total installed capacity for saline water desalination applications was around 31% [19]. These two technologies are the major technologies used in GCC countries, including 68% of the desalination capacities in 2012 [15], [55]. The other reason thermal phase change processes are more applicable in that region is that Persian Gulf water is known as 4H [15]. This means high salinity (around 45,000 ppm [6]), high turbidity, high temperature and high marine life. Moreover, due to the presence of radioactive materials because of the presence of warships and recent boron limitation rules in drinking-water, thermal phase change desalination technologies (MSF and MED) are mostly preferred over RO in that region [15]. However, some GCC countries such as Oman and KSA (Kingdom of Saudi Arabia) use RO for desalting the Indian Ocean and Red Sea water, which are more suitable feeds for RO than the Persian Gulf water [15].
Multi Stage Flash (MSF) distillation process (Figure 1-2) accounts for the major portion of desalinated municipal drinking water produced in the MENA (Middle East and North Africa) region [27]. Today it stands second place in the world after RO technology and is used primarily for desalting seawater [19]. This process has been in large scale commercial use for over thirty years, and because of a higher resistance against scaling as compare to MED technology, it has continually displaced the MED process from the 1950s (its development time) [30]. In 1957, four MSF plants with the total capacity of 9,084 m$^3$/day were installed in Kuwait [56]. The Al-Jubail plant (815,120 m$^3$/day) in Saudi Arabia (Figure 1-3) is the largest MSF plant in the world [57].

In this technology, as shown in Figure 1-2, seawater feeds are pressurized, heated, and discharged into a series of flashing chambers maintained slightly below the saturation vapour pressure, therefore a fraction of the feed water flashes into steam. The flashed steam passes through a mist eliminator and condenses on the exterior surface of the heat transfer tubing (condenser) that is located at the top of each stage. The condensed liquid then drips into trays and is collected as fresh water. All MSF plants consist of a heat source, heat recovery and heat rejection sections [58]; and typically their heat recovery sections are made of 19 to 28 stages (in modern and large MSF plants) [8]. Heat rejection section usually consists of three or more stages, which is used to control the recycled brine temperature [58]. Top brine temperature (TBT) in the heat input section is typically between 90 to 110°C [58], [59]; however, it can reach to 120 and 130°C with some modifications [60], [61]. Figure 1-2, shows the conventional MSF schematic design. Typical specific pumping power consumption for this
technology is mostly between 3.0 to 5.0 kWh/m³ for large modern MSF plants [15], [62]. The main advantages and disadvantages of this technology are listed as [8], [27], [30], [31], [34], [53], [63]:

**Advantages:**

- Produce large capacities of freshwater
- Independent of the salinity of the feed water
- Easy operation and very low performance degradation within the plant life
- Very high quality freshwater (less than 10 mg/L TDS)
- Minimal requirement for pre-treatment of the feed water as compared to the RO process
- Simple operational and maintenance procedures
- Higher resistance against scaling as compare to the MED process
- Long background of commercial use and reliability
- The possibility for being combined with other processes

**Disadvantages:**

- Expensive to build and operate
- Requires a high level of technical knowledge
- Highly thermal energy intensive process
- Operates at high temperature (top brine temperatures of 90–120°C)
- Low recovery ratio, means more feed water is required to produce the same amount of freshwater as compared to the MED process

![One of the 40 MSF units of Al-Jubail Desalination Plant – Kurt Kiefer [64]](image-url)
1.4.1.2 Multi Effect Distillation (Evaporation) (MED or MEE):

MED or MEE technology is the oldest industrial desalination method [65] and the most efficient thermal distillation process thermodynamically [66], although it stands in second place after MSF in the thermal desalination market [19]. The capacity of MED plants is typically more than 300 m$^3$/day as anything less than this is not financially viable [34].

In this system (Figures 1-4 and 1-5) the feed water is distributed onto the heat exchanger surfaces of the first effect. The heat source fluid (steam or hot liquid) that flows through the heat exchanger releases its energy to the distributed feed water and evaporates a portion of the feed water. The produced vapour then condenses in the heat exchanger of the second effect to evaporate more feed water in that effect. The brine from the first effect is then purged. At the second effect, the evaporated feed water goes on to power the third effect with the resulting brine being drained from the bottom of that effect. This process continues to the last effect with the corresponding produced vapour entering the condenser section and condensed by the incoming saline feed water acting as a coolant. Part of the preheated saline water is then sent to the various effects as feed water [2].

MEE processes have many possible configurations. For example, with respect to the evaporator design, it can use a rising film vertical or a falling film horizontal tube. With respect to the feed flow, it can be backward, parallel or forward. The effects arrangement, in turn, can be horizontal or use a stacked layout [65].
Chapter 1. Introduction of Desalination

The main difference between MED and MSF is related to the difference between boiling and flashing evaporation. Flashing technique (MSF process) requires much more feed water in comparison with boiling technique (MED Process) for producing the same amount of vapour; meaning flashing technique consumes more pumping power. Typical specific pumping power consumption for large scale MED plants is around 1.5 to 3 kWh/m$^3$ [8], [15], significantly lower than the MSF process.

The other difference is related to the top brine temperature (TBT) and scaling issue. Because of the design of MED processes, some cleaning procedures which can be applied for MSF, cannot be used in MED, thus, the best way to stand against scaling the problem is to keep the TBT around 65°C [67].

The main advantages and disadvantages of this technology are listed below [27], [34], [58], [63], [65]:

**Advantages:**

- Operates at low temperature
- High quality freshwater
- High thermal performance
- The possibility for being combined with other processes
- Low pumping power
- Very reliable process
- Handle normal levels of biological or suspended matter
- Minimum pre-treatment requirements
- Minimum labour requirement

**Disadvantages:**

- Expensive to build and operate (high CAPEX and OPEX)
- Susceptible to corrosion
- Low recovery ratio (not as low as MSF)
1.4.1.3 Vapour Compression Distillation (VCD)

The VCD process is mostly used for small to medium scale desalination plants [69]. They are often used in industries where fresh water is not readily available [31]. Generally, the heat for evaporation in VCD processes is provided by the compression phenomena. TVC (Thermal Vapour Compression) and MVC (Mechanical Vapour Compression) are the two configurations of VCD technology. In TVC processes, the vapour is compressed by a steam ejector (thermo-compressor), which needs live steam with enough pressure. In MVC processes, the vapour is compressed by a mechanically driven compressor which uses only electricity to operate; therefore, in cases that steam is not available as a heat source, this process can be an option.

VCD technology also can be used in combination with MED and MSF process, such as TVC-MED [70] or TVC-MSF [59]; a combination that can increase the system performance ratio over the stand alone MED or MSF system [27], [59]. TVC-MED and MVC-MED (Figures 1-6 and 1-7) are the most dominant processes due to the higher thermal performance of MED as compared with MSF processes; however TVC-MED system has the highest thermal performance among all steam driven processes. As it shows in Figure 1-7, in MVC-MED processes the produced vapour in the last effect is compressed in a mechanical compressor so this system is attractive in places where cooling sources are not available.

Typically an MVC-MED plant's capacity range is up to 3000 m$^3$/day [8], [69] and this range for a TVC-MED plant is around 20,000 m$^3$/day [69]. For reaching higher capacities, typically two or more plants will work in parallel. Curaçao Island
Chapter 1. Introduction of Desalination

12,000 m$^3$/d (1994), Umm Al Nar (UAE) 16,000 m$^3$/d (1998), Layyah (UAE) 36,370 m$^3$/d (2005) and Fujairah II (UAE) 38,670 m$^3$/d (2007) are utilising MED-TVC [71]. Tobruk desalination plant in Libya is one of the largest MED-TVC plants in the world (Figure 1-8).

**Figure 1-6** Schematic design of a Thermal Vapour Compression Multi Effect Distillation (TVC-MED) process for seawater application with horizontal falling film evaporators

The electrical power consumption of MVC-MED units is higher than TVC-MED (as mentioned above) and is typically between 7 and 12 kWh/m$^3$ [31]. MED-TVC’s electrical consumption is typically 1 kWh/m$^3$ [65], which is lower than conventional MED, as well. In contrast, MVC-MED does not require heat sources.

**Figure 1-7** Schematic design of a Mechanical Vapour Compression Multi Effect Distillation (MVC-MED) process for seawater application with horizontal falling film evaporators

Bijan Rahimi
As mentioned above, TVC-MED and MVC-MED are the most dominant steam driven processes as compared to other thermal desalination processes such as MSF and MED. Their main advantages and disadvantages can be listed as [27], [73]:

**Advantages:**

- Operates at low temperature
- High quality freshwater
- No need for steam line and cooling source (For MVC-MED)
- Highest thermal performance (For TVC-MED)
- Lowest pumping power (For TVC-MED)
- Very reliable process
- Thermo-compressor is very robust (For TVC-MED)
- Minimum labour requirement

**Disadvantages:**

- Expensive to build and operate (high CAPEX and OPEX)
- Susceptible to corrosion.
- High pumping power (For MVC-MED)
- Higher investment cost due to the cost of mechanical compressor (For MVC-MED as compared to MED and TVC-MED)
1.4.2 Processes without phase change

In these processes, as mentioned before, freshwater is extracted by passing saline water through membranes without involving phase change and RO (Reverse Osmosis) is the main technology which is under this category. Membrane based desalination plants, which make up the majority of plants worldwide [17], are largely powered by electricity derived from non-renewable sources.

An osmosis phenomenon is the transfer of the solvent (pure water) of a solution (saline water) based on a concentration gradient, through a semi-permeable membrane. For example, if freshwater and saline water separated by a semi-permeable membrane, due to the concentration gradient between the two medium, freshwater (low concentrate medium) is moving toward the saline water (high concentrate medium) for making an overall concentration balance (equilibrium) in the system. The principle of this moving is osmotic pressure. Therefore by gradually increasing the pressure on the saline water side (against the osmotic pressure direction), at a point which the applied pressure is to be equal to the osmotic pressure, the water flow across the membrane will be stopped. At this point, any increase in the saline water pressure side drives the flow in the opposite direction; meaning, instead of moving freshwater toward the saline water, the freshwater can be separate from the saline water. This process is called Reverse Osmosis (RO). The pressure difference between the applied pressure and the osmotic pressure is one of the key factors in the RO process to identify the freshwater mass flow rate that passes through the membrane (Figure 1-9) [39]. Depending on the feed water quality, the technology and the type of membranes, around 30% to 80% of the total feed water can be desalinated [34].
Typically RO processes include four major sections, namely pre-treatment, high pressure pumps, membranes and post treatment (Figure 1-10). A single stage RO's produced water has a salinity of less than 500 mg/L TDS [34]. Pre-treatment of the feed water is an essential component of the RO plant, as it's used to prevent scaling of the membranes. Because there is no requirement to heat or phase change, the major use of energy for RO is to pressurize the feed water; however, sometimes feed water preheating is used for decreasing the specific pumping power consumption [74], [75]. The typical specific electrical consumption of RO plants is between 3 to 7 kWh/m$^3$ [40], [76], [77].

RO is the fastest growing desalination process as compared with other desalination technologies due to the reduction in membrane costs and developments of energy recovery devices [78]. Its share of installed capacities was 63% in 2012 [19].

The main advantages and disadvantages of this technology are listed as [27], [34], [63].

**Advantages:**

- Quick to build and simple to operate
- Easy to extend the capacity by adding on extra modules
- Low energy consumption
- No need to heat sources
- Lower usage of chemicals for cleaning purposes
- No need to shut down the entire plant for scheduled maintenance

**Disadvantages:**

- Expensive membranes and low life expectancy (2-5 years)
- Sensitive toward changes in feed water salinity
- Need for high quality standard of materials and equipment
- Bacterial contamination is possible
- Pre-treatment of the feed water is essential
- High pressure operation which causes mechanical failure
1.5 Energy Consumption and Environmental Impacts

Desalination is considered an energy intensive process, however it represents only a small fraction of the total national energy consumption [77]. As a very general figure, energy consumption for desalinated water in a household environment is only around 3.2% of the total family energy consumption, including electricity, gas and fuels [79].

Desalination processes predominantly require large amounts of electrical and thermal energy. Concurrently water desalination, by both thermal and membrane processes, is becoming cheaper owing to material improvements, process improvements and an increase in competition [10].

The global rate of installation of water desalination facilities is increasing, with estimates of an approximate 55% increase in the number of plants installed in 2012 over the previous year [10]. The increased adoption of desalination represents a significant rise in global energy usage, due to the energy intensive processes required. For perspective, consider that the total amount of energy used for desalination worldwide has become comparable to the total energy requirement of a small industrialised country such as Sweden [80].

Around 30-50% of the product water cost is related to the energy cost of desalination process [77]; this share makes the energy saving issue an important key factor for water price reduction. Therefore, any improvements into the energy utilisation and efficiency of desalination processes are important, particularly...
those that incorporate the use of renewable energy sources such as industrial waste heat, geothermal and solar energy [3].

The minimum theoretical energy for desalination is the minimum required isothermal reversible work of separation to produce freshwater from seawater, which is independent of the process and is a function of the concentration and recovery factor; Figure 1-11 [32]. This theoretical threshold for 35,000 ppm NaCl solution is equal to 0.79 kWh/m$^3$ [79]; this value for a typical recovery of 50% is changed to 1.06 kWh/m$^3$ [32]. The actual energy consumption for desalination process is larger than the abovementioned quantities due to all irreversibilities related to the nature of the utilised desalination process. For example in thermal desalination plants the majority of these irreversibilities are related to heat losses. MSF and MED processes consume both thermal and electrical energy. For a typical MSF plants with the maximum live steam (as heat source) temperature of 120°C, the thermal energy consumption is around 12 kWh/m$^3$ of produce freshwater, while it is around 6 kWh/m$^3$ for an MED plant, which operate at lower temperatures (less than 70°C) [81]. The electrical consumption of these processes also should be considered to find the overall energy consumption. In contrast, RO processes only consume electrical energy in a range between 4 and 7 kWh/m$^3$ for seawater application [77], [79], [81], [82]. Thus, the overall energy consumption of RO processes is much closer to the minimum theoretical consumption threshold as compared to MED and MSF processes. However, in such cases that the thermal energy is free, the overall energy consumption of both MED and MSF processes which takes cost will be close to the theoretic threshold, as well.

The following equation [79], [83], [84] shows the irreversibility of thermal processes as a function of Boiling Point Elevation ($BPE$) and the temperature driving force.

$$w_{irr} = \frac{w_{rev}}{\eta_{Carnot}} \cdot \left(1 + \frac{\Delta T_{irr}}{BPE}\right) \quad (1-1)$$

In this equation, $\Delta T_{irr}$ should be more than the relevant $BPE$, otherwise no evaporation occurs [79]. For example, for a range of 35% to 45% recovery, as shown in Figure 1-9, the minimum required energy ($w_{rev}$) for 35,000 PPM NaCl solution is between 0.9 and 1.0 kWh/m$^3$ [32]. For a typical MED plant with top
Chapter 1. Introduction of Desalination

Brine temperature of 70°C, the average boiling point elevation is around 0.7°C [79]. Assuming a 1.5°C temperature difference driving force for the plant and a 75% Carnot efficiency (however the Carnot efficiency for thermal desalination plants is much lower than this figure), the actual work ($w_{irr}$) is between 1.8 and 2 kWh/m³.

![Figure 1-11 Theoretical minimum energy consumption for desalting pure NaCl solution. The highlighted range is the typical recovery range for an SWRO plant. [32]](image)

As mentioned before, 30-50% of the water cost is related to the energy consumption of desalination processes which is related to both electrical and thermal energy. For example, RO process has a high overall efficiency at the expense of consuming a large amount of electricity, especially for running the high pressure pumps. For thermal processes, such as MSF and MED, it is related to large thermal energy consumption, besides the electrical energy consumption. Therefore in respect of production rate, economic feasibility and environmental friendliness, the optimisation of desalination methods should be considered in the context of minimising energy consumption [2]. In this regard, incorporating renewable energy sources into desalination has been shown to be both technically and economically feasible and it should be considered an incontestable goal of ongoing research into desalination technology to incorporate renewable energy technologies henceforth [3].

Though more economical than renewable energy sources, the use of fossil fuels should also be framed within the context of the environmental implications including the greenhouse effect, carbon dioxide emissions and associated
environmental pollution. Both energy and environmental issues are linked together because the energy generation processes which produce the required energy for desalination processes has potential environmental impacts. However, the design and management of the desalination process by itself also can have environmental impacts [77]. For thermal processes such as MSF and MED, discharging fumes such as CO₂, NOₓ and SOₓ, and the rejected concentrated brine, are the two major impact factors on the environment [85]. Fumes are directly connected to the energy supply processes, so any use of alternative (sustainable) energy sources such as industrial waste heat, geothermal or solar energy instead of fossil fuels can substantially reduce these greenhouse gas emissions. As a perspective, around 1.4 – 1.8 kg CO₂ is emitted per each cubic meter of freshwater in current state-of-the-art SWRO plants [32], [86]–[88].

The environmental impacts of desalination plants are not limited to greenhouse gasses. Intake seawater system, brine disposal processes and energy dissipated (thermal impact) in the sea by thermal processes can affect on the marine life and the local elevation of salinity and temperature which need more attention [81].

Intake seawater system is associated with the impingement and entrainment of marine organisms [6], [28], [81]. By locating low velocity open surface intakes with a combination of appropriate meshed screens into deeper offshore waters and reducing the volume of intake water by re-using the power plant’s cooling water as feed for the desalination plant, the impacts from impingement and entrainment problems will be substantially minimised [28], [32], [81].

Brine rejection into the ocean can increase the local temperature and salinity. Many studies have shown that elevated salinity has little or no environmental impact [77] and pre-diluting the rejected brine with other waste streams such as power plant cooling water can solve the high salinity impact problem [81]. Pre-treatment, post-treatment and washing processes reject some chemicals such as heavy metals, anti-foams, anti-scalant, coagulants and cleaning chemicals to the ocean, and tracing the impact of these chemicals into the environment is an important issue which is recognised as an environmental hazard [89]. To avoid the impacts of thermal rejection, the maximum heat dissipation is conducted from
the waste streams before entering to the sea [81]. Sommariva, et al., [85] shows a clear relationship between the plant efficiency and thermal environmental impact, which shows the higher the plant efficiency, the lower the environmental impact.

Many researches have been conducted on the environmental impact of desalination [32], [77], [79], [81], [82], [85], [90] and ISO 14000 is an assessment tool for environmental behaviour of desalination plant standards [85] but they are still in their early development phase.
Chapter 2
Low Grade Sensible Heat Driven Distillation

This chapter introduces low grade sensible heat driven distillation processes particularly for saline water distillation. Typically heat sources with liquid medium and temperatures lower than 100°C are known as low grade sensible heat sources. The conventional Multi Effect Distillation (MED) is the best thermal processes with respect to thermal performance, which can be coupled with these kinds of heat sources. A previous invention called Boosted MED (B-MED) introduced a 22% improvement with respect to freshwater yield as compared to the low grade sensible heat driven conventional MED process. Both conventional and boosted MED processes are described in this chapter. The Next chapter introduces the new novel design called Flash Boosted MED (FB-MED) which brings a 43% improvement as compared to the conventional MED process.

Highlights:
- Introducing industrial waste heat and sensible heat sources
- Definition of high- and low-grade waste heat sources
- Conventional MED systems coupled with low grade sensible waste heat sources
- Introducing the Boosted MED (B-MED) process as the current novelty

Publication:
Majority of this chapter has been extracted from some sections of my below publications [2]-[4]:
2.1 Low Grade Sensible Heat Sources

Typically, most of the thermal desalination units in seawater application or as evaporation units in mineral refining industry are powered by live steam. Live steam is one of the most valuable and expensive industrial energy resources and any process that can decrease the consumption of this precious resource which in turn saves fuel consumption in the steam plant will significantly reduce production cost and greenhouse gas emission and thus global warming.

In seawater applications, around 30-50% of the desalinated water cost is related to the energy cost of desalination process [77] that is directly attributed to the live steam consumption in thermal desalination processes; this share underlines the importance of saving energy for reducing the water price and the desalination market in general. In industrial evaporation units, on the other hand, the price of freshwater production is not a direct issue. This is because the evaporation process subsumes under the main process and freshwater is not a direct outcome of the plant; so any expenses of the evaporation system will be reflected in the final product cost that is different for different industries. For this purpose the solution is to monitor the amount of saved live steam from the optimised evaporation units, which can in turn be translated as tangible savings and economic benefits.

With this purpose the optimisation of desalination (and industrial distillation) methods should be considered in the context of minimising energy consumption [2] for which the importance of low grade sensible waste heat sources becomes evident.

Waste thermal energy has always been an important issue in the process industries. Management of waste heat resources is one of the important subjects in process plants. In general, industrial thermal energy can be divided into high and low grade types. High grade thermal energies are referred as the heat which is viable to be recovered within the processes, and low grade heats are not and usually rejected to the environment [91]. Temperature wise, the threshold temperature range for low grade heat source is around 250°C [92]; therefore, any heat sources with temperatures lower than this can be classified as wastes.
However, in this work, we restrict our consideration of those low grade heats that are between 60°C and 100°C. This type of low grade heat sources is termed sensible heat sources if the waste heat medium is liquid, and is accompanied by a temperature drop during the heat transfer process. This kind of heat sources, on account of their low temperatures, is suitable for conventional Multi Effect Distillation (MED) process with a top brine temperature of around 70°C.

Low grade waste sensible heat sources are available in many industries and for plants located in the coastal area, they can potentially be used for seawater desalination (e.g. [66], [91], [93]–[97]). In some other plants, these waste heat sources can be used for the evaporation processes of the plant for internal usage, such as in alumina refinery plants (Chapters 5 and 6) [98]. Management of these kinds of heat source has always been an important issue in the process industries. Desalination with low grade sensible heat sources is not limited to industrial waste heat streams. Low grade geothermal heat sources with a wellhead temperature lower than 100 °C [99] can also be utilised for desalination purpose (e.g. [80], [95], [99]–[106]).

One of the main advantages of low-grade sensible heat sources is related to carbon dioxide emission and global warming issue. If the required energy hails from fossil fuel source then the freshwater production will contribute to carbon dioxide emission and consequently global warming. Low grade sensible heat sources such as waste heat from process plants and geothermal energy generate minimal greenhouse gasses.

In this chapter, the MED (Section 2.2) as the conventional process (which is the most efficient process to couple with low grade sensible heat sources) and Boosted MED [95] as a recently reported novel process (Section 2.3) are explained, and in the next chapter the latest novel Flash Boosted MED [2] is described as the superior technology for low grade heat driven desalination (evaporation) purposes as compared to both conventional and boosted MED systems.
2.2 Conventional Multi Effect Distillation (MED) Process

It is known that TVC-MED has the highest performance among all steam driven thermal desalination technologies [27], [65], [69]. It consumes less electrical energy and utilises the thermal energy in an optimised way as compared to the other thermal desalination technologies. However, in the absence of medium pressure (MP) steam and when the low grade sensible heat source (hot liquid medium) is the only available heat source, the conventional MED technology is a superior option.

In the temperature range of low grade sensible heat sources (< 100°C), MED is ideal as its top brine temperature varies between 60°C to 75°C [2]. This process is more efficient than other conventional thermal processes such as Multi Stage Flash (MSF) process. However, this process is inefficient as the outgoing heat source temperature is still sufficiently high [96]. Figure 2-1 shows a conventional MED system, coupled with a low grade sensible heat source. As it shows, in the conventional MED process, the feedwater is distributed onto the heat exchanger surfaces of the first effect which is heated by the sensible heat source. The temperature drop occurs in the heat-source stream. The produced vapour then condenses in the next-effect heat exchanger and act as a heat source for the feedwater which is distributed in this effect; and the brine is purged. This process continues to the last effect. At the end, the corresponding produced vapour enters the condenser section and is condensed by the incoming saline feedwater acting as a coolant.
Chapter 2. Low Grade Sensible Heat Driven Distillation

The preheated conventional MED system as shown in Figure 2-2, is another possible option which brings about an improvement as compared to the conventional MED system; however it has recently been shown that, in terms of low-grade sensible heat applications, the B-MED process (which is explained in the next section) is economically superior to the feed preheating MED configuration [107], and hence the latter will not be compared in this work.

**Conventional MED process with feed preheating module**

![Figure 2-2 Schematic design of a 4-effect preheated feed MED process coupled with a sensible heat source](image)

### 2.3 Boosted Multi Effect Distillation (B-MED) Process

The novel B-MED process, shown in Figure 2-3, utilises a ‘booster unit’, an evaporator unit which receives the heat source fluid which has already been used to provide the necessary thermal energy for the primary MED effects. The heat source fluid, which still contains a considerable amount of useful thermal energy, is used to evaporate more feed within the booster unit, which causes a higher temperature drop for the heat source medium as compared to the conventional MED process [95]–[97], [108], [109]. The vapour produced in the booster unit is then directed into an appropriate MED effect, supplementing the vapour and distillate production of the entire system. The B-MED process is particularly thermally efficient when utilising lower temperature heat sources, achieving
production rates that are up to 22% greater than that of the optimised MED alternative utilising the same heat source [2].

**Boosted MED**

![Diagram of Boosted MED system](image)

**Figure 2-3 Schematic design of a Boosted MED (B-MED) system**

In this regard, a state-of-the-art pilot plant consisting of two serially connected Alfa Laval's single-effect rising film plate evaporator/condenser modules has also been successfully tested at the National Centre of Excellence in Desalination Australia (NCEDA) and confirmed the fundamental potential of further improvement of the boosted and flash boosted configurations as compared to the
conventional MED process when it is coupled to low grade sensible heat sources (Figure 2-4) [110].

### 2.4 Process Simulation

Process simulations and validation of conventional MED and B-MED processes were produced based upon steady-state analysis methods used in combination with appropriate boundary conditions, which is available in detail in references [2], [4], [27], [95], [97], [107], [110]. The results of the process simulation indicate that the freshwater production of the B-MED process is indeed greater than the optimised conventional MED alternative, for a range of inlet heat source temperature which is between 65°C and 90°C [2], [3], [107]. Furthermore results indicate that the waste heat performance ratio, $PR_{WH}$ [2], [96], improves in much the same manner as the production rate, indicating that the B-MED process better utilises the energy available in the heat source fluid as compared to the conventional MED.

Insofar as low grade heat applications are concerned, the conventional performance ratio ($PR$) which holds that heat comes with a premium as it is consumed does not capture the essence of a desalination system driven by such heat sources, which only attracts a one-off investment cost, instead of an ongoing cost of thermal energy used (Equation 2-1). A waste–heat performance ratio ($PR_{WH}$), as in Eq. (2-2), is used instead which encourages the maximal use of the enthalpy of the low grade sensible heat source relative to the heat sink [2], [96]

$$PR = \frac{m_D \Delta h_{ref}}{m_{HS,1}(h_{f,HS,1,in} - h_{f,HS,1,out})} \quad (2-1)$$

$$PR_{WH} = \frac{m_D \Delta h_{ref}}{m_{HS,1} \Delta h_{avail}} = \frac{m_D \Delta h_{ref}}{m_{HS}(h_{f,HS,1,in} - h_{f,C,in})} \quad (2-2)$$

$\Delta h_{ref}$ is the specific reference enthalpy of the distillate that is equal to 2336 kJ/kg as an industrial benchmark [96]. $\Delta h_{avail}$ represents the maximum exploitable energy of the heat source relative to the lowest available temperature which in this case is the condenser inlet temperature.

In steam driven systems such as TVC-MED the gain output ratio ($GOR$) which is the ratio of the freshwater production rate to the heat source steam input,
facilitates an effective benchmark for comparing these systems. Any improvement on steam driven systems can be interpreted as the use of less steam to produce one unit of production rate. In these processes, the latent heat of steam is utilised as the driving force, the amount of which is calculated by knowing the steam flowrate at the prevailing steam temperature and pressure. Therefore, in an optimised steam driven process with a known amount of available latent heat energy, the maximum number of effects (which is limited by top brine temperature, cooling water temperature and design temperature differential across each effect [96]) brings about the maximum production rate. Therefore, in each case (with the same steam operational condition that includes temperature, pressure and flowrate) any improvement can be benchmarked against the amount of steam used under the same process conditions (i.e., same steam operating condition, number of effects and production rate). In this situation, less steam consumption per unit of freshwater means a more thermally efficient plant.

In contrast, in those cases where low grade sensible heat source is used instead of steam, a conventional MED process is the superior option and is accompanied by a temperature drop for the heat source stream. In these processes, the outlet heat source temperature and the amount of released energy vary according to the heat source inlet temperature, which shapes the optimisation of the conventional MED process [2], [96], [111]. In this context, having a higher number of effects does not guarantee higher production rates [2]. This is because with the same process conditions (such as having the same cooling water temperature, top brine temperature and design parameters), a higher number of effects means higher heat source outlet temperature which means a lower temperature drop in the heat source stream and a smaller amount of energy extraction at the first effect. In contrast, a lesser number of effects produce a higher temperature drop and larger energy extraction from the heat source but the production rate per effect cannot then be multiplied through a greater number of effects. Therefore, in our optimisation, a balance between the number of effects and heat source temperature drop is always necessary (with respect to the operational and design limitations that include top brine and cooling water temperatures and temperature differential across each effect [96]) so as to find
the configuration that maximises the production rate. Moreover, for this type of
heat source, benchmarking the system simply by means of the minimal heat
source energy consumption per unit production rate is irrational [96]. It is far better
to exploit the heat source as much as possible (insofar as operational and design
limitations permit) to bring about maximal production rate [96].

To this end, as shown (based on the relevant assumptions, Table 2-1) in Table
2-2 [2] each optimised MED and B-MED processes generally requires a different
number of effects to maximise the production rate [2], [95], [96], [109], [111].

<table>
<thead>
<tr>
<th>Assumptions of thermodynamic simulations [3]</th>
</tr>
</thead>
<tbody>
<tr>
<td>Maximum top brine temperature (°C) 70</td>
</tr>
<tr>
<td>Heat source flow rate, ( \dot{m}_{HS,1} ) (kg/s) 100</td>
</tr>
<tr>
<td>Heat source temperature (°C) 65, 70, 75, 80, 85, 90</td>
</tr>
<tr>
<td>Feed to vapour ratio ( R ) (35% evaporation) 2.857</td>
</tr>
<tr>
<td>( T_{C,in} ) (°C) / ( T_{C,out} ) (°C) 28 / 38</td>
</tr>
<tr>
<td>( \Delta P_{in}# ) (Pa) 500</td>
</tr>
<tr>
<td>( X_F ) (ppm), Feed salinity 35,000</td>
</tr>
</tbody>
</table>

# relevant pressure difference for steam injection purpose in B-MED process

<table>
<thead>
<tr>
<th>Table 2-2 Main flowrates for optimised MED and B-MED over the considered heat source temperature range [3]</th>
</tr>
</thead>
<tbody>
<tr>
<td>Process Type</td>
</tr>
<tr>
<td>---------------</td>
</tr>
<tr>
<td>Optimised MED</td>
</tr>
<tr>
<td></td>
</tr>
<tr>
<td>Optimised B-MED</td>
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</tbody>
</table>

Figures 2-5 and 2-6 shows the optimised conventional MED and B-MED processes coupled with a 65°C inlet heat source temperature [3]. In Figure 2-6, the optimised B-MED configuration is able to accommodate one more effect as compared to the optimised conventional MED process (Figure 2-5) on account of a lower temperature drop at the first effect of the primary MED section. This is made possible because of the booster unit which can process the heat source further, so as to realise an increase to the total production rate by around 22% as compared to the conventional MED [2], [3], [111]. The same optimisation
methodology [2], [95], [96], [109], [111] is also applied to the other heat source inlet temperatures.

Figure 2-5 Schematic design of an optimised conventional MED system (typical quantities as per the simulation for 65°C inlet heat source temperature) [3]

All details related to the pumping power (as shown in Figures 2-5 and 2-6) including the efficiencies and relevant pressure heads have been explained in detail in Chapter 3 and reference [2].

Figure 2-6 Schematic design of an optimised Boosted MED (B-MED) system (typical quantities as per the simulation for 65°C inlet heat source temperature) [3]
In this chapter, a new novel process called Flash Boosted MED (FB-MED) is introduced. Its application in saline water distillation and alumina refinery evaporation plant process is explained. This chapter also includes a detailed explanation about mathematical simulation methods and thermo-economic analysis. To have a wide economic comparison, there is a detailed thermo-economic analysis including cash flow analysis, capital and operating cost, Net Present Value (NPV), Internal Rate of Return (IRR) and Unit Product Cost (UPC) estimation. In the next chapter the results of both technical and thermo-economic analysis of the FB-MED process are compared with the conventional and boosted MEDs processes as the benchmarks for seawater application.

Highlights:

- Introduce the new novel Flash Boosted MED (FB-MED) process
- Mathematical simulation of FB-MED processes
- Thermo-Economic simulation of FB-MED processes

Publication:

Portions of this Chapter have been extracted from my below publications [2]–[4]:

3.1 Introduction

As mentioned previously, in cases of low grade sensible heat sources, it is far better to exploit the heat source as much as possible (insofar as operational and design limitations permit) to bring about maximal production rate [3]. A novel boosted MED system tailored for waste sensible heat streams has been recently reported [95], [96], [108], [109], [112]. In this system (Figure 2-6), a steam booster unit is installed to better exploit the waste heat stream so as to increase the freshwater yield. This booster unit (or an evaporator) is powered by the outgoing waste heat source of the primary MED plant. The generated vapour from the booster unit is then introduced into an appropriate effect of the primary MED plant. This scheme substantially heightens the production rate, but the extent of improvement is limited by the temperature drop across the booster unit [2].

To further exploit waste sensible heat, an improved system called Flash Boosted MED (FB-MED) has been developed as shown in Figure 3-1 [2], [111]. The improvement is derived from the ability of the system to extract the maximal energy from the waste heat and transform it into heightened freshwater production in the primary MED plant. To this end, multiple flashing chambers are installed to efficiently transform the waste energy into valuable steam. Referring to Figure 3-1, the outgoing heat source from the primary MED plant goes onto heat up the feed water via a liquid-to-liquid heat exchanger, which has been marginally preheated by the outgoing brine from the last flashing chamber. The heightening of the feed water temperature via this approach is only limited by the practical temperature of approach of the liquid-to-liquid heat exchanger and top brine temperature and this scheme ensures that it comes sufficiently close to the outlet temperature of the heat source leaving the primary MED plant. The heated feed water then goes through a series of flashing chambers. Each flashing chamber produces supplementary vapour which is directed to an appropriate primary MED effect according to the relevant pressure/temperature differences, thereby supplying more heat and increasing the amount of feedwater which can be vaporised across the MED effect. The resulting effect of this vapour injection scheme is an increased distillate production by around 50% across the series of primary MED effects as compared to the conventional MED system [2], which is
made possible only by the addition of several pumps, flashing chambers and a liquid-liquid heat exchanger [3].

Flash Boosted MED systems are not limited to the above mentioned configuration (Figure 3-1); there are 7 major configurations mentioned in the patent [111]. The other configurations are designed based on the combination of these 7 major configurations. One of those is Flash Boosted Thermal Vacuum Compression Multi-Effect Evaporation (FB-TVC-MEE) process which is designed for an alumina refinery plant in Western Australia (see Chapter 6).

A detailed mathematical simulation and thermo-economic evaluation of the FB-MED process has been explained in the following sections; and as mentioned before, for the mathematical simulation of the conventional MED and boosted MED (B-MED) processes, it can be referred to [27], [58], [95]–[97], [113], [114].

3.2 Mathematical Simulation

As with other thermal processes, mathematical simulation of this process includes mass, salinity and energy balance equations. Relevant equations in an
equation system are solved to calculate important parameters such as production rate, power consumption, waste heat performance ratio, $UA$ values, temperature gradients across each effect and other parameters, all of which are useful for thermo-economic evaluation. In section 3.2.4, the relevant solving procedure flowchart of the FB-MED process for seawater application has been shown. The GRG (Generalised Reduced Gradient) method [115] is used for solving the equation system, by imposing the boundary conditions from the law of thermodynamics and appropriate operational, technical and economic constraints.

The assumptions that are considered for mathematical simulation are as:

1. Steady state process
2. Negligible external heat loss across each effect and flashing chamber
3. Constant inlet feed water temperature and salinity
4. Negligible internal pressure/temperature losses of the vapour flow in the demister, transmission lines, and condensation inside the tubes
5. All condensation processes happen in constant pressure
6. The produced water is pure water
7. All primary MED effects have equal and constant recovery ratios
8. A 500 Pa pressure driving force is considered for injecting vapour from the flashing chambers to the relevant MED effect [107]
9. A constant 3°C temperature difference is considered inside each primary MED effect, between the condensed vapour temperature (heat source) and the outlet highly concentrated feed temperature [2], [3]

For mathematical simulation, the mass, salinity and energy balance equations are written for each part of MED effects and flashing chambers, which will be explained in the following sections. These parts include the first effect of the primary MED, the second to the last effect of the primary MED (includes both injected and non-injected effects), the condenser, and the battery of flashing chambers (Figure 3-1).

For the properties of water, steam and seawater, the REFPROP package (REFerence fluid PROPerties) developed by the National Institute of Standards
3.2.1 The primary MED section

This section includes the MED effects and the last condenser as shown in Figure 3-1. The mass and salinity balances for all MED effects (save the last condenser), is written as (\( k \in \{1, ..., n\} \)):

\[
m_{F,k} = m_{B,k} + m_{V,k}
\]
\[
m_{F,k} \cdot X_{F,k} = m_{B,k} \cdot X_{B,k}
\]

With reference to the definition of feed to vapour ratio (\( R \)) and recovery factor (\( RF \)) as:

\[
R = \frac{1}{RF} = \frac{m_F}{m_v}
\]

The outlet highly concentrated (which known as brine in seawater application) mass flowrate and salinity can be written as:

\[
m_{B,k} = (R - 1) \cdot m_{V,k} = \left(\frac{R-1}{R}\right) \cdot m_{F,k}
\]
\[
X_{B,k} = \left(\frac{R}{R-1}\right) \cdot X_{F,k}
\]

The energy balance for the primary MED section should be written separately for the first effect, the second to the last effects (including both injected and non-injected effects) and for the condenser section.

3.2.1.1 First effect

The inlets and outlets of the first effect of the primary MED section are shown in Figure 3-2. As explained before, the heat source medium is entered to this effect and after releasing its sensible thermal energy, its temperature drops and leaves the effect. The released energy is transferred to the inlet feed water, which is distributed from the top onto the hot coils or plates (depending on the type of evaporator). The outlet of the feed water side is the highly concentrated feed stream (called brine in seawater application), which is purged from the bottom.
The vapour, which is canalised to the second effect of the primary MED section, is used as the heat source.

3.2.1.1 Energy balance:

The energy balance for the first effect according to the Figure 3-2 is written as:

$$\dot{m}_{F,1} \cdot h_{F,1} + \dot{m}_{HS,1,in} \cdot h_{HS,1,in} = \dot{m}_{HS,1,out} + \dot{m}_{V,1} \cdot h_{V,1} + \dot{m}_{B,1} \cdot h_{B,1}$$

(3-6)

$h_{F,1}$ is the enthalpy of the feedwater that is a function of the temperature and the salinity of the feedwater (See Appendix A for seawater application).

$$h_{F,1} = h_{f_f(T_{F,1},X_{F,1})}$$

(3-6-1)

$h_{HS,1,in}$ and $h_{HS,1,out}$ are the heat source medium inlet and outlet enthalpies at the first effect, respectively. In case of using condensed steam (pure water) as the heat source, these enthalpies are only a function of temperature and can be found from the steam tables or relevant software such as REFPROP [116]; otherwise they are a function of both temperature and salinity of their medium and should be calculated based on their relevant equations.

$$h_{HS,1,in} = h_{f_{HS}(T_{HS,1,in},X_{HS,1})}$$

(3-6-2)

$$h_{HS,1,out} = h_{f_{HS}(T_{HS,1,out},X_{HS,1})}$$

(3-6-3)
Chapter 3. Flash Boosted MED, the New Novel Design

$h_{V,1}$ is the enthalpy of the produced vapour (Equation 3-6-4) which is slightly superheated in seawater applications and completely superheated in process liquor applications in alumina refinery plants (see Chapters 5 and 6), since the produced vapour pressure is equal to the effect’s pressure ($P_1$), but its temperature is more than the saturated temperature at $P_1$, by $BPE$ ($T_{V,1} = T_{B,1} = T_{sat}(P_1) + BPE_{B,1} > T_{sat}(P_1)$). In seawater application, $BPE$ is not more than 1°C (see Appendix B), therefore the degree of superheated is small and negligible, but in contrast in liquor applications in alumina refinery processes, the $BPE$ (which is a function of temperature and concentration, see Appendix B) can reach 10°C, which makes a substantial degree of superheated steam.

$$h_{V,1} = h_{gV(T_{B,1}, P_1)} \quad (3-6-4)$$

where, $h_{gV}$ can be found from superheated steam tables or relevant software such as NIST REFPROP [116].

$h_{B,1}$ is the enthalpy of the highly concentrated outlet at its temperature and salinity.

$$h_{B,1} = h_{fB(T_{B,1}, x_{B,1})} \quad (3-6-5)$$

After rearranging equation (3-6), we have:

$$m_{HS,1} \cdot (h_{HS,1,in} - h_{HS,1,out}) = m_{V,1} \cdot h_{V,1} + m_{B,1} \cdot h_{B,1} - m_{F,1} \cdot h_{F,1} \quad (3-7)$$

The left hand side of the above equation is the amount of released energy from the heat source in the first effect ($\dot{Q}_{HS,1}$). With the combination of equations (3-7) and (3-4) the energy balance equation for the 1st effect is written as:

$$\dot{Q}_{HS,1} = m_{F,1} \cdot \left( \frac{1}{R} \cdot h_{V,1} + \left( \frac{R-1}{R} \right) \cdot h_{B,1} - h_{F,1} \right) \quad (3-8)$$

3.2.1.1.2 Temperature-energy profile and UA value:

Figure 3-3 represents temperature-energy profile of the heat source and the feed mediums across the first effect of the primary MED section. In this effect, the heat source temperature drops across the evaporator, while the feed temperature firstly (preheating zone shown as zone 1) increases from the inlet temperature to the relevant boiling temperature (namely, $T_{B^*,1} = T_{sat}(P_1) + BPE_{B^*,1}$), according to
its concentration, at the prevailing pressure ($P_1$) (a sensible-sensible heat transfer) and then increases further during the evaporation process (evaporation zone, shown as zone 2) because of continuing concentration that brings about heightened Boiling Point Elevation ($BPE$), to the relevant boiling temperature (namely, $T_{B,1} = T_{sat}(P_1) + BPE_{B,1}$). Therefore due to the different type of heat transfer between preheating and evaporation zones, there are two different $UA$ values. In seawater applications, due to small changes in the boiling point elevation during the evaporation process ($BPE_{B^*,1} \approx BPE_{B,1}$) a conservative and reasonable approximation can be considered, namely that the feed temperature remains constant ($T_{B^*,1} = T_{B,1}$). In the above mentioned equations, $BPE_{B^*,1}$ and $BPE_{B,1}$ are the boiling point elevations of the inlet feed and the outlet highly concentrated streams, respectively.

In our cases (in both seawater desalination and process liquor (Chapter 5 and 6) applications), the preheating zone (zone 1) covers less than 5% of the total energy transferred and therefore to a good approximation it can be ignored (As a rough estimation, it can be seen that the enthalpy change for unit of mass of pure water in liquid phase for a 30°C increase in temperature from 30°C to 60°C is equal to 125.45 kJ/kg, while the energy needed for evaporation of unit of mass of pure water at 60°C is 2357.70 kJ/kg is around 19 times more than the sensible enthalpy (125.45 kJ/kg) change). Hence the overall $U$ value (heat transfer coefficient, (kW/m$^2$.K)) for the first effect is practically $U_{z2}$, that relates to the evaporation zone (zone 2), where sensible-latent heat transfer happens.

![Figure 3-3 Temperature-energy profile for the first effect of the primary MED section of the FB-MED system](image-url)
To calculate the \( UA \) value for the two different zones, the pinch point temperature of the first effect’s evaporator (as it shows in Figure 3-3) should be calculated. For this purpose the energy balance for the first zone (Figure 3-3) is written as:

\[
\dot{m}_{HS,1} \cdot (h_{f_{HS(T_{pinch},X_{HS,1})}} - h_{f_{HST(HS,1,out),X_{HS,1}}}) = \dot{m}_{F,1} \cdot (h_{f_{F(T_{F,1},X_{F,1})}} - h_{f_{F(F,1),X_{F,1}}}) \tag{3-9}
\]

The right hand side term is the amount of absorbed energy by the feed in the first zone, \( (\dot{Q}_F,1,z_1) \), and the left hand side is the amount of released energy from the heat source in the same zone, \( (\dot{Q}_{HS,1,z_1}) \). Then, \( T_{pinch} \) is interpolated from the below equation:

\[
h_{f_{HS(T_{pinch},X_{HS,1})}} = \frac{\dot{m}_{F,1}}{\dot{m}_{HS,1}} \cdot (h_{f_{F(T_{B,1},X_{F,1})}} - h_{f_{F(F,1),X_{F,1}}}) + h_{f_{HST(HS,1,out),X_{HS,1}}} \tag{3-10}
\]

Therefore, \( UA \) value can be calculated from:

\[
(UA)_{1,z_1} = \frac{\dot{Q}_{HS,1,z_1}}{\Delta T_{lm,1,z_1}} \tag{3-11}
\]

where:

\[
\Delta T_{lm,1,z_1} = \frac{(T_{HS,1,out} - T_{F,1}) - (T_{pinch} - T_{B,1})}{\ln \left( \frac{T_{HS,1,out} - T_{F,1}}{T_{pinch} - T_{B,1}} \right)} \tag{3-12}
\]

The same methodology can be applied for the second zone to find out the relevant \( UA \) value.

\[
(UA)_{1,z_2} = \frac{\dot{Q}_{HS,1,z_2}}{\Delta T_{lm,1,z_2}} \tag{3-13}
\]

where:

\[
\dot{Q}_{HS,1,z_2} = \dot{m}_{HS,1} \cdot (h_{f_{HST(HS,1,in),X_{HS,1}}}) - h_{f_{HST(pinch,X_{HS,1})}}) \tag{3-14}
\]

\[
\Delta T_{lm,1,z_2} = \frac{(T_{HS,1,in} - T_{B,1}) - (T_{pinch} - T_{B,1})}{\ln \left( \frac{T_{HS,1,in} - T_{B,1}}{T_{pinch} - T_{B,1}} \right)} \tag{3-15}
\]

As mentioned before, \( (UA)_{1,z_2} \gg (UA)_{1,z_1} \) which means to a good approximation:

\[
(UA)_1 \approx (UA)_{1,z_2} \tag{3-16}
\]
3.2.1.2 Second to the last effect

Figures 3-4 and 3-5 represent the inlets and outlets of the second to the last effects of both with and without vapour injected effects configurations. As mentioned above, the produced vapour in each effect is used as a heat source for the next following effect. The vapour inside the tubes condenses and its energy is released to evaporate the inlet feed water which is distributed from the top onto the hot coils or plates (depends on the type of evaporator). Depending on the degree of superheating, the superheated vapour de-superheats to the relevant saturation temperature at its prevailing pressure ($T_{\text{Sat}(P_k)}$) and then it condenses (at a constant pressure). As with the first effect, the produced vapour goes on to power the next following effect with the resulting high concentrate feed being drained from the bottom of that effect.

![Figure 3-4 Schematic design of the effects of the primary MED section (Except the first effect) without vapour injection from the flashing chambers ($k \in \{2, ..., n\}$)](image)

For the effects with vapour injection from the flashing chambers, the amount of injected vapour is added to the produced vapour of the previous effect, which will be used as a heat source of that effect (Figure 3-5). The driving force for injection is considered to be a 500 Pa pressure difference to make a natural vapour flow and compensate for the relevant pressure drops [3], [107]. The temperature of the mixed vapour streams is calculated based on an energy balance equation. Being conservative, it is assumed that the pressure of the mixed streams is equal to the lowest pressure, which means the pressure of the previous effect (Figure 3-5); therefore just as with the non-injected effects, a superheated vapour is
available but with a slightly more degree of superheating (as compared to the non-injected effects). With this assumption, for the effects with vapour injection, the condensation occurs at the previous effect’s pressure (the same as the non-injected effects).

Figure 3-5 Schematic design of the effects of the primary MED section (Except the first effect) with vapour injection from the flashing chambers \((k \in \{2, \ldots, n\}; \ i \in \{1, \ldots, j\})\)

### 3.2.1.2.1 Energy balance

As it shows in Figures 3-4 and 3-5, the energy balance for these effects can be figured out as:

\[
\dot{m}_{HS,k} \left( h_{HS,k,in} - h_{f, sat(P_{k-1})} \right) = \dot{m}_{V,k} \cdot h_{V,k} + \dot{m}_{B,k} \cdot h_{B,k} - \dot{m}_{F,k} \cdot h_{F,k} \tag{3-17}
\]

The left hand side of the above equation is the amount of released energy from the heat source \((\dot{Q}_{HS,k})\). As can be seen, the difference between the equation (3-17) and equation (3-7) is only related to the left hand side, which is related the different type of heat transfer. Since in the first effect the heat source medium temperature drops while its phase does not change, but for the second to the last effects the heat source medium which is vapour, is firstly de-superheated and then condensed inside the tubes. After rearranging and combination with Equation (3-4) it can be written as:

\[
\dot{Q}_{HS,k} = \dot{m}_{F,k} \left( \frac{1}{R} \cdot h_{V,k} + \left( \frac{R-1}{R} \right) h_{B,k} - h_{F,k} \right) \tag{3-18}
\]

As shown in Figure 3-5, for the injected effects the heat source inlet mass flowrate is calculated by \((i \in \{1, \ldots, j\})\):

\[
\dot{m}_{HS,k} = \dot{m}_{V,k-1} + \dot{m}_{V',i} \tag{3-18-1}
\]
where ‘i’ is the relevant flashing chamber’s number that in an optimised seawater application it can be equivalent to ‘k – 2’, however it depends on the design parameters, heat source temperature drops and feed water properties.

The heat source inlet energy for the injected effects can be also written as:

$$h_{HS,k,in} = \frac{m_{V,k-1}h_{V,k-1}+m_{V',i}h_{V',i}}{m_{V,k-1}+m_{V',i}}$$  \hspace{1cm} (3-18-2)

where $h_{V',i}$ is the enthalpy of the produced flashed vapour in the flashing chambers which is explained in section 3.2.2.2.

3.2.1.2.2 Temperature-energy profile and UA value:

As Figure 3-6 shows the temperature-energy profile across the second effect to the last effect. As explained before, the heat source of these effects is the superheated vapour from the previous effect. Referring to Figure 3-6, in zone 1 (de-superheating zone) superheated vapour is de-superheating to the relevant saturated temperature of the previous effect. Then it condenses. In contrast, the trend of feed temperature profile is the same as in the first effect if the saturation pressure of the feed stream is less than the effect’s pressure, as shown in Figure 3-6(A). Therefore, in this case, the feed temperature increases to the relevant boiling temperature in preheating zone (zones 1 and 2) and then its temperature increases further during the evaporation process (however, as mentioned before, it remains constant in seawater application). In zone 3, during the de-superheating process, the feed water temperature raises up to $T_{F^*,k}$. In both seawater applications and process liquor in alumina refineries, the feed water temperature cannot reach its boiling point temperature in this zone, therefore $T_{F^*,k} < T_{B^*,k}$. The area, however, that is needed for de-superheating has a significant effect on the capital cost of the evaporator and this will be discussed in detail in Chapters 5 and 6 which the superheating degree is much higher than seawater application. In these effects, as mentioned before, a reasonable approximation for the overall $U$ value is simply related to the evaporation zone (zone 3).

Sometimes the feed saturation pressure is more than the effect's pressure, as what is shown in Figure 3-6(B). In this case, before the feed is being distributed into the tubes, a negligible amount of feed (less than 1% in our cases, see
Chapter 3. Flash Boosted MED, the New Novel Design

Chapter 5 and 6) is flashed while its pressure and temperature decrease. For these effects the inlet feed temperature is considered to equate to the relevant feed boiling temperature, \( T_{B^*,k} \) (which is related to its concentration) that corresponds to the pressure in the relevant effects, shown in Figure 3-6(B). After temperature reduction, once it distributes into the tubes, evaporation occurs. As shown in Figures 3-6(B), for these effects, the amount of energy required for de-superheating is negligible (see Chapters 5 and 6).

During the de-superheating process, the feed water temperature raises up to \( T_{B^{**},k} \) as shown in Figure 3-6(B). In our case, due to the small amount of energy which is released in the de-superheating zone, a reasonable approximation can be considered, namely \( T_{B^*,k} = T_{B^{**},k} \).

\[ h_{f\left(T_{F^*,k},x_{F,k}\right)} = \frac{m_{HS,k}}{m_{F,k}} \left( h_{HS,k,\text{in}} - h_{g_{\text{sat}}(p_{k-1})} \right) + h_{f\left(T_{F,k},x_{F,k}\right)} (3-19) \]
where \( h_{HS,k,in} = g_v(T_{HS,k,P_{k-1}}) \).

Therefore, \( UA \) value can be calculated from:

\[
(UA)_{k,x1} = \frac{m_{HS,k} \left( h_{HS,k,in} - h_{g_{sat}(P_{k-1})} \right)}{\Delta T_{lm,k,x1}} \quad (3-20)
\]

where:

\[
\Delta T_{lm,k,x1} = \frac{(T_{sat(P_{k-1})} - T_{F^*,k}) - (T_{HS,k} - T_{F,k})}{\ln\left(\frac{T_{sat(P_{k-1})} - T_{F^*,k}}{T_{HS,k} - T_{F,k}}\right)} \quad (3-21)
\]

In zone 2, the feedwater temperature raises up to its relevant boiling point while the vapour inside the tube is condensed. Therefore \( UA \) value can be found from:

\[
(UA)_{k,x2} = \frac{m_{F,k} \left( h_{fF(T_{B^*,k,X_F,k})} - h_{fF(T_{F^*,k,X_F,k})} \right)}{\Delta T_{lm,k,x2}} \quad (3-22)
\]

where:

\[
\Delta T_{lm,k,x2} = \frac{(T_{sat(P_{k-1})} - T_{B^*,k}) - (T_{sat(P_{k-1})} - T_{F^*,k})}{\ln\left(\frac{T_{sat(P_{k-1})} - T_{B^*,k}}{T_{sat(P_{k-1})} - T_{F^*,k}}\right)} \quad (3-23)
\]

The quality of the vapour inside the tube at the threshold between the second and the third zone also can be calculated from the enthalpy of vapour at that point, according to the energy balance equation, which is written as:

\[
h'^{g}_{k} = h_{g_{sat}(P_{k-1})} - \frac{m_{F,k}}{m_{HS,k}} \left( h_{fF(T_{B^*,k,X_F,k})} - h_{fF(T_{F^*,k,X_F,k})} \right) \quad (3-24)
\]

Therefore:

\[
\chi = \frac{h'^{g}_{k} - h_{g_{sat}(P_{k-1})}}{h_{g_{sat}(P_{k-1})} - h_{g_{sat}(P_{k-1})}} \quad (3-25)
\]

which in our applications \( \chi \) is not less than 98%.

At the end, for zone 3, the relevant \( UA \) value can be calculated as:

\[
(UA)_{k,x3} = \frac{m_{HS,k} \left( h'^{g}_{k} - h_{g_{sat}(P_{k-1})} \right)}{\Delta T_{k,x3}} \quad (3-26)
\]
where:

$$\Delta T_{lm,k,3} = \frac{(T_{sat(p_{k-1})} - T_{g^*,k}) - (T_{sat(p_{k-1})} - T_{B,k})}{\ln \left(\frac{T_{sat(p_{k-1})} - T_{g^*,k}}{T_{sat(p_{k-1})} - T_{B,k}}\right)}$$  \hspace{1cm} (3-27)

In seawater application since evaporation occurs in a constant temperature, thus a constant temperature difference between the heat source and feedwater ($\Delta T_{lm,k,3} = T_{sat(p_{k-1})} - T_{B,k}$) exists.

When feed stream flashes at the entrance of the effect as shown in Figure 3-6(B), the $UA$ value for zone 1 can be calculated from:

$$(UA)_{k,z1} = \frac{\dot{m}_{HS,k} \left( h_{gV(THS,k)P_{k-1}} - h_{gsat(P_{k-1})} \right)}{\Delta T_{lm,k,z1}}$$  \hspace{1cm} (3-28)

where:

$$\Delta T_{lm,k,z1} = \frac{(T_{HS,k} - T_{g^*,k}) - (T_{sat(P_{k-1})} - T_{g^*,k})}{\ln \left(\frac{T_{HS,k} - T_{g^*,k}}{T_{sat(P_{k-1})} - T_{g^*,k}}\right)}$$  \hspace{1cm} (3-29)

In zone 2, we have:

$$(UA)_{k,z2} = \frac{\dot{m}_{HS,k} \left( h_{gsat(P_{k-1})} - h_{fsat(P_{k-1})} \right)}{\Delta T_{lm,k,z2}}$$  \hspace{1cm} (3-30)

where:

$$\Delta T_{lm,k,z2} = \frac{(T_{sat(P_{k-1})} - T_{g^*,k}) - (T_{sat(P_{k-1})} - T_{B,k})}{\ln \left(\frac{T_{sat(P_{k-1})} - T_{g^*,k}}{T_{sat(P_{k-1})} - T_{B,k}}\right)}$$  \hspace{1cm} (3-31)

### 3.2.1.3 Condenser

The condenser (Figure 3-7) is located after the last effect to condense the last remained produced vapour (of the last effect) and in case of injection from the last stage of the flashing chamber (stage number ‘$j$’); the final mixture will be conducted to the condenser. In seawater application the coolant is seawater. In our case for all seawater application simulations the inlet seawater temperature has been considered to be 28°C. The cooling water absorbs the released energy from the condensation of the superheated vapour inside the tubes. A portion of the outlet cooling water which has been preheated in the condenser is used then
as the feedwater for the desalination process (in seawater application) and the remained is purged to the sea. In alumina refinery application (Chapters 5 and 6) the outlet cooling water is rejected to the evaporation pond.

\[ \dot{m}_{HS,cond} = \dot{m}_{V,n} + \dot{m}_{V',j} \]

\[ h_{HS,cond,in} = T_{HS,cond,in} \]

\[ \dot{m}_{C} \]

\[ h_{C, out} \]

\[ T_{C, out} \]

\[ X_{C} \]

\[ m_{HS,cond} \]

\[ h_{f, sat(p_n)} \]

\[ T_{sat(p_n)} \]

**Figure 3-7 Schematic design of the condenser with vapour injection**

### 3.2.1.3.1 Energy balance

The amount of required coolant mass flowrate (\( \dot{m}_{C} \)) is calculated from the energy balance equation, which is written as:

\[ \dot{m}_{HS,cond} \cdot (h_{HS,cond,in} - h_{HS,cond,out}) = \dot{m}_{C} \cdot (h_{C, out} - h_{C,in}) \quad (3-32) \]

where the left hand side term is the amount of energy released from the hot side (\( \dot{Q}_{HS,cond} \)) and:

\[ \dot{m}_{HS,cond} = \dot{m}_{V,n} + \dot{m}_{V',j} \quad (3-32-1) \]

\[ h_{HS,cond,in} = \left( \frac{\dot{m}_{V,n}h_{V,n} + \dot{m}_{V',j}h_{V',j}}{\dot{m}_{V,n} + \dot{m}_{V',j}} \right) \quad (3-32-2) \]

\[ h_{HS,cond,out} = h_{f, sat(p_n)} \quad (3-32-3) \]

\( h_{C,in} \) and \( h_{C,out} \) are the enthalpies of the inlet and outlet cooling waters which are a function of their temperatures and salinities, respectively.

\[ h_{C,in} = h_{f,C(T_C,in,X_C)} \quad (3-32-4) \]
\[ h_{C,\text{out}} = h_{f_c(T_{C,\text{out}}, X_C)} \quad (3-32-5) \]

3.2.1.3.2 Temperature-energy profile and UA value

As Figure 3-8 shows, the cooling water temperature increases across the condenser while the trend of the inlet superheated vapour temperature profile is the same as the other effects, so that zone 1 (namely the de-superheating zone) is negligible once again, with the overall heat transfer coefficient being practically \( U_2 \), where latent-sensible heat transfer happens. The impact of de-superheating is being considered as extra required heat exchanger area for the alumina refinery application (Chapters 5 and 6).

In zone 1 (Figure 3-8) the cooling water temperature will reach \( T_{C^*} \) which can be interpolated from the energy balance equation, which for this zone is:

\[
\dot{m}_{\text{HS,cond}} \cdot (h_{\text{HS,cond,in}} - h_{\text{g,\text{sat}(P_n)}}) = \dot{m}_C \cdot (h_{C^*} - h_{C,\text{in}}) \quad (3-33)
\]

Rearranging to:

\[
h_{C^*} = \frac{\dot{m}_{\text{HS,cond}}}{\dot{m}_C} \cdot (h_{\text{HS,cond,in}} - h_{\text{g,\text{sat}(P_n)}}) + h_{C,\text{in}} \quad (3-34)
\]

where:

\[
h_{C^*} = h_{f_c(T_{C^*}, X_C)} \quad (3-34-1)
\]
Chapter 3. Flash Boosted MED, the New Novel Design

$T_C$, is then calculated from the interpolation of equation (3-34-1).

With the same methodology of the other effects, $UA$ value for zone 1 is calculated by:

$$ (UA)_{k,z1} = \frac{m_{HS,cond} (h_{HS,cond,in} - h_{sat(p_n)})}{\Delta T_{lm,cond,z1}} $$

(3-35)

where:

$$ \Delta T_{lm,cond,z1} = \frac{(T_{sat(p_n)} - T_C) - (T_{HS,cond,in} - T_{C,in})}{\ln \frac{T_{sat(p_n)} - T_{C,in}}{T_{HS,cond,in} - T_{C,in}}} $$

(3-36)

In zone 2, the cooling water (cold side medium) temperature raises up to the outlet temperature while the vapour (hot side medium) is condensed. Therefore the relevant $UA$ value is calculated as:

$$ (UA)_{k,z2} = \frac{m_{HS,cond} (h_{gsat(p_n)} - h_{f sat(p_n)})}{\Delta T_{lm,cond,z2}} $$

(3-37)

where:

$$ \Delta T_{lm,cond,z2} = \frac{(T_{sat(p_n)} - T_C) - (T_{sat(p_n)} - T_{C,out})}{\ln \frac{T_{sat(p_n)} - T_{C,out}}{T_{sat(p_n)} - T_C}} $$

(3-38)

3.2.2 Flashing section

This section is the heart of the flash boosted MED (FB-MED) system. This section harvests the remained available energy from the heat source (as compared to the conventional and boosted MEDs processes), and distributes it to the primary MED effects with the aim of increasing the production rate. As mentioned before, the generated vapor from each stage of flashing is injected into the judicious effect of the primary MED plant for further boosting as in the previous steam boosted scheme. It is discernible that this scheme permits the waste heat source outlet temperature to approach the mixing point temperature between the flashing chamber outlet and the condenser outlet temperatures as close as is practicable. This process therefore engenders a maximally efficient scheme to exploit the potential of waste heat, far more than the conventional and boosted MED technologies [2].
The flashing section includes a liquid-liquid heat exchanger, a battery of flashing chamber, a brine recirculation process to decrease the make-up feed water flowrate and a deaerator to reduce the dissolved gases in the make-up water stream (Figure 3-9).

**3.2.2.1. Liquid-liquid heat exchanger**

This heat exchanger (Figure 3-10), that can be a plate type heat exchanger, harvests the remaining available energy which is not collected from the heat source by the first effect of the primary MED section, and delivers it to the flashing chambers. A typical 3°C approach temperature has been considered in all simulations of this work for this heat exchanger.

By knowing the heat source condition the amount of required flashing chambers' feed flowrate ($\dot{m}_{c,\text{hex}}$) is calculated from the energy balance of this heat exchanger, which is:

$$\dot{m}_{\text{hex}} \cdot (h_{\text{hex,in}} - h_{\text{hex,out}}) = \dot{m}_{c,\text{hex}} \cdot (h_{\text{c,hex,out}} - h_{\text{c,hex,in}}) \quad (3-39)$$

where:

$$\dot{m}_{\text{hex}} = \dot{m}_{HS,1} \quad (3-39-1)$$

$$h_{\text{hex,in}} = h_{HS,1,\text{out}} = h_{fS(T_{HS,1,\text{out}},X_{HS,1})} \quad (3-39-2)$$

$$h_{\text{hex,out}} = h_{fS(T_{\text{hex,out}},X_{HS,1})} \quad (3-39-3)$$
3.2.2. Flashing chambers

As known, when a pressurised liquid is entered to a flashing chamber with a pressure lower than its saturated pressure, the liquid becomes superheated and vigorously flashes (Figure 3-11). During the flashing phenomena a portion of the liquid evaporates while the temperature and pressure decreases until the equilibrium condition are achieved [124].

The mass and salinity balances for a flashing chamber are written as:

\[
\dot{m}_{FC,i,in} = \dot{m}_{V,FC,i} + \dot{m}_{FC,i,out} \tag{3-40}
\]

\[
\dot{m}_{FC,i,in} \cdot X_{FC,i,in} = \dot{m}_{FC,i,out} \cdot X_{FC,i,out} \tag{3-41}
\]

The energy balance is:

\[
\dot{m}_{FC,i,in} \cdot h_{FC,i,in} = \dot{m}_{FC,i,out} \cdot h_{FC,i,out} + \dot{m}_{V,FC,i} \cdot h_{V,FC,i} \tag{3-42}
\]

where:

\[
h_{FC,i,in} = h_{f,FC}(T_{FC,i,in},X_{FC,i,in}) \tag{3-42-1}
\]

\[
h_{FC,i,out} = h_{f,FC}(T_{FC,i,out},X_{FC,i,out}) \tag{3-42-2}
\]

\[
h_{V,FC,i} = h_{g}(P_{i},T_{V,FC,i}) = h_{g}(P_{i},T_{FC,i,out}) \tag{3-42-3}
\]
Therefore, with reference to the equations (3-40) and (3-42) the amount of generated flashed vapour is calculated by:

\[
\dot{m}_{VF,FC,i} = \dot{m}_{FC,i,in} \cdot \left( \frac{h_{FC,T_{FC,i,in},X_{FC,i,in}} - h_{FC,T_{FC,i,out},X_{FC,i,out}}}{h_{R(P_i,T_{FC,i,out})}} \right)
\]

In the abovementioned equations, \( P_i \) is the pressure of the relevant flashing vessel, and:

\[
T_{FC,i,out} = T_{sat(P_i)} + BPE_i + NEA_i
\]

\( BPE \) is the boiling point elevation (Appendix B) and \( NEA \) or Non-Equilibrium Allowance for flashing chambers is a function of flashing temperature range, saturation temperature, mass flow rate of brine per unit of chamber width, brine level inside the flashing chamber, and the design of the flash chamber such as chamber length, width and the orifice type of the brine transfer device from chamber-to-chamber (Appendix B) [125]. Under typical operating conditions, \( NEA \) varies between 0.03°C (as in the first few stages) to 0.8°C (as in the last low-temperature stages) in an MSF desalination plant [126]. In the present simulations for the seawater application, \( NEA \) is varied from 0.3°C (as in the first chamber) to 0.6°C (as in the last chamber), but the results that we present here pertains to a uniform 0.6°C \( NEA \) for all flashing chambers so as to be conservative [2]. For the alumina refinery evaporation plant application (Chapters 5 and 6), a fixed 0.5°C \( NEA \) is considered [98].
For the purpose of simulation and to make a connection between the flashing vessels and the primary MED effects, the pressure of each flashing chamber will be set based on the pressure of the judicious MED effect which the flashed vapour is injected to. Therefore a pressure difference ($\Delta P_{\text{inj}}$) between 500 Pa and 1 kPa is considered to be sufficient to have a natural flow [107].

$$P_i = P_k + \Delta P_{\text{inj}}$$  \hspace{1cm} (3-45)

$\Delta P_{\text{inj}}$ is considered to be 500 Pa and 1.0 kPa for seawater and alumina refinery applications, respectively [2], [3], [98].

### 3.2.2.3. Brine recirculation

For the aim of decreasing the fresh feedwater flowrate to the flashing section (make-up water), a brine recirculation configuration is used as shown in Figure 3-9. For this purpose, a rejected brine stream is then needed to control the mass and salinity balance in the flashing section [27].

The recycling of brine emanating from the series of flashing chambers effectively suppresses the amount of released non-condensable gases and thereby further improves the heat transfer efficacy and attenuates the venting and vacuuming power consumptions [2]. It also can decrease the chemical additive consumption and the size of pretreatment and deaeration facilities for the feed stream [27], [60], [61].

As shown in Figure 3-12, to compensate the water loss due to the evaporation and brine rejection in the flashing section, a make-up water stream is needed. The mass balance for the mixing point is written as:

$$\dot{m}_{FC, MK} = \dot{m}_{FC, F} - \dot{m}_{FC, R}$$  \hspace{1cm} (3-46)

where from section 3.2.2.1:

$$\dot{m}_{FC, F} = \dot{m}_{c, \text{hex}}$$  \hspace{1cm} (3-46-1)

and from the flashing chambers, we know that:

$$\dot{m}_{c, \text{hex}} = \dot{m}_{FC, j, out} + \dot{m}_{V, FC, \text{total}} = \dot{m}_{FC, j, out} + \sum_{i=1}^{j} \dot{m}_{V, FC, i}$$  \hspace{1cm} (3-46-2)
For the rejection point, we have:

\[ m_{FC,R} = m_{FC,j,out} - m_{FC,\text{drain}} \]  

(3-47)

where:

\[ m_{FC,\text{drain}} = m_{FC,j,out} \cdot DR \]  

(3-47-1)

where \( DR \) is the drain ratio which is guessed by the solver (see section 3.2.4), therefore:

\[ m_{FC,R} = (1 - DR) \cdot m_{FC,j,out} \]  

(3-48)

Which by knowing the drain ratio \( DR \) and the outlet mass flow rate from the last flashing chamber, the amount of recycled brine mass flowrate is calculated.

The salinity balance for the mixing point is written as:

\[ m_{FC,\text{MK}} \cdot X_{FC,\text{MK}} = m_{FC,F} \cdot X_{FC,F} - m_{FC,R} \cdot X_{FC,R} \]  

(3-49)

In our seawater application a 70,000 ppm is considered for \( X_{FC,R} \) [2]–[4]. This is the maximum permissible value of the outlet salinity in order to prevent calcium sulfate formation [27]. \( X_{FC,\text{MK}} \) is the salinity of the make-up water after deaeration which to a good approximation it can be considered to equate to \( X_C \) (see section
Chapter 3. Flash Boosted MED, the New Novel Design

3.2.2.4) that in our considered seawater application is 35,000 ppm [2]–[4]. Therefore by knowing the number of flashing chambers and the relevant overall temperature difference across the flashing chambers and with reference to the relevant mass flowrate from equations (3-46) to (3-49) as the relevant boundary conditions, \( X_{FC,F} \) can be calculated by the solver.

3.2.2.4. Deaerator

As mentioned above, the make-up water stream in the brine recirculation system is needed to compensate both water losses due to the evaporation and brine rejection. This make-up water should be properly deaerated and vented to avoid accumulation of \( NCG \) s, otherwise these \( NCG \) s are then released in the flashing chambers and transferred to the primary MED effects through the injection lines which will decrease the heat transfer coefficient in the heat exchangers surfaces and decrease the thermal performance of the process [127]. For the purpose of a proper venting system, a deaerator is needed to remove a portion of these dissolved gasses from the feed seawater inlet to the flashing chamber (Figure 3-13) to attenuate the venting power consumptions. Moreover, deaeration will reduce the oxygen content of the make-up water which will cause corrosion problems in the other devices [128]. Nowadays, the deaerator is a rectangular cross section vessel, including an empty space with a lower pressure at the top, where the flash takes place, and a packed section with Pall rings at the bottom, where the striping process occurs [129]. The feedwater is fed from the top to this flashing vessel, and then distributed onto the corrugated plate while the stripping steam is fed from the bottom and comes into contact with the distributed feed in a counter current direction [128]. A recent industrial experimental investigation showed the possibility of operating without stripping steam feeding [129], which has been considered in our seawater application.

In our seawater application, a 5.9 kPa operating pressure [129] has been considered for the deaeration vessel, which due to the low equivalent temperature drop (2°C to 3°C in our seawater application) to a good approximation the flashing evaporation rate can be ignored in our deaerator [2]–[4]. Therefore there is no mass flowrate change, and the salinity also remains constant (Figure 3-13), means:
\[ X_{FC, MK} \approx X_C \] (3-50)

The temperature drops, therefore the enthalpy changes from \( h_{C, out} \) to \( h_{FC, MK} \)

\[ h_{C, out} = h_f(T_{C, out}, X_C) \] (3-51)

\[ h_{FC, MK} = h_f(T_{FC, MK}, X_C) \] (3-52)

3.2.3 Overall mass, salinity and energy balances

The overall mass, salinity and energy balances are essential steps in the mathematical simulation. As shown in Figure 3-14, for an FB-MED, there are two inlets (the main heat source inlet and the cooling water inlet streams) and five outlets (heat source outlet, freshwater outlet, brine outlets from the primary MED and flashing chambers, and cooling water outlet). The overall mass balance is calculated as:

\[ \sum \dot{m}_{inlets} = \sum \dot{m}_{outlets} \] (3-53)

\[ \dot{m}_{C,in} = \dot{m}_{C, out} + \dot{m}_{B, total} + \dot{m}_{FC, drain} + \dot{m}_{D, total} \] (3-54)

where, \( \dot{m}_{B, total} \) is the total outlet brine which is extracted from the primary MED section; \( \dot{m}_{FC, drain} \) is the rejected brine from the flashing section; \( \dot{m}_{D, total} \) is the total freshwater production rate of the process which is:

\[ \dot{m}_{D, total} = \sum_{k=1}^{n} \dot{m}_{V,k} + \sum_{i=1}^{j} \dot{m}_{V',i} \] (3-54-1)

and:

\[ \dot{m}_{C,in} = \dot{m}_C \] (3-54-2)
Chapter 3. Flash Boosted MED, the New Novel Design

\[ \dot{m}_{C,\text{out}} = \dot{m}_C - \dot{m}_{F,\text{total}} - \dot{m}_{FC,\text{MK}} \]  

(3-54-3)

The overall salinity balance is:

\[ (\dot{m}_{C,\text{in}} - \dot{m}_{C,\text{out}}) \cdot X_C = \dot{m}_{B,\text{total}} \cdot X_{B,\text{total}} + \dot{m}_{FC,\text{drain}} \cdot X_{FC,\text{drain}} \]  

(3-55)

At the end, for the energy balance we have:

\[ \sum E_{\text{inlets}} = \sum E_{\text{outlets}} \]  

(3-56)

\[ \dot{m}_{HS,1} \cdot (h_{HS,1,\text{in}} - h_{b,\text{hex,\text{out}}}) = \dot{m}_{C,\text{out}} \cdot h_{C,\text{out}} - \dot{m}_{C,\text{in}} \cdot h_{C,\text{in}} + \dot{m}_{FC,\text{drain}} \cdot h_{FC,\text{drain}} + \dot{m}_{B,\text{total}} \cdot h_{B,\text{total}} + \dot{m}_{D,\text{total}} \cdot h_{D,\text{total}} \]  

(3-57)

3.2.4 Solving procedure

The general solving procedure flow chart is shown in Figure 3-15. The GRG method (Generalized Reduced Gradient) [115] is used to solve all above-mentioned equations by imposing the boundary conditions from the law of thermodynamics and appropriate operational, technical and economic constraints as explained in the following chapters [2]–[4], [98].

The assumptions of thermodynamic simulation are listed in table 3-1. By using these assumed parameters, and relevant boundary conditions which will be described in the next chapters, the solver will produce the results such as the total freshwater production rate (\(\dot{m}_{D,\text{total}}\)), total highly concentrated flowrate, which
Chapter 3. Flash Boosted MED, the New Novel Design

is important in alumina refinery process, waste heat performance ratio \( PR_{WH} \), temperature profiles across the effects and \( UA \) values. These technical results then are used for the thermo-economic evaluation as explained in section 3.4.

<table>
<thead>
<tr>
<th>Item</th>
<th>Parameter</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>( T_{HS,1,in} )</td>
<td>Heat source inlet temperature (°C)</td>
</tr>
<tr>
<td>2</td>
<td>( m_{HS} )</td>
<td>Heat source mass flow rate (kg/s)</td>
</tr>
<tr>
<td>3</td>
<td>( T_{C,in} )</td>
<td>Cooling water inlet temperature (°C)</td>
</tr>
<tr>
<td>4</td>
<td>( X_{C,in} )</td>
<td>Feedwater salinity (ppm)</td>
</tr>
<tr>
<td>5</td>
<td>( R )</td>
<td>Feed to vapour ratio</td>
</tr>
<tr>
<td>6</td>
<td>( TBT )</td>
<td>Top Brine Temperature (°C), which is 70°C for MED process</td>
</tr>
<tr>
<td>7</td>
<td>( \Delta P_{inj} )</td>
<td>Pressure difference for steam injection purpose (Pa)</td>
</tr>
<tr>
<td>8</td>
<td>( Min \Delta T_e )</td>
<td>Minimum temperature difference across each MED effect (°C)</td>
</tr>
<tr>
<td>9</td>
<td>( Min \Delta T_{HS,1} )</td>
<td>Minimum heat source temperature drop across the first effect (°C)</td>
</tr>
<tr>
<td>10</td>
<td>( Min \Delta T_{hex} )</td>
<td>Minimum temperature drop across liquid-liquid heat exchanger (°C)</td>
</tr>
<tr>
<td>11</td>
<td>( \Delta T_{approach} )</td>
<td>Liquid-liquid heat exchanger approach temperature (°C)</td>
</tr>
<tr>
<td>12</td>
<td>( Min \Delta T_c )</td>
<td>Minimum temperature difference across the condenser (°C)</td>
</tr>
<tr>
<td>13</td>
<td>( X_{max} )</td>
<td>Maximum allowable outlet salinity (ppm), which is 70,000 (ppm) in seawater application</td>
</tr>
<tr>
<td>14</td>
<td>( X_{FC,j,\text{out}} )</td>
<td>Flashing section brine outlet salinity (ppm)</td>
</tr>
<tr>
<td>15</td>
<td>( m_{vent} )</td>
<td>Venting flowrate which is considered as a percentage of ( m_{HS,\text{cond}} )</td>
</tr>
<tr>
<td>16</td>
<td>( P_{deaerator} )</td>
<td>The deaerator operating pressure</td>
</tr>
</tbody>
</table>
3.3 Pumping Power Analysis:

Electrical power consumption is one of the most important considerations, stemming primarily from pumping power consumption in desalination plants. Pumping power is calculated according to the equation below [2].

\[ \text{Pumping Power} = \sum \text{Pump Efficiency} \times \text{Flow Rate} \]
Chapter 3. Flash Boosted MED, the New Novel Design

\[ Pumping\ \text{Power} \ (kW) = \frac{\Delta P \ (kPa) \cdot V \ (m^3/s)}{\eta_{\text{pump}} \cdot \eta_{\text{motor}}} \]  

(3-58)

where, \( \eta_{\text{pump}} \) and \( \eta_{\text{motor}} \) are the pump and motor efficiencies and equate to 0.7 and 0.9, respectively in the present simulation.

\[ p \ (kWh/m^3) = \frac{\text{Pumping Power} \ (kW)}{M_{D,\text{total}} \ (m^3/h)} \]  

(3-59)

where \( p \) is the specific pumping power in \( kWh/m^3 \) and \( M_{D,\text{total}} \) is the total production rate of the plant in \( m^3/h \).

As schematically shown in Figures (2-1), (2-3) and (3-1), in conventional, boosted and flash boosted MEDs, the pumping power accounts for the following [2]:

3.3.1 Saline water pump

This pump conveys saline water through the condenser of the primary MED and feeds the MED effects. The pressure drop across the condenser is considered to be 0.5 bar [2]. The overall pressure difference in this pump is taken to be 1.5 bar (0.5 bar for the condenser and 1.0 bar to access the atmosphere) [2].

3.3.2 Brine recirculation pump

This pump is used in the flash boosted MED system to recirculate the brine through the brine heater (liquid-liquid heat exchanger) and the flashing chambers. Sommariva et al. [62] showed the brine recirculation pump head contributions in the Al-Taweelah B plant that are related to the tube bundles, water boxes, brine heater, spray pipe, control valves, pipeline, geodetical and pressure heads. All the pressure drops, save those associated with the tube bundles and water boxes which are absent in the present flashing chamber section, and the reported pressure head, are adopted from; this accordingly amount to 1.7 bar [62]. The germane pressure head is calculated based on the flashing chamber inlet and outlet saturated pressure difference.

3.3.3 Distillate extraction and brine blowdown pumps

These pumps are used to extract the distillate (freshwater) and brine from the primary MED plant. For these pumps a 2 bar differential pressure is considered sufficient [2].

Bijan Rahimi
3.3.4 Heat source medium pump

This pump is used for pumping the heat source liquid through the evaporator (first effect) of the primary MED plant and the liquid-liquid heat exchanger (for the flash boosted MED). A 0.5 bar drop in the evaporator and a 0.5 bar in the liquid-liquid heat exchanger has been considered for FB-MED process (Figure 3-1) [2]. The total pressure drop for the conventional MED is accordingly 0.5 bar [2] that is related to the evaporator’s pressure drop (Figure 2-1). For B-MED processes (Figure 2-3) a 1 bar pressure drop has been considered for the two evaporators [2].

3.3.5 Drain pump

The drain pump is used for brine rejection from the brine recirculation stream of the flash boosted MED. The pumping power of this stream is negligible because of the small flow rate. However, a 2 bar pressure differential has been considered for this pump for pumping from vacuum to the atmospheric pressure [2].

3.3.6 Make-up water pump

The make-up water pump is used to provide sufficient pressure to the make-up water that exits the deaerator. A 2 bar differential pressure has been considered for this pump [2].

3.3.7 NCG extraction vacuum pump

One of the important issues for low grade sensible heat driven desalination processes is related to the venting process, which serves to remove the generated non-condensable gasses \( \text{(NCG)} \) and the air leakage into the system. In the steam heat driven desalination process, such as TVC-MED or MSF plants, steam ejectors are used for vacuuming and non-condensable gas removal; but in low grade sensible heat driven application there is no pressurized steam for driving ejectors, and so two options are available. The first option is to use water eductors or air ejectors, and the second is to use water ring vacuum pumps. In the present simulation, the second option is adopted [2]–[4] and the corresponding pumping power consumption is evaluated according to the manufacturer’s catalogues for water-ring vacuum pumps [130]. The amount of non-condensable gas released is a function of feedwater composition, evaporator
design and working conditions. Literatures expounding the methods for estimating the amount of non-condensable gasses abound [131]–[133], but in terms of design, it is typical to consider an over designed venting system for handling the air leakage into the system [134]. As a matter of comparing between the conventional MED and flash boosted MED, a cascaded venting scheme is considered for both systems which means that the non-condensable gases are extracted from each effect and cascaded downstream and eventually into the condenser where all accumulated $NCG$s are extracted by the vacuum pump. On account of the low top brine temperature and the make-up/brine recirculation ratio of the present flashing chambers and also the use of deaerated make-up water, the amount of generated $NCG$s in the present flashing chambers as compared to the released $NCG$s in the primary MED effects is relatively low. In the present simulation the amount of $NCG$ released in the flashing chambers is considered based on the ratio between the make-up water of the flashing chambers and the feedwater of the primary MED plant that is varied between 15 to 25% (for seawater application) [2]–[4]. Therefore a venting system is considered for handling all $NCG$s released from both the MED plant and flashing chambers. Following routine considerations the venting system is designed to remove 1% of the vapor in the condenser as $NCG$s [135].

### 3.4 Thermo-Economic Analysis

Many works have reviewed or evaluated the economics of desalination technologies. Hitherto the focus of such works predominantly being a review the state of the art [14], [53], [80], [136]–[148]. Less frequent are reports on the economic viability of upcoming technologies that have yet to reach the stage of prototyping. Furthermore, most authors focused on the unit cost of distillate when comparing the economy of desalination technologies, with less priority given to annual cash flows, capital investment requirements and operating costs as a consequence. According to Badiru and Newnan [149], [150], greater importance should be placed upon cash flows and the inherent monetary value of plant installations, implying that capital budgeting metrics such as net present value ($NPV$) and internal rate of return ($IRR$) are of greater use for comparison of engineering projects than estimates of unit product cost. This section reports a generalised method which was used to quantitatively compare the economic
value of desalination installations using MED, B-MED and FB-MED, via such metrics as $NPV$ and $IRR$ [3].

Capital budgeting metrics such as net present value ($NPV$), internal rate of return ($IRR$) and incremental rate of return ($\Delta IRR$) can be used to quantitatively assess whether an investment has greater economic value than an alternative investment. By assuming that all conventional MED, B-MED and FB-MED installations are subject to the same site-specific conditions, the use of these metrics can be extended to assess which process is more ‘economical’ for the assumed conditions. To produce such metrics, an estimate of the annual cash flows is necessary; hence estimates of the initial investment, as well as annual operating expenses are required. The following builds upon the method for estimating capital costs previously reported in [2].

3.4.1 Capital cost

The initial investment required for a desalination plant is assumed to be equal to the total capital cost of the plant, which is assumed to be inclusive of such costs as engineering, commissioning, erection, equipment, instrumentation, electronics and control. Reputable data of EPC (Engineering, Procurement and Construction) costs is available for conventional MED and MSF desalination plants from the GWI Desalting Plant Inventory [55], and an estimate of total plant capital cost as a function of production capacity can be interpolated from such data [2], [3]. For this analysis, IDA census data [55] were deemed suitable for the purposes of a generalised analysis. A review by Park et al. [141] asserts that location variance has little appreciable effect upon the cumulative costs of a desalination plant, hence all data was considered regardless of plant location. The spread of data pertaining to the capital costs of MSF and MED plants indicated that a regression function could be derived for both MED and MSF plants as a function of daily plant capacity in the bounded region which is shown in Figure 3-16 and highlight the dense conservative limit of data points [3]. To improve the accuracy of the regression analysis, obvious outliers were removed, leaving only the conservative values within the upper limit of the high density region. The regression equation obtained for the remaining data therefore generates a conservative estimate of capital costs for plants of production...
capacity up to 10,000 m$^3$/day - a range that encompasses the capacities of all the simulated cases. Equation (3-60), obtained via the regression analysis on the adjusted data gives an estimate of the capital cost of both MED and MSF plants as a function of their respective daily production capacities [3].

Equation (3-60) specifies the cost of the FB-MED plant.

$$TCC_{MED,MSF} (US$) = \Psi_{Dt} = 3054 \times D_t^{0.9751}$$  \hspace{1cm} (3-60)$$

where $TCC$ is the total capital cost of the desalination plant and $D_t$ is the total production rate of a conventional plant in m$^3$/day. This function ($R^2 = 0.994$) in the bounded region of the relevant data (which is shown with two dotted lines) shows that the conservative capital cost can be considered as a function of plant production capacity in this region [2], [3].

Following the same methodology as Rahimi, et al. [2], [3], the capital cost of a FB-MED plant is estimated to be equal to the sum of cost of the primary MED section and the flashing chambers, with an augmenting cost factor to account for the increased heat transfer requirements of the MED effects due to supplementary steam injection. Equation (3-61) specifies the cost of the FB-MED plant. The first term relates to the primary MED effects of the FB-MED, based upon the vapour production of these effects. The second term contained within brackets is an augmenting factor that accounts for the increased heat transfer requirements of the primary MED effects due to vapour injection; and the third
term specifies the cost of the flashing chambers which augment the vapour production of the system.

\[
TCC_{FB-MED}(\text{US$}) = \Psi_{D_t,FB-MED-\Delta P_{FC}} + \left[0.20 \cdot (\Psi_{D_t,FB-MED} - \Psi_{D_t,FB-MED-\Delta P_{FC}})\right] + 0.41 \cdot \Psi_{D_{FC}} \quad (3-61)
\]

Similarly the total capital cost of a B-MED plant is approximated as the sum of the capital cost of the primary MED section, specified as the first term of Equation (3-62) [3], augmented for the increased heat transfer requirements by the second term, and the cost of the booster unit is then specified by the third term. The cost of booster unit is approximated as the cost of an MED plant with a daily production capacity equal to the production capacity of the booster unit itself. This cost is then discounted by a factor of 69%, as it is assumed that the cost of the booster unit is inclusive only of evaporator, equipment and piping costs [2], [3], [151].

\[
TCC_{B-MED} \quad (\text{US$}) = \Psi_{D_t,B-MED-\Delta P_{Booster}} + \left[0.20 \cdot (\Psi_{D_t,B-MED} - \Psi_{D_t,B-MED-\Delta P_{Booster}})\right] + 0.69 \cdot \Psi_{Booster} \quad (3-62)
\]

### 3.4.2 Operating costs

Annual operating expenses constitute the remaining costs of a desalination facility. The annual operating expenses were assumed to be comprised of electrical, labour, chemical additive, maintenance and spares, and insurance costs. Constancy of operating conditions, external economic conditions and plant reliability is assumed, such that annual operating expenses remain constant over a plant’s useful life, albeit with an assumed escalation factor to account for inflation in the cost of goods and services.

**3.4.2.1 Electricity**

Electrical expenses are approximated to the cost of electricity consumed by all process pumps, as these should constitute the main contribution to the total electrical energy consumption of a desalination plant. The rate of consumption of electricity is known for each process pump owing to existing process simulations [2]. The annual electrical cost \(AEC\) is specified by equation (3-63) [3].

\[
AEC \quad (\text{US$} /\text{yr}) = EUP \left(\frac{\text{US$}}{\text{kWh}}\right) \cdot p \left(\frac{\text{kWh}}{\text{m}^3}\right) \cdot D_t \left(\frac{\text{m}^3}{\text{day}}\right) \cdot f \cdot 365 \left(\frac{\text{day}}{\text{yr}}\right) \quad (3-63)
\]

where the Electricity Unit Price, \(EUP\), is assumed to have a value of 0.151 AU$/kWh based upon recently published Australian rates [3], [152], which was
converted to US$ based upon recent exchange rates [153] of 1.086 AU$ per US$.
The plant availability, \( f \), is assumed to be 95% per year [2], [3].

Referring to section 3.3 and the relevant Figures (2-1), (2-3) and (3-1) it can be seen that the MED, B-MED and FB-MED processes all require five pumps inclusive of the saline water intake pump, heat source pump, non-condensable gas (vacuum) pump, distillate pump and brine blowdown pump. The improved thermal efficiency and production rates of the B-MED are enabled by an increased duty requirement of these pumps. The improved FB-MED process is enabled by increased duty on these pumps, as well as by the addition of brine recirculation pump, make-up water pump and drain pump which service the flashing chambers which boost the production of the primary MED effects. All details related to the pumping power calculation including the efficiencies and relevant pressure heads have been explained in detail in section 3.3 [2].

3.4.2.2 Labour

Labour costs are based upon actual cost data, and the estimate is made that one full time equivalent of skilled personnel is required per 4543 m\(^3\)/day plant capacity [3]. The average cost of such a water services sector employee in Australia is 97,905 US$ per year [3], [154]. All simulations considered for the purposes of this evaluation have production rates below 4543 m\(^3\)/day, hence it is assumed the labour cost is fixed at the prescribed value [3].

\[
\text{Labour Cost}_{\text{Annual}} \left( \frac{\text{US}\$}{\text{yr}} \right) = 97905 \left( \frac{\text{US}\$}{\text{yr}} \right) \cdot f \quad (3-64)
\]

3.4.2.3 Chemicals

Chemical additives are necessary for ensuring the effectiveness and maintainability of the desalination process, the potability of product, and constitute a necessary annual expense. The primary MED feedwater has specific chemical dosing requirements, constituting the total chemical expense of a conventional MED. The FB-MED has additional chemical requirements due to the required chemical treatment of the flashing chamber feed. The specific chemical dosing cost of the primary MED feed is considered to be 0.0223 US$/m^3$ based upon values from [3], [27], [145], and the specific chemical cost for the flashing chamber feed is assumed to be 0.0198 US$/m^3$ based upon values from [3], [27],
Chapter 3. Flash Boosted MED, the New Novel Design

[138], [144], [145]; assuming that the chemical requirements of the flashing chamber feed are similar to the feed of MSF plant evaporators at low temperature.

3.4.2.4 Maintenance, spares and insurance

The yearly cost of maintenance activities, spares and insurance is assumed to be 1.5% of the total capital cost of the plant [155], as specified by equation (3-65) [3]. The cost of maintenance and spares of any desalination plant are highly dependent upon factors such as the timing of cash flows as well as the proficiency of engineering services and management. Therefore the annual costs of a plant utilising the well-known MED process could differ greatly to the cost of similar plants utilising the FB-MED or B-MED process, especially in the infancy of these technologies. However, without any justifiable alternative, such costs are approximated as a percentage of the total capital cost of the plant [3].

\[
\text{Maintenance, Spares, Insurance Cost (\text{US$ yr})} = 1.5\% \times TCC \tag{3-65}\]

where the total capital cost, \(TCC\), is determined as a function of production capacity in the manner described in section 3.4.1.

3.4.3 Cash flows and capital budgeting metrics

Annual operating expenses, income taxes, loan repayments and incomes are assumed to comprise the annual cash flows over the lifetime of the desalination plants, with all incomes and expenses assumed to increase at a rate of 3% per year [3] from the starting year to account for inflation. Cash flows are then used to determine capital budgeting metrics so that the processes can be compared. It is assumed that the initial investment is equal to the total capital cost of the desalination plant, and this cost is incurred exhaustively at the beginning of the plant’s useful life.

3.4.3.1 Incomes

It is assumed that the incomes of the desalination plant in a period are equal to the incomes made on the sale of the volume of water equal to the total production of the plant within that same period, sold at the average market price [3].
where \( WMP \) is the market price of water product (distillate) available from an external supplier, assumed as being equal to the Australian average water price of 2.72 AU$/m^3 [156], which is then converted into US$ based upon recent exchange rates [153]. A 3% escalation rate is applied to calculate the income for the following years [3].

### 3.4.3.2 Costs

Major costs include operating cost, loan repayment and income tax. Operating cost has been explained in section 3.4.2. Simple straight line depreciation is assumed to apply to the capital cost of the plant. For the loan repayment, it is assumed that a bank loan covers 50% of the cost of the initial investment, equal to the total capital cost of the plant, with a 2% supplementary fee to cover miscellaneous bank fees, and with interest incurred annually at a rate of 8% upon borrowed money [3]. Annual fixed payments (Equation 3-67) are made to cover the cost of loan repayment, assuming the loan is to be fully paid off by the end of the plant’s useful life, which is assumed to be 30 years for all cases. Equation (3-68) specifies the capital recovery factor, \( CRF \), a factor which enables the calculation of a single amortising value for an amount that is to be fully amortised within ‘\( n \)’ years, and which incurs interest at an annual rate, \( i \) [27], [144], [150].

\[
\text{Loan Repayment} = (1 + S) \cdot F \cdot TCC \cdot CRF
\]  

\[
CRF = \frac{i(1+i)^n}{(1+i)^n-1}
\]  

Income tax is calculated assuming the tax rate is a flat 30% of taxable income [157] (that in this case is the summation of income, operating cost, depreciation and bank interest); however, income tax will often affect the alternatives similarly, allowing to compare the choices without considering income taxes. The total cash flow in a year is then equal to the difference of the total annual benefits and total annual expenditures, adjusted for inflation and for tax.

With cash flows specified over the lifetime of each plant design, the economic metrics can then be calculated. These metrics, often referred to as capital budgeting metrics, allow investments and projects to be compared for their
economic value. Net Present Value (NPV) is widely considered to be the superior and most appropriate metric for ranking the economic value of investments and projects [149], [150], [158], and it is taken as the primary means of comparison in this analysis. NPV, as specified by equation (3-69) [150], is the sum of the annual cash flows of a project or investment, discounted to a present day value by means of an interest rate, i.

\[ NPV = \text{Present Value of Benefits} - \text{Present Value of Costs} = \sum_{yr=1}^{n} \frac{\text{Cash Flow}_{yr}}{(1+i)^{yr}} - TCC \]  
(3-69)

The Incremental Rate of Return (ΔIRR) metric, and incremental analysis can be the preferred method of comparing investments and engineering project [149], [150]. Incremental analysis compares the cash flows of two specific investments, or engineering projects, over the lifetime of the projects. The incremental rate of return, ΔIRR, is the value of the interest rate which yields an NPV of zero for the differential cash flows of two projects [150]. The calculated rate of return is compared with the preselected Minimum Attractive Rate of Return (MARR), and if higher, the higher investment alternative is the economically superior choice; otherwise the lower investment is the superior choice [150]. The MARR is assumed to be equal to the interest rate on borrowed money of 8%, in projects accompanied by normal business risk and uncertainty [3], [150].

The unit product cost (UPC) of water (Equation 3-70) is also used as an appropriate metric for comparing desalination plants and processes [17], [27], [137], [140], [142], [144]. The use of the UPC metric is appropriate for ranking the production costs of desalination plants; however, the plant with the lowest specific production costs is not inherently the plant that will be the most profitable. Taxation, asset depreciation, interest charges and other factors too affect the profitability of an engineering project. A desalination plant is a monetary investment, as well as an engineering endeavour, and therefore it must be framed and compared to alternatives in terms of profitability and worth. Pasqual [158] asserted that the consensus on ranking monetary investments was that the metrics such as NPV were the best means of ranking investment alternatives.

\[ UPC = \frac{TCC \times CRF + OPEX}{365 \cdot f \cdot D_t} \]  
(3-70)
Chapter 3. Flash Boosted MED, the New Novel Design

where $OPEX$ refers to the operating expense incurred in the first year of operation, and the 3% escalation rate is being ignored here [3].
Chapter 3. Flash Boosted MED, the New Novel Design
Chapter 4

Application of the Flash Boosted MED Process in Seawater Desalination

The novel FB-MED desalination processes that utilise low grade sensible heat sources has been modelled and shown to be more thermally efficient means of desalinating water than the conventional MED and the Boosted MED processes. The novel Flash Boosted MED (FB-MED) process is capable of higher production rates than both conventional and boosted MEDs, enabled by the addition of process components and by an increase in specific electrical power consumption. A simple method of estimating the capital and operational costs of MED, B-MED and FB-MED desalination installations has been presented in the previous chapter. A generalised comparison of the economics of these three processes is conducted, asserting the economic viability of the novel desalination processes.

Highlights:

- The novel FB-MED process maximally exploit the potential of waste sensible heat source
- Up to 50% more freshwater yield compared with the conventional MED process
- 7% reduction in the specific capital cost as compared to the conventional MED process
- The unit product cost of FB-MED is 6% less than the conventional MED process
- The net present value of FB-MED is 53% greater than the conventional MED’s
- FB-MED is both technically and economically superior to both conventional and boosted MEDs

Publication:

This Chapter consists of the result sections of my below publications [2]–[4]:

Chapter 4. Application of the Flash Boosted MED Process in Seawater Desalination

4.1 Introduction

Process simulations and validation of conventional MED, B-MED and FB-MED were produced based upon steady-state analysis methods used in combination with appropriate boundary conditions (Table 4-1) [2], [95], [110]. The relevant equations and a detailed explanation are available in Chapter 3 [2], [95], [96]. The results of the process simulation and thermo-economic evaluation are presented in this chapter. They indicate that the freshwater production of the FB-MED process is indeed greater than the optimised conventional MED and the optimised B-MED alternatives, for a range of design cases as shown in Table 4-2 [2]. Furthermore, results indicate that the waste heat performance ratio, \( PR_{WH} \) [2], [96], improves in much the same manner as the production rate, indicating that the novel FB-MED process better utilises the energy available in the heat source fluid. This asserts the greater thermal efficiency of the novel process, which can also be typified by an increased production rate which, however, is gained via an increase in the specific electrical power consumption.

<table>
<thead>
<tr>
<th>Table 4-1 Assumptions of thermodynamic simulations [3]</th>
</tr>
</thead>
<tbody>
<tr>
<td>Maximum top brine temperature (°C)</td>
</tr>
<tr>
<td>Heat source flow rate, ( m_{HS} ) (kg/s)</td>
</tr>
<tr>
<td>Heat source temperature (°C)</td>
</tr>
<tr>
<td>Feed to vapour ratio (35% evaporation)</td>
</tr>
<tr>
<td>( T_{cond,\text{in}} ) (°C) / ( T_{cond,\text{out}} ) (°C)</td>
</tr>
<tr>
<td>( \Delta P_{\text{inj}} ) (Pa)</td>
</tr>
<tr>
<td>( BPE+NEA ) in each flashing chamber (°C)</td>
</tr>
<tr>
<td>Liquid-to-Liquid heat exchanger approach temperature (°C)</td>
</tr>
<tr>
<td>Minimum temperature drop for the heat source medium (°C)</td>
</tr>
<tr>
<td>( X_F ) (ppm), Feed salinity</td>
</tr>
<tr>
<td>( X_{BFC} ) (ppm), Flashing chamber outlet salinity</td>
</tr>
</tbody>
</table>

4.2 Simulation results

Table 4-1 delineates all the assumptions inherent in the present simulation. The heat source medium is water and its flow rate has been fixed at 100 kg/s in all simulations. The inlet seawater temperature at the condenser is 28°C and the salinity of the seawater is 35000 ppm. The top brine temperature is considered to be 70°C, so that the top brine temperature is capped at 70°C. For the liquid-to-liquid heat exchanger the minimum temperature of approach is taken to be 3°C. For vapor injection to happen from the flashing chamber to the judicious effect in
the primary MED plant, a 500 Pa pressure difference is considered to be sufficient [96].

Figures 4-1 to 4-3, show the schematic layout of optimised MED, B-MED and FB-MED systems which are coupled to a 65°C sensible heat source [3]. All results including the production rate, number of effects, pumping power consumptions and waste heat performance ratio for a range of sensible heat source temperature between 65°C and 90°C are shown in Table 4-2 [3]. Table 4-3 also includes the main flowrates of the considered processes [3].

As shown in Table 4-2, the performance of the flash boosted MED scheme with those of the optimised conventional and boosted MED processes has been compared. The number of effects for the last three schemes as in Table 4-2 (namely optimized conventional, boosted MED and flash boosted MED) generally increases with temperature, so as to maximize freshwater production. Subject to the same heat source inlet temperature, the heat source temperature drop across the primary MED plant, for the three schemes presently considered, is generally different so that each technology maximises its respective freshwater production. Specifically, the heat source temperature drop across the primary plant is generally the smallest for the flash boosted MED scheme, as the remaining heat source enthalpy can be better exploited at the flashing chambers. This is then followed by the boosted MED process, whereby the steam booster unit can still further process the remaining waste heat. The largest heat source temperature drop is verily found across the optimised conventional MED plant, as this is the
only avenue to utilize the waste heat. While more effects can certainly be built into the conventional MED system at any particular heat source inlet temperature, as delineated in Table 4-2, this is done at the expense of freshwater production, as less heat via a temperature drop can be exploited [96]. Consequently both the boosted and flash boosted MED systems can realize a successively larger number of effects than the conventional MED system for the same heat source inlet temperature. For clarity, in Figure 4-1 for the conventional MED plant, the important process temperatures stemming from the simulation are included. The same is done to both the boosted MED scheme as in Figure 4-2, and flash boosted MED scheme as in Figure 4-3.

Figure 4-2 Schematic design of an optimised Boosted MED (B-MED) system (typical quantities as per the simulation for 65°C inlet heat source temperature) [3]

Figure 4-4 highlights the characteristics of the flash boosted MED scheme in relation to the conventional and steam boosted MED processes. Over the range of temperatures from 65°C to 90°C, the flash boosted scheme is 10% to 20% better in terms of freshwater production than the boosted MED process. This potential is better discerned where the production increment rates of both the flash boosted MED and boosted MED schemes are compared against the conventional MED system. The production rate increment decreases gradually for the boosted MED process from around 22% at 65°C heat source inlet temperature to close to 13% at 90°C. For the flash boosted MED scheme, on the other hand, the production rate increment begins at around 34% at 65°C and peaks at about 43% at 75°C, and then pares back to around 33% at 90°C. The
reduction is enjoined by the top brine temperature constraint of 70°C at the first effect of the primary MED plant and forces the number of primary MED effects to remain the same for heat source temperatures higher than 80°C [2].

As shown in Figure 4-4, the freshwater production rate of the FB-MED process over the entire heat source temperature range is more than both the optimised conventional and boosted MED. The B-MED enjoys its highest thermal performance in the lower temperature range while the FB-MED performance first increases till 80°C heat source temperature and then slightly decrease; however over the entire temperature range it remains the superior technology.

Insofar as low grade heat applications are concerned, the conventional performance ratio (See section 2.4) which holds that heat comes with a premium as it is consumed does not capture the essence of a desalination system driven by such heat sources, which only attracts a one-off investment cost, instead of an ongoing cost of thermal energy used [2]. A waste-heat performance ratio [96], or \( PR_{WH} \), as in equation (2-2), is used instead which encourages the maximal use of the enthalpy of the low grade heat source relative to the heat-sink. Figure 4-5 shows the waste-heat performance ratio for all three systems. From this figure, the waste-heat performance ratio of the flash boosted MED experiences a

---

Figure 4-3 Schematic design of an optimised Flash Boosted MED (FB-MED) system, (typical quantities as per the simulation for 65°C inlet heat source temperature) [3]
substantial jump at 75°C. This mirrors the boost in production rate at 75°C as shown in Figure 4-4 [2].

<table>
<thead>
<tr>
<th>Heat source inlet temperatures (°C)</th>
<th>65</th>
<th>70</th>
<th>75</th>
<th>80</th>
<th>85</th>
<th>90</th>
</tr>
</thead>
<tbody>
<tr>
<td>Optimised Conventional MED</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Freshwater yield (m³/day)</td>
<td>707</td>
<td>998</td>
<td>1341</td>
<td>1694</td>
<td>2103</td>
<td>2573</td>
</tr>
<tr>
<td>Normalised Pumping Power (kWh/m³)</td>
<td>2.02</td>
<td>1.95</td>
<td>1.56</td>
<td>1.51</td>
<td>1.27</td>
<td>1.09</td>
</tr>
<tr>
<td>Number of MED effects</td>
<td>4</td>
<td>4</td>
<td>5</td>
<td>5</td>
<td>6</td>
<td>7</td>
</tr>
<tr>
<td>Heat source temperature drop (°C)</td>
<td>12.9</td>
<td>18.2</td>
<td>20.5</td>
<td>25.8</td>
<td>28.2</td>
<td>31.2</td>
</tr>
<tr>
<td>$PR_{WH}$</td>
<td>1.18</td>
<td>1.47</td>
<td>1.77</td>
<td>2.03</td>
<td>2.30</td>
<td>2.59</td>
</tr>
</tbody>
</table>

| Optimised Boosted MED               |     |     |     |     |     |     |
| Freshwater yield (m³/day)           | 865 | 1217| 1603| 2034| 2462| 2920|
| Normalised Pumping Power (kWh/m³)  | 2.39| 1.82| 1.63| 1.46| 1.37| 1.26|
| Number of MED effects/Injected Effect’s Number | 5/5 | 6/5 | 6/5 | 7/6 | 7/6 | 8/7 |
| Production Rate Ratio*              | 22% | 22% | 20% | 20% | 17% | 13% |
| Normalised Power Consumption Ratio* | 18% | -7% | 4%  | -3% | 8%  | 15% |
| Heat source temperature drop (°C)   | 17.6| 19.7| 24.7| 29.8| 34.8| 39.9|
| $PR_{WH}$                           | 1.44| 1.80| 2.12| 2.43| 2.69| 2.94|

| Optimised Flash Boosted MED         |     |     |     |     |     |     |
| Freshwater yield (m³/day)           | 945 | 1417| 1922| 2426| 2928| 3430|
| Normalised Pumping Power (kWh/m³)  | 3.21| 2.42| 2.01| 1.75| 1.60| 1.50|
| Number of MED effects / Flashing stages | 5/4 | 7/6 | 9/8 | 10/9| 10/9| 10/9|
| Production Rate Ratio*              | 34% | 42% | 43% | 43% | 39% | 33% |
| Normalised Power Consumption Ratio* | 59% | 24% | 29% | 16% | 27% | 37% |
| Heat source temperature drop (°C)   | 19.5| 24.6| 29.8| 34.9| 40.0| 45.1|
| $PR_{WH}$                           | 1.58| 2.09| 2.54| 2.90| 3.20| 3.45|

*Relative to the optimised conventional MED alternative

Figure 4-6 presents the temperature-energy profiles of the heat source medium of both conventional and Flash Boosted MED systems. For the conventional MED the heat source outlet temperature is around 9°C higher than that of the FB-MED system, although the temperature drop over the first effect is higher. For both systems, the heat source temperature decreases due to the sensible heat transfer. However for the feed, the heat transfer includes both sensible (increasing the feed inlet temperature from 38°C (Table 4-1) to the relevant boiling point temperature) and latent heat transfers (that relates to the
evaporation at the vapour temperature that is slightly lower than the boiling point temperature due to the $BPE$).

Table 4-3 Main flowrates for MED, B-MED and FB-MED over the considered heat source temperature range [3]

<table>
<thead>
<tr>
<th>Type</th>
<th>Flowrates (kg/s)</th>
<th>Heat Source Inlet Temperature (°C)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Condenser Inlet/Outlet</td>
<td>65</td>
</tr>
<tr>
<td>MED</td>
<td>117.1/93.9</td>
<td>165.2/132.6</td>
</tr>
<tr>
<td></td>
<td>Freshwater</td>
<td>8.1</td>
</tr>
<tr>
<td></td>
<td>Brine</td>
<td>15.1</td>
</tr>
<tr>
<td>B-MED</td>
<td>Condenser Inlet/Outlet</td>
<td>160.7/132.4</td>
</tr>
<tr>
<td></td>
<td>Freshwater</td>
<td>9.9</td>
</tr>
<tr>
<td></td>
<td>Brine</td>
<td>18.4</td>
</tr>
<tr>
<td>FB-MED</td>
<td>Condenser Inlet/Outlet</td>
<td>180.0/150.4</td>
</tr>
<tr>
<td></td>
<td>Freshwater</td>
<td>10.8</td>
</tr>
<tr>
<td></td>
<td>Brine</td>
<td>18.8</td>
</tr>
</tbody>
</table>

As shown in Figure 4-6, the FB-MED process has a higher potential to exploit the heat source energy as compared to other processes (MED and B-MED). As it
shows, for 75°C heat source inlet temperatures, FB-MED can exploit 21% and 45% more energy from the heat source as compared to the B-MED and the conventional MED, respectively.

As shown in both Figures 4-1 and 4-3, the first effect's higher heat source outlet temperature of the FB-MED allows it to maintain more effects as compared to the conventional MED process. The liquid-liquid heat exchanger can then exploit more energy from the heat source and hence the outlet heat source temperature of the FB-MED is around 9°C less than conventional MED. This exploited energy is transferred to the flashing chamber feed stream and gradually released over the train of flashing chambers and injected into the proper effects of the primary MED section of the FB-MED system and thereby increasing the freshwater generation [2].

Figure 4-5 Waste heat performance comparison among the three systems and the percentage of increment of boosted and flash boosted MED’s performance ratios compare to the optimised conventional MED for an assortment of heat source temperatures

The trend of the energy transferred over all sections of the systems (namely, MED, B-MED and FB-MED) is shown in Figure 4-7 (dashed line separates the systems). Because of the higher temperature drop over first effect of the conventional MED heat source (Figure 4-6), the amount of energy released in the first effect of the conventional MED is around 192% more than that at the first effect of the FB-MED system at 75°C configuration. In contrast, it is around 18% more than that at the first effect of the B-MED system. The amount of released
energy decreases from effect to effect for conventional MED. The same happens for B-MED, except for the 5th effect, at which steam injection occurs. For FB-MED, this decrement continues until the 3rd effect (i.e. before injection occurs) and, starting from the 4th effect, the energy increases due to steam injection and this trend continues to the end of the process. As shown in Figure 4-7 the total amount of released energy in the primary MED section of the FB-MED system is more than that of the conventional MED system and as mentioned before, this extra energy is realised through the flashing chambers that effectively process more energy from the heat source by using the liquid-liquid heat exchanger (first column as shown in Figure 4-7 for FB-MED). The same happens for B-MED, however the amount of released energy from the booster section is much lower (3.1 times less) than the amount of released energy in the liquid-liquid heat exchanger of the FB-MED process, which results in a lower improvement as compared to the FB-MED system.

Figure 4-8 shows the $UA$ value distribution for all considered systems. For the FB-MED's liquid-liquid heat exchanger, the heat transfer rate is around 9.53 MW (Figure 4-7), coupled with a constant approach temperature of 3°C for both sides of the heat exchanger (Table 1), 3.18 MW/°C $UA$ value is required. Since the major part of heat transfer occurs in both the first effect and in the condenser pertains to sensible-latent heat transfer instead of latent-latent heat transfer as in other effects this results in a higher $UA$ values for those effects compared with that of the first effect and condenser [2].

Referring to section 3.3, the normalized pumping power consumptions of the FB-MED as compared to conventional MED and B-MED are shown in Figure 4-9. It demonstrates that the FB-MED process has the highest auxiliary power consumption, which is mainly related to the brine recirculation pump. The impact of pumping power consumption appears in the operating costs, which will be discussed in the following sections.
Figure 4-6 Temperature profile for the 1st effect of the conventional MED, the booster and the 1st effect of the B-MED and both the 1st effect and the liquid-liquid heat exchanger of the FB-MED system for 75°C heat source inlet temperature configuration.
Chapter 4. Application of the Flash Boosted MED Process in Seawater Desalination

Figure 4-7 Energy released of each effect of all three systems for 75°C heat source inlet temperature configuration

Figure 4-8 UA-value of each effect of all three systems for 75°C heat source inlet temperature configuration

Bijan Rahimi
4.3 Thermo-Economic analysis

Continuing on from the work of Wang, et al. [95] and Rahimi, et al. [2], [3] into B-MED and FB-MED, several simulated conventional MED, B-MED and FB-MED design cases were optimised for production rate and were then analysed using the methods described in Chapter 3. An assumption-based generalised analysis was employed to assess whether the FB-MED processes could be used for desalination installations that had greater economic potential than plants utilising the conventional MED or the B-MED processes. Table 4-4 specifies all the significant assumptions used in conducting the present analyses.

A cash flow analysis was performed upon the annual cash flows of each MED, B-MED and FB-MED plant design, in a manner described in the prior section. Each unique plant design can be characterised by the temperature of heat source fluid, the number of primary MED effects utilised and the number of flashing chambers utilised (FB-MED only).
4.3.1 Capital costs

The capital cost of FB-MED plant designs increases with increased heat source fluid temperature and with the addition of each flashing chamber, which gives rise to: 

All relevant thermo-economic simulation results have been summarised in Table 4-5 [3].

Table 4-5 Thermo-economic simulation results for MED, B-MED and FB-MED over the considered heat source temperature range [3]

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Process</th>
<th>Heat Source Inlet Temperature (°C)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>65</td>
</tr>
<tr>
<td>Total Capital Cost (US$)</td>
<td>MED</td>
<td>1,833,501</td>
</tr>
<tr>
<td></td>
<td>B-MED</td>
<td>2,216,058</td>
</tr>
<tr>
<td></td>
<td>FB-MED</td>
<td>2,300,687</td>
</tr>
<tr>
<td>Annual Electrical Expenses (US$/yr)</td>
<td>MED</td>
<td>68,945</td>
</tr>
<tr>
<td></td>
<td>B-MED</td>
<td>99,789</td>
</tr>
<tr>
<td></td>
<td>FB-MED</td>
<td>146,072</td>
</tr>
<tr>
<td>Annual Labour Cost (US$/yr)</td>
<td>MED</td>
<td>93,010</td>
</tr>
<tr>
<td></td>
<td>B-MED</td>
<td>93,010</td>
</tr>
<tr>
<td></td>
<td>FB-MED</td>
<td>93,010</td>
</tr>
<tr>
<td>Annual Chemicals Cost (US$/yr)</td>
<td>MED</td>
<td>15,182</td>
</tr>
<tr>
<td></td>
<td>B-MED</td>
<td>18,578</td>
</tr>
<tr>
<td></td>
<td>FB-MED</td>
<td>19,092</td>
</tr>
<tr>
<td>Annual Maintenance, Spares and Insurance (US$/yr)</td>
<td>MED</td>
<td>27,503</td>
</tr>
<tr>
<td></td>
<td>B-MED</td>
<td>33,241</td>
</tr>
<tr>
<td></td>
<td>FB-MED</td>
<td>34,510</td>
</tr>
<tr>
<td>Total Annual Operating Costs (US$/yr)</td>
<td>MED</td>
<td>204,640</td>
</tr>
<tr>
<td></td>
<td>B-MED</td>
<td>244,618</td>
</tr>
<tr>
<td></td>
<td>FB-MED</td>
<td>292,684</td>
</tr>
</tbody>
</table>
to an associated increase in production capacity. Though the capital costs of an optimised FB-MED and B-MED plant are predominantly greater than the MED alternative, the normalised capital cost is also lower, as seen in Figure 4-10. The normalised capital cost can be as much as 7% lesser for FB-MED, and as much as 1% lesser for B-MED, as compared to the MED alternative [3]. Inherently this reflects the improvements in the thermal efficiency of the process. Less investment is required per unit product for the B-MED and FB-MED processes due to the greater thermal efficiency in the utilised desalination process.

![Figure 4-10](image)

**Figure 4-10** A comparison of the normalised capital costs of optimised MED, B-MED and FB-MED plants for each considered heat source temperature. The values are indicative of the percentage decrease in the specific capital cost of the B-MED or FB-MED, as compared to the conventional MED alternative for each specific heat source temperature. [3]

### 4.3.2 Operating costs

Results show that the annual operating cost increases with heat source fluid temperature and is greater for both B-MED and FB-MED than the operating cost of the MED alternative, as can be seen in Figure 4-11 [3]. The general increase in operating cost can be attributed to several factors (such as auxiliary pumping power), all of which are associated with the increased production rates of the FB-MED and B-MED. It can be seen also that the component operating expenses increase for the B-MED and FB-MED plants for each heat source temperature (except for labour cost which has a fixed value as mentioned before). As Figure 4-11 shows, the electrical expenses constitute the greatest increase in operating
expenditure for FB-MED system. The FB-MED system has several more pumps to handle flows which enable the increased thermal efficiency of the system, which can be seen by comparing Figures 4-1 and 4-3 [3]. The make-up water, brine recirculation and drain pumps can be seen in Figure 4-3, the operation of each constituting an extra duty requirement to the FB-MED system. Furthermore the pumps that exist in both MED and FB-MED plants have a greater duty in the FB-MED system. The heat source pump has a greater duty requirement because of the greater pressure difference due to the extra heat exchanger (Figure 4-3) [2]. Due to the increased production of the plant, the brine, freshwater, NCG (Non Condensable Gas) extraction and saline water pumps also have increased duties, and therefore greater electrical consumptions [3]. In general, according to Figure 4-11 it can be seen that the annual electrical cost (which is mostly related to the total pumping power consumption, Table 4-2) for MED, B-MED and FB-MED is varying between 33 to 50% of the relevant total annual operating cost (Table 4-5) [3]. This means that pumping power only exerts an effect on a portion of the total operating costs which is no more than half of it [3].

Chemical expenses increase for the FB-MED is generally due to the increased production of the system compared to the MED, but also due to the addition of the flashing chamber feed and its chemical dosing requirements. Expenses related to insurance, maintenance and spares also increase due to the increased capital cost of the FB-MED plant. The component operating costs are also greater for B-MED, and for much the same reasoning [3].

Despite the overall increase in operating costs, the specific operating costs for B-MED and FB-MED processes are comparable to that of the MED process, as can be seen in Figure 4-12 [3]. This is due to the higher production rates of B-MED and FB-MED as compared to the conventional MED. The figures for the B-MED process are systematically smaller, as it requires a smaller increase in electrical consumption than the FB-MED alternative. Likewise the more moderate gains in production mean that chemicals and maintenance, spares and insurance expenses increase less dramatically than the FB-MED alternative.
4.3.3 Unit product cost

Figure 4-13 details the decrement in $UPC$ for B-MED and FB-MED compared to MED. Generally, B-MED and FB-MED have an appreciably lower $UPC$ at lower heat source fluid temperatures, but this diminishes as the temperature increases. Both novel processes show a marked improvement in $UPC$ over conventional MED, owing to the improved efficiencies of the B-MED and FB-MED processes.
4.3.4 Net present value and incremental analysis

The \( NPV \) of the alternatives is plotted for a range of assumed interest rates (Figure 4-14). The \( \Delta IRR \) at the intersection with the abscissa is also calculated and shown only for FB-MED versus B-MED in Figure 4-15. The investment alternative with the higher \( NPV \) for a given range of interest rates is the superior option for those assumed conditions. If no option has a \( NPV \) greater than zero for a specific interest rate, then it is preferable to commit to neither option. Clearly the FB-MED has a greater \( NPV \) over the range of interest rates, making it the economically superior option to both B-MED and MED; however B-MED is also a better option to conventional MED in all cases. For example, for the FB-MED plants utilising a 75°C inlet heat source, the \( IRR \) is 20.3% [3]. This implies that at an interest rate of 20.3%, the \( NPV \) is equal to zero, and that beyond this threshold interest rate value it is preferable to do nothing. For interest rates below this threshold value, the FB-MED plant is the most valuable choice for the assumed conditions. In this case, the \( IRR \) of the conventional MED and B-MED are 19.0 and 19.2%; respectively, both less than the \( IRR \) of FB-MED. As shown in Figure 15 for the entire heat source temperature range the \( \Delta IRR \) between FB-MED and B-MED is more than 8% (the assumed \( MARR \) [3]), thus as mentioned in Chapter 3, it means that the FB-MED process as the higher cost alternative is the better choice.
Generally beyond the first one or two flashing chambers, the net present value of FB-MED plants increases with the addition of each flashing chamber, as can be seen in Figure 4-16. Such results are expected considering both the decreased normalised capital and operational costs associated with the FB-MED process that improve with the addition of each flashing chamber. The results indicate primarily that the potential for monetary savings associated with the improved thermal efficiency of the process is not outweighed by the necessary increase in investment [3]. As such, the FB-MED designs that have the highest production rate, and that are inherently the most efficient in production, are also the designs that possess the greatest economic value.

Figure 4-17 compares the optimised MED, B-MED and FB-MED plant designs and indicates the incremental improvement in $NPV$ over the MED, according to the available market price for water, that is 2.50 US$/m^3$ (equivalent of 2.72 AU$/m^3$ [156]). As it shows, the FB-MED process has a considerably greater $NPV$, being up to 53% greater than that of the MED alternative [3]. Similarly, the
NPV of B-MED installations can be up to 30% greater than that of conventional MED [3]. We also note that UPC, in general, cannot clarify the profitability of two systems [3]. Referring to both Figures 4-13 and 4-17, at 70°C heat source inlet temperature both B-MED and FB-MED's unit product costs are the same yet the NPV of FB-MED is 18% more than that of the B-MED process [3].

Figure 4-15 Net Present Value (NPV) vs. interest rate plots for the difference between FB-MED and B-MED for all the relevant heat source temperatures. It also highlights the incremental internal rate of return, ∆IRR. [3]

Figure 4-16 NPV vs. number of installed flashing chambers for all considered temperatures of the heat source fluid. Note that the plant designs utilising zero flashing chambers are conventional MED plants, whilst all remaining plant designs are FB-MED. The percentage values above some data points indicate the percentage increase in NPV for that FB-MED design compared with the conventional MED alternative (0 number of flashing chamber) for that particular heat source temperature. [3]
Chapter 4. Application of the Flash Boosted MED Process in Seawater Desalination

4.4 Conclusions

A novel FB-MED desalination process has been conceptualised and simulated, and shown to achieve higher production rates and performance ratios than the conventional MED and B-MED processes over an extensive range of low grade sensible heat source temperatures. A generalised method for estimating the costs of conventional MED, B-MED and FB-MED plants has been developed and numerous configurations have been analysed accordingly. Measures of profitability have been obtained for each design so that the economic viability of FB-MED designs can be assessed in terms of the number of installed flashing chambers. The results of the analysis indicate that the increased production capacity and thermal efficiency of the B-MED and FB-MED processes outperform the increased capital and operational requirements inherent to these processes. Results of the analysis indicate that the optimised B-MED and FB-MED systems can have a net present value (\(NPV\)) that is up to 30% and 53% respectively greater than the conventional MED, depending on the temperature of the utilised heat source [3]. We have also shown that \(NPV\) is a better measure of profitability than Unit Product Cost (\(UPC\)), as \(NPV\) is able to differentiate two propositions even when \(UPC\) for those two propositions is the same. In general it can be postulated therefore, that the novel FB-MED process is both technically and
economically superior to conventional MED and B-MED processes when they are coupled with low grade sensible heat sources, and are therefore viable future desalination technologies [3].
Evaporation is an essential process in alumina refineries both to produce water for such applications as residue washing processes, and re-concentrating process liquor back to the main process circuit for further digestion. This chapter reports on the application of a novel low grade heat driven re-concentration process. It consists of both flashing and falling film techniques, and is meant to couple to one of the available waste streams in the evaporation unit as a heat source and thereby re-concentrating the process liquor. The heat source has been identified as the process condensate stream. The condenser of the re-concentration process is cooled by barometric water from the recovery lake. The generated freshwater is either discharged or sent to the freshwater lake so as to manage the water balance of a refinery. For 200 m$^3$/hr process condensate simulated at two fixed temperatures of 75°C and 85°C, the novel process generates up to 36% more re-concentrated process liquor and about 10% less specific capital cost compared to the conventional multi effect falling film evaporation system.

**Highlights:**

- The novel FB-MEE distillation process has been introduced for alumina refinery application.
- It has been benchmarked against the optimised conventional MEE process.
- This process can improve the freshwater and reconcentrate production rates by 36%.
- Its thermal performance increases by 36%.
- Its specific capital cost reduces by 12%.

**Publication:**

This chapter is identical to my conference paper in 10th International Alumina Quality (AQW) Workshop, Perth, Australia, 19th-23rd April 2015, 327-336 [98] and my published paper in Hydrometallurgy journal [159] as below details:

5.1 Introduction

The evaporation process is one of the key parts of an alumina refinery that also consumes huge amount of thermal energy. It is responsible for the water balance in the whole plant, providing feed stream for digestion and freshwater which may be utilised as washing water in other sections. Its role can therefore be summarised as to balance water and discharge impurities. Evaporation unit is typically run by available low pressure steam and is one of the energy intensive sections in an alumina refinery plant [160]. Therefore any improvement on this section can decrease the steam consumption, which in turn saves the fuel consumption in the steam plant which then reduces greenhouse gas emission.

The design of evaporator is purely based on its suitability to handle process liquor which is viscous and prone to scaling. Therefore, both tube falling film and flashing vessel techniques (evaporators) can be used. Recently the plate falling film evaporator is also used by Aluminum Company of Shanxi and Guizhou Alumina [161], [162] to improve the evaporation efficiency in evaporation sections with low concentration feed streams. It is worthwhile to mention that a state-of-the-art pilot plant consisting of two serially connected Alfa Laval's single-effect rising film plate evaporator/condenser modules [110] is also being tested in Worsley Alumina Refinery Plant to study and address the potential scaling issues when it is coupled to different process streams (Figure 5-1).

Figure 5-1 The two serially connected Alfa Laval's single-effect rising film plate evaporators modules (JWP-16-C Series) [110]
Both techniques, namely evaporation and flashing, have been shown in Figures (5-2 and 5-3). Independent of the evaporator types, a Multi Effect Evaporation (MEE) process is the preferred choice to reduce energy consumption. In the falling film technique, the feed liquor enters the evaporator from the top, which is specially designed to distribute the feed into the tubes. Gravity pulls the film down along the inner walls of the heater tubes, so that the fast moving and thinning film gives rise to a high heat transfer coefficient [163]. At the bottom of the heat exchanger the feed stream is re-concentrated. The outgoing vapour from the first effect can be the heat source of the second effect, and this trend continues through to the last condenser.

Alternatively, a combination of heat exchangers and flash vessels known as Multi Stage Flash (MSF) can also be used (Figure 5-3). In this system, the feed liquor is firstly preheated by going through a series of heat exchangers (preheaters) and is then introduced to a battery of flash vessels. The flashed vapours from the vessels are condensed in the preheaters, and the highly concentrated feed from the last outlet is the final product of the evaporation unit. The first heat exchanger in this system uses the available live steam as the heat source. The first heat exchanger uses the low pressure steam as the heat source.

The aim of this chapter is to introduce a novel process [2]–[4], [111] that can be coupled with waste heat sources in an alumina refinery plant to reduce the steam consumption of the evaporation unit. This reduces fuel consumption, greenhouse gas emission, and production cost [98]. It should be clarified that the novelty herein lies with the process design instead of the process units themselves, the latter are simply standard technologies. The fact that standard process units are used augurs well for scale-up adoption in refineries. For illustration purpose, we elect to focus on one available waste heat stream that comes out from an evaporation unit of Worsley Alumina Refinery Plant. This stream is the outlet process condensate stream of the main evaporation unit (Figure 5-3, MSF Process). This novel technique can potentially harvest all available sensible waste streams to run an auxiliary evaporation unit so as to reduce the load of the main evaporation unit which in turn saves the steam consumption and therefore coal consumption of Worsley Alumina Refinery Plant [98].
Rahimi, et al. [2]–[4], [111] developed a process that can be coupled with low grade waste sensible heat streams to boost freshwater production as compared to the conventional MEE process. Figures (5-4) shows the Flash Boosted MEE (FB-MEE) process when it is coupled with the outlet process condensate of the evaporation unit [98]. The available temperature is around 85°C. The simulation has also been done for 75°C.

As shown in Figure 5-4, the process condensate (inlet sensible heat source) is firstly used to power the primary MEE section, and then to heat the feed stream which supplies the flash vessels. Each vessel produces flashed vapour which is directed to an appropriate primary MEE effect according to the relevant pressure differences, thereby supplying more heat and increasing the amount of feed which can be evaporated across the MEE effect. The result of this vapour injection scheme is an increase in freshwater production and re-concentrate flow rate across the primary MEE effects. The number of falling film evaporation effects can be set based on the available temperature difference between the
heat source and the cooling water temperature and the relevant boundary conditions \[2\], \[95\].

Figure 5-4 Schematic design of a Flash Boosted Multi Effect Evaporation (FB-MEE) unit (parallel feed) (Three falling film effects and two flash vessels) coupled with an 85°C low grade heat source \[98\]

5.2 Numerical Analysis and Validation

A simulation model has been developed to quantify the efficiency of the flash boosted MEE system \[2\], \[98\]. The mathematical model includes mass, concentration and energy balance for each effects. Mass and concentration balances are written as \( k \in \{1, \ldots, n\} \):

\[
m_{F,k} = m_{V,k} + m_{HC,k}
\]

\[
m_{F,k} \cdot X_F = m_{HC,k} \cdot X_H
\]

Energy balance in the first effect can be written as below.

\[
\dot{Q}_1 = \dot{m}_{F,1} \cdot h_{F,1} + \dot{m}_{HS} \cdot h_{HS,\text{in}} - \dot{m}_{HS} \cdot h_{HS,\text{out}} + \dot{m}_{V,1} \cdot h_{g(T_{HC,1-P_{sat,1}}} + \dot{m}_{HC,1} \cdot h_{HC,1}
\]

\( \dot{Q} \) is the total heat transfer (kW), \( \dot{m}_{F} \), \( \dot{m}_{HS} \), \( \dot{m}_{V} \), and \( \dot{m}_{HC} \) are the feed, heat source, generated vapour and re-concentrated feed mass flow rates (kg/s), respectively. \( h_f \) and \( h_g \) are the enthalpies of saturated liquid and steam (kJ/kg), respectively.

Figures 5-5 and 5-6 illustrate the temperature-energy profile across the first effect through to the last effect and the condenser of both conventional MEE and FB-MEE processes for 85°C heat source inlet temperature. The process conditions considered conform to the assumptions listed in Table 5-1. In the first effect, the heat source temperature drops across the evaporator, while the feed liquor...
temperature firstly increases from the inlet temperature to the relevant boiling temperature ($T_{HC}$) (preheating zone), according to its concentration at the prevailing pressure ($P_{sat}$) and then increases further during the evaporation process because of continuing concentration that brings about heightened Boiling Point Elevation ($BPE$) (evaporation zone). Therefore due to the different types of heat transfer between preheating and evaporation zones, there are two different $UA$ values. In this case, the preheating zone covers less than 4% (Figures 5-5 and 5-6) of the total energy transferred and therefore to a good approximation it can be ignored. Hence the overall $U$ (heat transfer coefficient, (kW/m$^2$.K)) value for the first effect pertains to that of the evaporation zone.

For the second effect, $k \in \{2, \ldots, n\}$, where $n$ is the total number of effects, we have

$$Q_k = \dot{m}_{V,k-1} \cdot h_{g(T_{HC,k-1},p_{sat,k-1})} + \dot{m}_{F,k} \cdot h_{f,F,k} = \dot{m}_{V,k-1} \cdot h_{g(T_{HC,k-1},p_{sat,k-1})} + \dot{m}_{V,k} \cdot h_{g(T_{HC,k},p_{sat,k})} + \dot{m}_{HC,k} \cdot h_{f,HC,k}$$

(5-4)

$h_{f,D}$ is the enthalpy of distillate at its saturation temperature (kJ/kg). Figures 5-5 and 5-6 show the temperature-energy profile across the second and third effect. Because the heat source of these effects is the superheated vapour from the previous effect, the superheated vapour de-superheats to the relevant saturated temperature of the previous effect, and then it condenses. In contrast, the trend of feed temperature profile is the same as in the first effect if the saturation pressure of the feed stream is less than the effect’s pressure, as with the second effect of the FB-MEE process as shown in Figure 5-6. Therefore, in this case, the feed temperature increases to the boiling point. The temperature then increases further during the evaporation process. In this case, to an excellent approximation, the overall $U$ value for these effects is simply that of the evaporation zone.

Sometimes the feed saturation pressure is more than the effect’s pressure, as in the second effect of the conventional MEE and the third effect of the FB-MEE process. In this case, before the feed is being distributed into the tubes, a negligible amount of feed (less than 1%, which corresponds to about 7°C of sensible heat in our case) is flashed as the pressure and temperature decrease.
Chapter 5. Application of the Flash Boosted MEE Process in Alumina Refinery Plant

For these effects the inlet feed temperature is considered to equate to the relevant feed boiling temperature (which is related to its concentration) that corresponds to the pressure in the second effect and, in Figures 5-5 and 5-6, is equal to 42.6°C. After temperature reduction, once it distributes into the tubes, evaporation occurs. As shown in Figures 5-5 and 5-6, for these effects the amount of energy required for de-superheating is negligible.

For the second and third effects of both processes, the area needed for de-superheating has a significant effect on the capital cost of the evaporator. This will be discussed in detail in section 5.3. In this case because the degree of superheating is less than 10°C, de-superheaters cannot practically be used, and additional heat exchanger area has to be factored in [164].

In the FB-MEE process the impact of the injected vapour from the flash vessels that mixes with the superheated vapour from the relevant upstream MEE effect, has to be considered in arriving at the mixed superheated vapour temperature.

For the condenser,

$$\dot{Q}_{\text{cond}} = m_{v,n} \cdot h_{g(T_{HC,n}, P_{sat,n})} + m_{\text{cond}} \cdot h_{f_{\text{cond,in}}} = m_{v,n} \cdot h_{f_{D(T_{sat,n})}} + m_{\text{cond}} \cdot h_{f_{\text{cond,out}}}$$  \hspace{1cm} (5-5)

$m_{\text{cond}}$ is the cooling water mass flow rate (kg/s). As again Figures 5-5 and 5-6 show the cooling water temperature increases across the condenser but the trend of the inlet superheated vapour temperature profile is the same as in the other effects, so that the de-superheating zone is negligible once again. The overall heat transfer coefficient is practically that of the latent-to-sensible heat transfer zone. De-superheating requires extra heat exchanger area.

In all the above equations,

$$T_{\text{sat}} = T_{HC} - BPE$$  \hspace{1cm} (5-6)

$BPE$ is the Boiling Point Elevation at the outlet of the re-concentrated process liquor in each effect, and is based on onsite specifications. $T_{\text{sat}}$ and $T_{HC}$ are the saturation temperature and the boiling temperature of the re-concentrate (°C). $BPE$ generally depends on the temperature and concentration of the re-concentrated solution. In this context, on account of the large internal vapour volume within the falling film tube evaporator and the fact that the bulk of the
evaporation happens at the latter part of the evaporator where it enjoys a more substantial temperature difference between the sensible heat source and the feed, we consider a pragmatic situation where the internal tube temperature throughout the entire tube, that essentially represents that of the evaporated vapour, corresponds to the outlet re-concentrate temperature. Hence, in this simulation as a conservative and pragmatic assumption, the temperature difference between the outlet re-concentrate boiling temperature and the heat source temperature at the pinch point for the first effect in both processes is assumed to be 3°C. For the other effects this 3°C temperature difference is assumed between the outlet re-concentrate boiling temperature and the heat source condensation temperature. An 8°C boiling point elevation has been considered for the re-concentrate which is equivalent to the BPE of 20wt% NaOH [1], [119], [121]–[123]. The BPE also dictates the degree of superheating, which in this case is 8°C and remains less than 10°C which is the threshold for deploying de-superheaters [164].

### Table 5-1 Assumptions of thermodynamic simulations [98]

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Heat source flow rate, $m_{HS,1}$ (m$^3$/hr)</td>
<td>200</td>
</tr>
<tr>
<td>Heat source temperature range (°C)</td>
<td>75, 85</td>
</tr>
<tr>
<td>Feed to vapour ratio (50% evaporation)</td>
<td>2.0</td>
</tr>
<tr>
<td>$T_{C,in}$ (°C) / $T_{C,out}$ (°C)</td>
<td>30 / 38</td>
</tr>
<tr>
<td>$\Delta P_{inj}$ (Pa)</td>
<td>1000</td>
</tr>
<tr>
<td>Liquid-to-Liquid heat exchanger approach temperature (°C)</td>
<td>3.0</td>
</tr>
</tbody>
</table>

# relevant pressure difference for steam injection purpose in FB-MEE processes

As shown in Table 5-1, for the purpose of simulation, the heat source medium is the process condensate from the evaporation plant at 75 or 85°C and its flow rate is 200m$^3$/hr. The inlet/outlet coolant temperatures at the condenser are 30°C/38°C respectively and it is supplied from a recovery lake close to the site. For the liquid-to-liquid heat exchanger in the FB-MEE process the minimum temperature of approach is taken to be 3°C. The inlet feed is 10wt% process liquor (1.6°C BPE [1], [119], [121]–[123]) which is re-concentrated to 20wt% (8.0°C BPE [1], [119], [121]–[123]). The feed to vapour ratio ($R = \dot{m}_F / \dot{m}_V$) for each MEE effects is taken to be 2.0.

For the flash vessel section (that is applicable only to the FB-MEE process), the energy balance equation is expressed as Equation (5-7) [98], $i \in \{1, \ldots, j\}$, where $i$ is the flash vessel number and $j$ the number of flash vessels.
\[ \dot{m}_{F,i,in} \cdot h_{F,i,in} = \dot{m}_{V,i} \cdot h_{g(T_F,out,i,P_{sat,i})} + (\dot{m}_{F,i,in} - \dot{m}_{V,i}) \cdot h_{F,i,out} \] (5-7)

\( \dot{m}_{F,i,in} \) and \( P_{sat,i} \) are the inlet feed mass flow rate (kg/s) and the saturated pressure (kPa) of the \( i^{th} \) flash vessel and

\[ T_{Sat} = T_{F,out} - (BPE + NEA) \] (5-8)

\( T_{F,out} \) is the temperature of the outlet feed of the flash vessel. \( NEA \) (Non-Equilibrium Allowance) for flash vessels is a function of flash temperature range, saturation temperature, mass flow rate of high concentrate outlet per unit of vessel width, its level inside the flash vessel, and the design of the flash vessel such as vessel length, width and the orifice type as the transfer device from vessel-to-vessel [2]. For this application, a fixed 0.5°C \( NEA \) is considered. Therefore in each flash vessel a fixed 8.5°C (namely \( BPE + NEA \)) is present [98]. The injection from each flash vessel to the relevant MEE effect is being considered with 1 kPa saturation pressure difference between the vessels (Table 5-1) [98].
Chapter 5. Application of the Flash Boosted MEE Process in Alumina Refinery Plant

Bijan Rahimi

Figure 5-6 Schematic of the temperature-energy profiles of the FB-MEE process for 85°C heat source inlet temperature

The total amount of freshwater generated is the sum total of the condensate from each of the MEE effects (Equation 5-9).

\[ M_{D,totai} = \sum_{k=1}^{n} m_{V,k} + \sum_{i=1}^{j} m_{V,i} \]  

(5-9)

The Generalised Reduced Gradient (GRG) method, which imposes boundary conditions from the thermodynamic laws and operational, technical and economic constraints, is used to solve the above equation set [98]. The abovementioned method has been validated before in [2]–[4], [95].

5.3 Results and Discussions

5.3.1 Production rate and waste heat performance ratio

Figures 5-4 and 5-7 show the schematics of both the conventional MEE and FB-MEE coupled with an 85°C low grade heat source [98]. They include all the important process temperatures in the simulation. The higher temperature drop for the heat source in the case of the FB-MEE process as compared to the conventional MEE process, means that more waste heat has been effectively
utilised to improve the production rate. Figure 5-8 [98] benchmarks the present FB-MEE scheme with the standard conventional MEE process over a range of heat source inlet temperatures. At 85°C heat source inlet temperature, both the freshwater yield and re-concentrated process liquor yield are increased by 36% with the FB-MEE process [98].

A waste–heat performance ratio [96] \( PR_{WH} \) - Equation (5-10) - which measures the efficacy of waste heat usage is also used to benchmark the performances. The specific enthalpy of evaporation of 2336 kJ/kg is a typical benchmark to translate the waste heat utilised to an equivalent amount of steam used [2]. The denominator represents the maximum exploitable energy of the heat source relative to the lowest available temperature which in this case is the condenser inlet temperature.
\( PR_{WH} = \frac{2336 \cdot m_{D$total}}{m_{HS}(h_{f,HS,in}-h_{f,cond,in})} \) (5-10)

Figure 5-8 also shows a 34% and 36% increment in the waste heat performance ratio of the FB-MEE process compared to the conventional MEE [98], consistent with the production rate improvement.

### 5.3.2 Auxiliary pumping power

The major rate of electrical consumption is related to the pumping power consumption. The pumping power is calculated according to the equation below.

\[ \text{Pumping Power (kW)} = \left( \frac{\Delta P \cdot \dot{V}}{\eta} \right) \] (5-11)

\( \Delta P \), \( \dot{V} \) and \( \eta \) are the total head (kPa), volumetric flow rate (m\(^3\)/s) and the efficiency of the pump, respectively. As shown in Figures 5-4 and 5-7, the main pumps of this process are summarised below.

#### 5.3.2.1 Cooling water pump

This pump conveys the cooling water through the condenser. The required head for this pump is taken to be 1.5 bar [2]. A 75% efficiency has been considered for this pump based on site specification [98].

#### 5.3.2.2 Re-circulation pump

This pump is used in the FB-MEE system to recirculate the re-concentrate through the liquid-liquid heat exchanger and the flash vessels. The head of this pump is a function of pressure drop across the heat exchanger, spray pipe, control valves, pipelines, geodetical and pressure heads (that is calculated based on the inlet and outlet saturated pressure difference across the flash vessels) [2], [62]. Therefore the required head for all the above mentioned heads save the pressure head has been considered to be 1.7 bar [2], and that the total head is calculated by adding the pressure head specific to the process. According to the site specifications, an 80% efficiency has been considered for this pump [98].

#### 5.3.2.3 Condensate extraction pump

This pump is used to extract the condensate from the primary MEE unit. The required head for this pump is taken to be 2 bar [2]. The efficiency is considered to be 70% [98].
5.3.2.4 Heat source pump

This pump is used for both conventional MEE and FB-MEE processes, and pumps the main heat source medium through the evaporator (first effect) of the primary MEE plant and the liquid-liquid heat exchanger of the FB-MEE process. The required head for this pump is taken to be 0.5 bar for the first effect (evaporator) and 0.5 bar for the liquid-liquid heat exchanger in the FB-MEE process [2], with a 70% efficiency [98].

5.3.2.5 Re-concentrate extraction pump (Drain pump)

There are two extraction pumps for the re-concentrated liquor - one to extract from the primary MEE and the other from the flash vessel unit. The required head for this pump is 2 bar [2]. Based on site specification an 80% efficiency has been considered [98].

5.3.2.6 Feed pump

The feed pump pumps the feed stream to the evaporation unit. A 1 bar head and an 80% efficiency have been considered for this pump [98].

5.3.2.7 NCG extraction vacuum pump

In order to remove the generated non-condensable gasses (NCG) and the air leakage into the system, a water ring vacuum pump has been considered to remove 1% of the vapour in the condenser as NCG's [135]. The corresponding pumping power consumption is evaluated according to the manufacturer's catalogs for water-ring vacuum pumps [2].

Figure 5-9 shows the breakdown of the specific pumping power for both systems to highlight the share of each pump in the total specific pumping power [98]. The FB-MEE consumes 53% and 24% more than that of the conventional MEE at 75 and 85°C, respectively. The main difference between the two processes is related to the re-circulation pumps deployed in the FB-MEE process. The substantial drop from 53% at 75°C to 24% at 85°C is because of the extra primary MEE effect that can be installed into the FB-MEE process on account of the higher heat source inlet temperature as compared to the conventional MEE [98].
5.3.3 Capital cost analysis

Capital costs depend on the material and equipment specifications. For the purpose of comparison, the focus herein is on the relative capital cost increase of the FB-MEE compared to the conventional MEE process and its impact on the specific capital costs. The latter is defined as the ratio of the gross capital cost to the total freshwater production rate, or equivalently to the re-concentrate production rate since the feed-to-vapour ratio is 2, which means that there is an equal production of freshwater and re-concentrated process liquor. For conventional thermal desalination plants (both MEE and MSF processes), the capital cost can be approximated based on the GWI/IDA database (IDA, 2011). For the capacity range up to 10,000 m³/day, this results in the following cost function [3].

\[
Cap_{\text{Cost}}^{\text{MEE,MSF}}(US\$) = \psi_{D_v} = 3054 \times D_v^{0.9751}
\]  

(5-12)

where \(D_v\) is the total production rate of the plant in m³/day and \(\psi_{D_v}\) is the cost function. The capital cost increase is due to the array of flash vessels as in the FB-MEE process. The flash vessels are major parts of a typical MSF process; therefore the cost can be estimated by the cost equation by treating the total vapour produced from the flash chambers as the equivalent MSF plant capacity. However, in contrast to MSF plants, no condenser tubing and no heat rejection
section is required for the array of flash vessels, which are also major cost items in thermal desalination plants [2], [165].

The capital costs for MSF and MEE systems are subdivided into five main sectors [165], [166]: Capital cost of the evaporator (40% of the total capital cost), equipment piping (29%), erection (14%), engineering and commissioning (10%), and electrical, instrumentation and control (7%); therefore the total capital cost for the FB-MEE process can be estimated as follow [2], [3]

\[
\text{Capital Cost}_{\text{FB-MEE}} = \psi_{D_{t,\text{FB-MEE}}} - D_{FV} + [0.20 \cdot (\psi_{D_{t,\text{FB-MEE}}} - \psi_{D_{t,\text{FB-MEE}}} - D_{FV})] + 0.41 \cdot \psi_{D_{FV}} \quad (5-13)
\]

\(D_{FV}\) is the production rate (vapour) of all flash vessels and \(D_{t,\text{FB-MEE}}\) is the total production rate of the FB-MEE plant. As mentioned in section 5.3, save the first effect, all vapour introduced to the evaporation effects have around 8°C superheat. This means that extra heat exchanger area for de-superheating in those evaporation effects has to be considered, and will give rise to a higher capital cost that impact on the evaporator section which is 40% of the total plant capital cost [165], [166]. From practical experience [164] 1% extra heating area is needed per 1°C of superheat to account for future scaling problem due to the presence of superheat steam. For this purpose, an area coefficient (\(\phi\)) is applied to address the extra heat exchanger area of the evaporation effects (except the first effect) as shown in Table 5-2.

<table>
<thead>
<tr>
<th></th>
<th>First Effect ((\phi_1))</th>
<th>Second Effect ((\phi_2))</th>
<th>Third Effect ((\phi_3))</th>
<th>Condenser ((\phi_{cond}))</th>
</tr>
</thead>
<tbody>
<tr>
<td>MEE</td>
<td>1</td>
<td>1.08</td>
<td>N/A</td>
<td>1.08</td>
</tr>
<tr>
<td>FB-MEE</td>
<td>1</td>
<td>1.08</td>
<td>1.09</td>
<td>1.09</td>
</tr>
</tbody>
</table>

\[
A_{\text{calc.}} = A_1 + A_2 + A_3 + \cdots + A_n + A_{\text{cond.}}. \quad (5-14)
\]

\[
A_{\text{actual}} = \phi_1 \cdot A_1 + \phi_2 \cdot A_2 + \cdots + \phi_n \cdot A_n + \phi_{\text{cond.}} \cdot A_{\text{cond.}}. \quad (5-15)
\]

\[
Y = \frac{A_{\text{actual}}}{A_{\text{calc.}}} \quad (5-16)
\]

\(A_{\text{calc.}}\) and \(A_{\text{actual}}\) (in m²) are the total heat exchanger area of the primary MEE section without and with considering the superheated steam impact, respectively.
and \((\alpha)\) is the area coefficient that should be applied to the evaporator capital cost (40% of the total capital cost), and so the total capital cost equation is:

\[
Capital\ Cost\ (US\$) = (0.4 \cdot Y + 0.6) \cdot \psi_{Dt}
\]

This correlation should be applied to the three \(\psi_{Dt}\) terms in Equation (5-13) except for \(\psi_{DFV}\), which relates to flash vessels.

Following this correlation, the FB-MEE scheme incurs around 20% higher total capital costs as compared to a conventional MEE unit (Figure 5-10). However, the specific capital cost ratio (Equation 5-18) is actually reduced by around 12% over the application range due to the increased production rate:

\[
Specific\ Capital\ Cost\ Ratio = \frac{Capital\ Cost\ Ratio}{Production\ Rate\ Ratio}
\]

5.4 Conclusion

A novel flash boosted MEE process has been developed for the exploitation of waste heat streams in an alumina refinery plant to minimise the steam consumption of evaporation units and reduce overall fuel consumption. Its performance has been benchmarked with a conventional MEE system in terms of freshwater and re-concentrated process liquor production and waste–heat performance ratio. The flash boosted MEE scheme has been demonstrated to be around 36% better than the optimised conventional MEE process with respect to the production rate over the range of waste heat stream temperatures considered. The specific electrical power consumption of this system is 53% and 24% for the inlet heat source temperatures of 75 and 85°C, respectively. The
main reason for the reduction in the pumping power consumption for the 85°C case is because of the extra effect installed in the primary MEE section. A capital cost analysis indicates 10 to 12% decrement for the specific capital cost of this system over the conventional system for the same production rate.
In this chapter, we have developed a novel flash boosted evaporation process that utilises the available waste heat sources of the evaporation units to reduce live steam consumption for evaporation purpose in alumina refineries. The results show a significant reduction in the live steam consumption of the evaporation plant in one of the major alumina refineries in Western Australia. This system can be attached to the main evaporation process without any process interruption. In this case, the available waste stream is utilised to recover the duty of one of the selected evaporation units, and it shows that the selected unit can be completely turned off, resulting in 82% of its relevant live steam consumption being saved. A detailed capital cost analysis and pumping power consumption evaluation is presented as well. This technology is also applicable to mineral refineries in general that involves re-concentrating liquor and water balance management.

**Highlights:**

- A novel TVC-MEE distillation process has been introduced for refinery applications.
- This process can save 82% of the selected evaporation unit’s available live steam.
- It has been benchmarked against the conventional TVC-MEE process.
- Its thermal performance increases by 98% against conventional process.
- Its specific capital cost reduces by 11% compared with conventional process.

**Publication:**

This Chapter is identical to my published article in Applied Thermal Engineering journal [167] as referenced below detail.

6.1. Introduction

Alumina itself is the precursor to aluminum which is 8% of the Earth by mass [168] typically found in the form of bauxite. Aluminum is the most abundant metal and the second most produced metal in the world, after steel [169], [170]. Bauxite is the essential raw material to produce alumina. It contains approximately 30% alumina in the form of aluminum hydroxide and Australia is the world leading producer of it; in 2012 Australia produced 79.4 Mt of bauxite from five bauxite mines, which is 65% more than China - the second largest bauxite producer [171].

Alumina is produced via the Bayer process which was invented and patented in 1887 by Karl Joseph Bayer [168], and it has since changed very little over a century [172] and China with 20.3 Mt of alumina in 2012 remains the world's largest alumina producer [171]. In this process the use of hot caustic soda is essential because it dissolves alumina, allowing the dissolved solution to be separated from undissolved waste for further processing. As a general explanation of the Bayer process, crushed bauxite is subjected to high-temperature digestion in a concentrated caustic solution. The resulting liquor (pregnant liquor), is then clarified and filtered to remove the mud and other insoluble residues. The clarified pregnant liquor is then cooled and seeded with solid aluminum hydroxide to precipitate the alumina. The resulting spent liquor is concentrated by evaporation and recycled for further bauxite digestion [172]–[174] which is an important part of the Bayer process in alumina refinery plants, both to make water for residue washing process, and to re-concentrate the spent liquor for the bauxite digestion process. This process consumes a significant amount of thermal energy. The evaporation section is also responsible for controlling the water balance in the whole plant. Its main heat source is the available live steam and as compared to the other sections of the plant, it is the most energy intensive process, accounting for approximately 25-30% of energy consumption per ton of alumina [160]. Therefore alumina refineries are large users of gas, coal and other relevant fuels for their energy requirements. For example in Australia in 2012, alumina refineries consumed 221 PJ of energy [171]. Any improvement to this section that can decrease steam consumption, which in turn saves fuel consumption in the steam plant, will significantly reduce
greenhouse gas emission and production cost. In 2012, 0.7 tonnes of CO$_2$-e was produced per tonne of alumina in Australia [171].

Evaporation units in alumina refineries typically include tube falling film or flash vessel techniques. The design of an evaporator is purely based on its suitability to handle process liquor which is viscous and prone to scaling. Recently plate type falling film evaporator is also used by Aluminum Company of Shanxi and Guizhou Alumina [161], [162] to improve the efficiency in evaporation sections with low concentration feed streams. In this regard our state-of-the-art pilot plant consisting of two serially connected Alfa Laval's single-effect rising film plate evaporator modules [110] is also being tested under realistic process conditions to investigate potential scaling issues when it is coupled to different process streams (Figure 6-1).

![Figure 6-1 The two serially connected Alfa Laval's single-effect rising film plate evaporator modules (JWP-16-C Series)](image)

Both flashing and evaporation techniques have been shown in Figures 6-2 and 6-3. As it is shown, a combination of heat exchangers and flash vessels known as Multi Stage Flash (MSF) can be used. In this system, the feed liquor is firstly preheated by going through a series of heat exchangers (preheaters) and is then introduced to a battery of flash vessels. The flashed vapours from the vessels are condensed in the preheaters, and the highly concentrated feed from the last outlet is the final product of the evaporation unit. The first heat exchanger in this system uses the available live steam as the heat source.

Instead of flash vessels, the falling film technique with a higher thermal performance [27] can be used. However, it is more prone to scaling, especially at
the higher temperature as compared to the Multi Stage Flash (MSF) process. In this technique (Figure 6-3), the feed liquor distributes onto the evaporator from the top. Gravity pulls the film down along the inner walls of the evaporator tubes, so that the fast moving and thinning film gives rise to a high heat transfer coefficient [163]. At the bottom of the heat exchanger the highly concentrated feed stream is collected. The outgoing vapour from the first effect can be the heat source for the second effect, and this trend continues through to the last condenser which makes up a Multi Effect Evaporation (MEE) process.

As it is shown in Figure 6-2, in the MSF configuration, the vapour which is produced in the last flash vessel is injected into the barometric condenser and is mixed with the cooling water supply stream. In the selected evaporation plant, the inlet temperature of the vapour to the barometric condenser is suitable to run a Multi Effect Evaporation-Thermal Vapour Compression (MEE-TVC) process with the aim of saving energy. This chapter introduces a novel process that can ultimately be coupled with these available waste heat streams in the evaporation units of an alumina refinery plant to recover the load of a selected evaporation
unit, so as to reduce the live steam consumption (which is the main heat source of the evaporation process) as compared to both the current MSF process and the conventional MEE-TVC process.

### 6.2. Process Description

The selected evaporation plant includes some evaporation units, which uses the flashing technique. The purpose of our novel design is to utilise the available flashed vapour waste stream to recover the load of a selected evaporation unit and eliminate its live steam consumption. Table 6-1 summarises the selected evaporation units' specifications. However in view of confidentiality, the quantities that are shown are different from actual specifications, but are meant to be generic and representative. For this purpose the simulation has been done for unit mass flowrate of live steam. Also the feed liquor stream has been replaced with pure caustic soda (NaOH) solution which is more conservative due to its higher Boiling Point Elevation (\(BPE\)) as compared to the actual feed liquor stream.

<table>
<thead>
<tr>
<th>Table 6-1 Assumed specifications</th>
</tr>
</thead>
<tbody>
<tr>
<td>Live Steam</td>
</tr>
<tr>
<td></td>
</tr>
<tr>
<td></td>
</tr>
<tr>
<td>Feed Inlet (NaOH)</td>
</tr>
<tr>
<td></td>
</tr>
<tr>
<td></td>
</tr>
<tr>
<td>Equivalent Process Liquor Outlet</td>
</tr>
<tr>
<td>(highly concentrated NaOH)</td>
</tr>
<tr>
<td>Available Waste Heat Stream</td>
</tr>
<tr>
<td>(Flashed Vapour)</td>
</tr>
<tr>
<td></td>
</tr>
<tr>
<td>Cooling Water Supply</td>
</tr>
</tbody>
</table>

As shown in Table 6-1, the selected MSF evaporation unit is considered to consume 1 kg/s live steam to re-concentrate 18.45 kg/s of 19.0wt% NaOH feed stream to 16.30 kg/s of 21.5wt% NaOH. As shown in Figure 6-2 the considered available waste stream is the flashed vapour from the last flash vessel of the other evaporation unit. This stream has enough potential to recover the load of the selected unit. For benchmarking purpose an optimised conventional TVC-MEE process as shown in Figure 6-4 is considered to harness the available waste heat source. Based on our process simulation, which is detailed below, it consumes 35% of the available live steam (0.35 kg/s as opposed to the original 1 kg/s consumption) via a thermal vapour compressor (TVC) to compress around 56%
of the abovementioned available flashed vapour in order to recover the duty of the selected evaporation unit. The remaining flashed vapour (namely, 44%) from the abovementioned flashed vapour stream is then purged as a waste stream to the extant barometric condenser. The process parameters in Figure 6-4, other than those specifications as in Table 6-1, are the results of our process simulation.

In contrast, as shown in Figure 6-5, our Flash Boosted Thermal Vapour Compression Multi Effect Evaporation (FB-TVC-MEE) process is used to harvest the same flashed vapour stream. In this process, our simulation as detailed in the section below indicates that only around 18% of the selected unit live steam (0.18 kg/s as opposed to the original 1 kg/s consumption) is needed to compress 28% of the available flashed vapour stream. The remaining flashed vapour (namely 72%) is then summarily injected into the second effect of the MEE system. This augmentative steam injection process has been elucidated in our previous works known respectively as the Boosted MEE (B-MEE) and Flash Boosted MEE (FB-MEE) processes [2], [4], [95], [96], [98], [107]–[111]. Essentially this is a special application of our earlier reported FB-MEE scheme that is coupled to the TVC, whereby the primary MEE process is limited to two effects, and the battery of flash chambers being reduced to one flash chamber. Again the process parameters in Figure 6-5, other than those specifications as in Table 6-1, are the results of our simulation.
Chapter 6. A Novel Flash Boosted Thermal Vapour Compression MEE Process for Alumina Refineries

Bijan Rahimi

Figure 6-5 Schematic design of the FB-TVC-MEE process coupled with the available waste stream of the considered alumina refinery plant. Other than plant specifications, all figures are based on our simulations.

The virtue of flash vapour injection in the FB-TVC-MEE process is to minimise the suction load of the TVC which then economises on live steam consumption. As shown in Figures 6-4 and 6-5, the live steam consumption of the novel FB-TVC-MEE process is 49% less than the optimised conventional TVC-MEE process (0.18 kg/s as compared with 0.35 kg/s live steam consumption). Due to the higher flashed vapour recovery of the FB-TVC-MEE process as compared to the conventional TVC-MEE process, the process condensate (wash water) production rate increases by 23%.

This novel FB-TVC-MEE process will be compared with the conventional TVC-MEE process and the current MSF process, in terms of the amount of saved live steam, Gain Output Ratio ($\text{GOR}$), specific pumping power consumption and specific capital cost.

6.3. Mathematical Simulation and Capital Cost Analysis

6.3.1 Technical simulation

Our simulations encompass mass, concentration and energy balances for the MEE effects, the condenser and the thermo-compressor. The Generalised Reduced Gradient (GRG) method [115] is used to solve the accompanying equations by imposing boundary conditions stemming from the law of thermodynamics and appropriate operational, technical and economic...
constraints as shown in Table 6-2. Validated simulations and their methodologies have been reported in details in our previous works [2], [95], [96], [107], [109]–[111]. The basic simulation has been validated against market available data from a reputable manufacturer [95], [175] and our pilot plant [110] (Figure 6-1).

<table>
<thead>
<tr>
<th>Table 6-2 Assumptions and boundary conditions of our thermodynamic simulation</th>
</tr>
</thead>
<tbody>
<tr>
<td>Recovery factor for each MEE effect ($1/R$)</td>
</tr>
<tr>
<td>Temperature difference across the condenser, ($^\circ$C)</td>
</tr>
<tr>
<td>Minimum pressure difference for steam injection purpose (kPa)</td>
</tr>
<tr>
<td>Temperature of approach across the evaporators, ($^\circ$C)</td>
</tr>
<tr>
<td>Feed (NaOH) inlet/outlet concentration (wt %)</td>
</tr>
<tr>
<td>Maximum available waste flashed vapour mass flowrate (kg/s)</td>
</tr>
<tr>
<td>Maximum available live steam mass flowrate use as the motive steam in the thermo-compressor (kg/s)</td>
</tr>
</tbody>
</table>

Due to the high boiling point elevation ($BPE$) of NaOH as the feed liquor, which is 9.4°C for pure 21.5wt% NaOH, and the optimised saturated temperature difference between the steam after vapour compression and the cooling water supply (heat sink), the number of effects in the MEE process is limited to two effects (Figures 6-4 and 6-5). The feed liquor (NaOH) enters the MEE effects at 19.0wt% and 58.4°C and is concentrated to 21.5wt%, being identical to the extant inlet and outlet conditions of the selected evaporation unit for process liquor. A low recovery factor (12%) is chosen to match the extant flashing process so as to harmonise with the entire operation. We note however that the potential of an MEE process, in terms of recovery factor, is much more than that [27].

The simulation aims to find the lowest possible required live steam mass flowrate (to wit, the motive steam of the thermo-compressor), the required TVC suction flowrate and the injected vapour mass flowrate in order to have the same concentrate production rate as the selected evaporation unit. It is also calculated that 70.7°C saturated steam temperature is the corresponding optimised heat source inlet temperature to have a two-effect MEE to meet the required concentrate production rate.

The mass and concentration balances are expressed as, $k \in \{1, 2\}$:

$$\dot{m}_{F,k} = \dot{m}_{HC,k} + \dot{m}_{V,k} \quad (6-1)$$
\[ \dot{m}_{F,k} \cdot X_{F,k} = \dot{m}_{HC,k} \cdot X_{HC,k} \]  

(6-2)

for which:

\[ \dot{m}_{HC,k} = \left( \frac{R-1}{R} \right) \cdot \dot{m}_{F,k} \]  

(6-3)

\[ X_{HC,k} = \left( \frac{R}{R-1} \right) \cdot X_{F,k} \]  

(6-4)

The energy balance for the first effect is written as:

\[ \dot{m}_{HS,1} \cdot \left( h_{HS,1,in} - h_{f,(P_d)} \right) = \dot{m}_{V,1} \cdot h_{V,1} + \dot{m}_{HC,1} \cdot h_{HC,1} - \dot{m}_{F,1} \cdot h_{F,1} = U_1 \cdot A_1 \cdot \Delta T_{lm,1} \]  

(6-5)

where \( h_{f,(P_d)} \) is the saturated enthalpy of the discharge steam condensate upon leaving the first effect; \( h_{HS,1,in} = h_{g,(P_d,T_d)} \) is the enthalpy of superheated steam at discharge pressure and temperature; \( h_{V,1} = h_{g,(P_1,T_{HC,1})} \) is the superheated vapour at effect’s pressure and the outlet high concentrate temperature; \( h_{HC,1} \) and \( h_{F,1} \) are the enthalpies of the high concentrate outlet and feed inlet streams at their relevant concentration and temperature, respectively.

The energy balance for the second effect which receives waste flash vapour by injection, in the case of FB-TVC-MEE, is expressed as:

\[ \dot{m}_{HS,2} \cdot \left( h_{HS,2,in} - h_{f,(P_1)} \right) = \dot{m}_{V,2} \cdot h_{V,2} + \dot{m}_{HC,2} \cdot h_{HC,2} - \dot{m}_{F,2} \cdot h_{F,2} = U_2 \cdot A_2 \cdot \Delta T_{lm,2} \]  

(6-6)

For which,

\[ \dot{m}_{HS,2} = \dot{m}_{V,1} + \theta \dot{m}_{V,inj} \]  

(6-7)

and,

\[ h_{HS,2,in} = \frac{\dot{m}_{V,1} \cdot h_{V,1} + \theta \dot{m}_{V,inj} \cdot h_{V,inj}}{\dot{m}_{V,1} + \theta \dot{m}_{V,inj}} \]  

(6-8)

\[ h_{V,2} = h_{g,(P_2,T_{HC,2})} \]  

(6-9)

\[ h_{HC,2} = h_{f,HC,(X_{HC,2},T_{HC,2})} \]  

(6-10)

\[ h_{F,2} = h_{f,F,(X_{F,2},T_{F,2})} \]  

(6-11)
\( \theta \) is a flag that distinguishes TVC-MEE from the FB-TVC-MEE process, so that \((\theta = 0)\) for TVC-MEE and \((\theta = 1)\) for FB-TVC-MEE.

The energy balance for the condenser is written as:

\[
\dot{m}_{V,2} \cdot (h_{V,2} - h_{f(p_2)}) = \dot{m}_C \cdot (h_{C,\text{out}} - h_{C,\text{in}}) = U_{\text{cond}} \cdot A_{\text{cond}} \cdot \Delta T_{\text{lm,cond}} \tag{6-12}
\]

where \( h_{C,\text{in}} \) and \( h_{C,\text{out}} \) are the enthalpies of the inlet and outlet cooling water at their relevant temperatures, respectively.

Figures 6-6 and 6-7 illustrate the temperature-energy profiles across the first effect, second effect and condenser for the conventional TVC-MEE and the FB-TVC-MEE processes, with the values stemming from our simulation. The heat source of these effects and the condenser is the superheated vapour from the previous effect or the discharge steam from the thermo-compressor. Referring to the first effect (Figures 6-6 and 6-7), in zone 1 the superheated vapour first de-superheats and then condenses over zones 2 and 3. In contrast, the feed temperature increases to the relevant boiling temperature and remains constant over the evaporation process. In this effect (namely effect 1), there are 3 different zones with three different modes of heat transfer. In zone 3, the feed is boiling and therefore its concentration is changing from 19.0% to 21.5% (Table 6-2), and because the boiling point elevation is a function of the concentration, \( BPE \) is changing during the evaporation process, so this process as shown in Figures 6-6 and 6-7 is not a constant temperature evaporation process. The \( UA \) values for zones 1 and 2 are not more than 8% of the total \( UA \) value in this effect for the FB-TVC-MEE process (3% for the conventional TVC-MEE process), and therefore to a good approximation their relevant heat transfer coefficients can be ignored, so that the overall \( U \) value for this effect \( (U_1) \) is simply \( U_{z3} \), which pertains to zone 3, where latent-latent heat transfer prevails (see Appendix C).

In the second effect, there are two zones. In zone 1 the superheated vapour first de-superheats and then condenses over zone 2. In our simulation for the second effect the impact of flashed feed stream is ignored. Flashing happens when the feed saturation pressure is higher than the pressure of the second effect. Hence a negligible amount of feed (less than 1%, that commensurate with 5.7°C of sensible heat in our case), is flashed while its pressure and temperature decrease
(namely, from 59°C to 53.3°C). For the second effect in both the FB-TVC-MEE and TVC-MEE the inlet feed temperature is considered to be 53.3°C, namely to the relevant feed boiling point temperature (which commensurate with its concentration) that corresponds to the pressure in the second effect (Figures 6-6 and 6-7).

After flashing the feed is boiling while its concentration is changing from 19.0% to 21.5% (Table 6-2). As mentioned above due to the different BPE, the temperature is varying during the evaporation process. Again, as shown in Figures 6-6 and 6-7, to a very good approximation the overall $U$ value for this effect ($U_2$) is simply $U_{z2}$.

Referring to Figures 6-6 and 6-7, the cooling water temperature increases across the condenser but the trend of the superheated vapour temperature profile is the same as in the previous effects, so that zone 1 (namely the de-superheating zone) is negligible once again, with the overall heat transfer coefficient ($U_{\text{cond.}}$) being practically $U_{z2}$, where latent-sensible heat transfer happens (See Appendix C).

In our simulation as a conservative assumption a 3°C temperature difference has been considered between the concentrate outlet boiling temperature and the heat source condensation temperature.

For the TVC, the required motive steam and produced discharge steam are calculated according to Figure 6-8 [176], with the assumed suction ($P_s$), motive ($P_m$)and discharge ($P_d$) pressures, and the required expansion ($\text{EXP} = P_m/P_d$), compression ($\text{CMP} = P_d/P_s$) and entrainment ($\text{ENT}$) ratios. The two equations below are used to calculate the required motive and suction mass flowrates.

$$\dot{m}_m = \dot{m}_s/\text{ENT} \quad (6-13)$$

where $\text{ENT}$ is the entrainment ratio to be found from Figure 6-8 [176]. And

$$\dot{m}_d = \dot{m}_s + \dot{m}_m \quad (6-14)$$

The discharge enthalpy and temperature can be found from the energy balance equation which is written as:
where, \( h_m = h_g(P_m, T_m) \) and \( h_s = h_g(P_s, T_s) \). The discharge temperature \( T_d = f(h_d, P_d) \), which is a function of the discharge pressure and enthalpy, can be obtained from steam property routine. This temperature shows the degree of superheating which enjoins extra heat exchanger area to cater for de-superheating.

\[
h_d = \frac{\dot{m}_m h_m + \dot{m}_s h_s}{\dot{m}_s + \dot{m}_m} \quad (6-15)
\]

The overall amount of wash water generated for the sum total of the condensate flowrate from each of the MEE effects:

\[
\dot{M}_{D, \text{total}} = \sum_{k=1}^{2} \dot{m}_{V,k} + \theta \cdot \dot{m}_{V, \text{inj}} + \dot{m}_s \quad (6-16)
\]

\[
\dot{M}_{HC, \text{total}} = \sum_{k=1}^{2} \dot{m}_{HC,k} \quad (6-17)
\]
As mentioned above, the Generalised Reduced Gradient (GRG) method is used to solve equations (6-1) to (6-17) to find the lowest possible motive steam flowrate to produce the same highly concentrated flow rate as the conventional MSF process (The original process) by imposing the boundary conditions for both the conventional TVC-MEE and FB-TVC-MEE processes. In all the above equations, and for the properties of steam and vapour, the REFPROP package (REference fluid PROPerties) as developed by the National Institute of Standards and
Technology (NIST) [116] has been used. The properties of NaOH are consulted from [1], [118]–[123]. Extensive details related to the simulation of the MEE process can be found in [2], [27].

6.3.2 Capital cost analysis

The overall capital cost is a key parameter for the financial decision makers. For conventional thermal desalination plants the capital cost can be approximated based on the GWI/IDA database [55]. For the capacity range up to 10,000 m$^3$/day, the following conservative cost function applies [3].

$$\Psi_{D_t} = 3054 \times D_t^{0.9751} \quad (6-18)$$

where $D_t$ is the total distillate production rate of the plant in (m$^3$/day) and $\Psi_{D_t}$ is the conservative cost function that yields the total capital cost of the relevant plant. The capital costs for thermal distillation processes (MSF and MEE systems) can be divided into two main sections [165], [166], namely capital cost of the evaporator (40% of the total capital cost), and remaining items that include equipment piping, erection, engineering and commissioning as well as electrical, instrumentation and control (60%). While this cost function stems from seawater applications with different choice of material compared with process liquor operation (i.e. the use of cast iron), but since we are only interested in percentage cost differential comparison between two thermal distillation processes, the comparative procedure remains valid.

Equation (6-18) is used to calculate the capital cost of TVC-MEE as a function of process condensate (distillate) production rate, but for FB-TVC-MEE because of the injected flashed vapour to the second effect, its impact on the capital cost should be evaluated. This procedure has been developed in our previous work [2], [3]. Accordingly, the capital cost of an FB-TVC-MEE system can be analysed with the help of the full cost function of Equation (6-18) by considering the process condensate capacity of the FB-TVC-MEE system, save the injected flashed vapour flowrate to the second effect. For the vapour injected from the waste flashed vapour stream, only condensation takes place in the MEE section and no additional evaporation surface needs to be considered for the vapour generation in this section, insofar as capital cost analysis is concerned. By assuming similar
overall heat transfer coefficients for both evaporation and condensation processes [27], only half of the typical heat transfer area is therefore needed to process the additional vapour injection, which is then translated to a 50% discount of the earlier introduced cost breakdown for the evaporator (namely 50% of 40% for the cost of evaporator) [2][3]. This overall factor is then applied to the capital cost differential between an MEE plant with a process condensate production capacity of an FB-TVC-MEE plant and an MEE plant with a process condensate production capacity of an FB-MEE plant less the total vapour injection rate from the flashed vapour waste stream as shown in Equation (6-19) [2],[3].

\[ CC_{(FB-TVC-MEE)} = \Psi(D_{t(\text{FB-TVC-MEE})} - D_{\text{inj}}) + 0.2 \cdot (\Psi(D_{t(\text{FB-TVC-MEE})} - D_{\text{inj}}) - \Psi(D_{t(\text{FB-TVC-MEE})} - D_{\text{inj}})) \] (6-19)

where \( D_{t(\text{FB-TVC-MEE})} \) is the total process condensate (distillate) flow rate (m³/day) of the FB-TVC-MEE and \( D_{\text{inj.}} \) is the injected flashed vapour flow rate (m³/day) into the second effect of the MEE section.

The discharge steam from the TVC, and for that matter, all the vapours (namely injected flashed vapour and vapour produced from the MEE process) has an inherent high degree of superheat. Consequently extra heat exchanger area for de-superheating in those evaporation effects has to be considered, and gives rise to a higher capital cost that impact on the evaporator section which constitutes 40% of the total plant capital cost [165], [166]. From practical experience [164], 1% extra heating area is needed per 1°C of superheat in order to account for both de-superheating and future scaling problem. For this purpose, according to the degree of superheat, a corresponding coefficient factor is therefore applied to cover the extra heat exchanger area of the evaporation effects as follow [9].

\[ A_{\text{calc.}} = A_1 + A_2 + A_{\text{cond}} \] (6-20)
\[ A_{\text{actual}} = \beta_1 \cdot A_1 + \beta_2 \cdot A_2 + \beta_{\text{cond}} \cdot A_{\text{cond}} \] (6-21)

and,

\[ Y = \frac{A_{\text{actual}}}{A_{\text{calc.}}} \] (6-22)

where \( \beta \) is the local evaporator coefficient factor which is considered as the extra percentage of required area per degree of superheat. \( A_{\text{calc.}} \) and \( A_{\text{actual}} \) (m²) are
the total heat exchanger areas of the MEE section with and without considering
the impact of superheated steam, respectively, and $\gamma$ is the overall area
coefficient factor that should be applied to the capital cost of evaporator (40% of
the total capital cost), and so the resultant total capital cost equation is as follow
[9].

\[
TCC\ (US\$) = (0.4 \cdot \gamma + 0.6) \cdot (CC) \tag{6-23}
\]

where $CC$ can be the capital cost for either TVC-MEE (Equation 6-18) or FB-TVC-
MEE (Equation 6-19) process. The specific capital cost can then be calculated
by Equation (6-24). Namely

\[
SCC\ (US$/\text{m}^3/\text{day}) = \frac{TCC}{D_t} \tag{6-24}
\]

where $D_t$ is the process condensate (distillate) production rate ($\text{m}^3$/day).

6.4. Results and Discussion

Based on all the abovementioned methods and assumptions, Table 6-3 and
Figure 6-9 illustrate the amount of steam saving of the two proposed processes
as compared to the current process. Evidently, the FB-TVC-MEE process is
superior with around 82% saving of the available live steam and its ability to
recover all available flashed vapour waste streams.

| Table 6-3 Available live steam, steam consumptions and recovered flashed vapour for the
considered processes |
<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Available Live Steam (For the Selected Evaporation Unit (MSF))</td>
<td>1.00</td>
<td>1.00</td>
<td>1.00</td>
</tr>
<tr>
<td>Live Steam Consumption (kg/s)</td>
<td>1.00</td>
<td>0.35</td>
<td>0.18</td>
</tr>
<tr>
<td>Available Flashed Vapour Waste Stream (kg/s)</td>
<td>1.55</td>
<td>1.55</td>
<td>1.55</td>
</tr>
<tr>
<td>Recovered Flashed Vapour Waste Stream (kg/s)</td>
<td>0</td>
<td>0.87</td>
<td>1.55</td>
</tr>
</tbody>
</table>

For the optimised TVC-MEE process in Figure 6-4, there is no steam injection
into the second effect and consequently around 44% of the flashed vapour waste
stream remains unused (0.68 kg/s out of 1.55 kg/s). Figure 6-5 shows an
optimised FB-TVC-MEE process which fully consumes all the flashed vapour
waste stream, whereby 28% of which is subjected to TVC and the balance is summarily injected into the second effect. Figure 6-10 elucidates the optimisation pathway progressing from an optimised TVC-MEE to an optimised FB-TVC-MEE. Verily the live steam consumption (motive steam flowrate) decreases linearly from 0.35 kg/s for the optimised TVC-MEE (which corresponds to nil injected flashed vapour flowrate) through to 0.18 kg/s for the optimised FB-TVC-MEE process (at 1.11 kg/s injected flashed vapour flowrate). Concomitantly the amount of unused flashed vapour waste decreases linearly to zero as the process evolves from an optimised TVC-MEE to an optimised FB-TVC-MEE.

The Gain Output Ratio (GOR) is a key factor which compares the thermal performance of all three processes. This factor is defined as the amount of favorable production which is produced by one unit of consumed live steam.

\[
GOR = \frac{\dot{m}_p}{\dot{m}_m}
\]  

(6-25)

where \(\dot{m}_m\) is the motive steam flowrate and, \(\dot{m}_p\) is the total produced process condensate excluding the injected vapour from the waste stream, as the latter has always been available. Figure 6-11 shows the GOR for all three considered evaporation processes. Verily, the GOR of the FB-TVC-MEE is the highest and around 5.7 times more than the selected evaporation unit and 98% better than the TVC-MEE process which means it can utilise the live steam more effectively.
Figure 6-10 Optimisation trajectory as the process evolves from an optimised TVC-MEE to an optimised FB-TVC-MEE.

With respect to pumping power consumption, as indicated in Figures 6-4 and 6-5, there are five essential pumps to supply the cooling water (pump no. 1), extract the process condensate (pump no. 2) and concentrate (pump no. 3) from the MEE effects, feed the effects (pump no. 4) and extract steam condensate from the first effect (pump no. 5). The pumping power is calculated according to the equation below.

\[ \text{Pumping Power (kW)} = \left( \frac{\Delta P \cdot \dot{V}}{\eta} \right) \]  

(6-26)

where \( \Delta P \), \( \dot{V} \) and \( \eta \) are the total head (kPa), volumetric flow rate (m\(^3\)/s) and the overall efficiency of the pump, respectively. Table 6-4 includes the considered pressure heads and efficiencies for the relevant abovementioned pumps. The specific pumping power of FB-TVC-MEE is compared with TVC-MEE as in Figure 6-12. A detailed calculation of power consumption and its methodology has been explained in [2], as well. As shown in Figure 6-12, the main duty is related to the cooling water supply pump (pump no. 1). With respect to the process condensate production rate, due to the higher flashed vapour recovery in FB-TVC-MEE as compared to the TVC-MEE process, the specific pumping power is only 5% more than that of TVC-MEE.
Chapter 6. A Novel Flash Boosted Thermal Vapour Compression MEE Process for Alumina Refineries

Table 6-4 Pump specifications

<table>
<thead>
<tr>
<th>Pump No.</th>
<th>Duty</th>
<th>Overall Efficiency</th>
<th>Pressure Head (barg)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Cooling water intake pump</td>
<td>0.70</td>
<td>5.1</td>
</tr>
<tr>
<td>2</td>
<td>Process condensate extraction pump</td>
<td>0.70</td>
<td>8.5</td>
</tr>
<tr>
<td>3</td>
<td>High concentrate liquor extraction pump</td>
<td>0.63</td>
<td>12.3</td>
</tr>
<tr>
<td>4</td>
<td>Feed pump</td>
<td>0.77</td>
<td>1.0</td>
</tr>
<tr>
<td>5</td>
<td>Steam condensate pump</td>
<td>0.60</td>
<td>12.2</td>
</tr>
</tbody>
</table>

Following the abovementioned method for comparing the capital costs of both TVC-MEE and FB-TVC-MEE processes as in section 6.3, Table 6-5 delineates the extent of superheating in each effect of both processes which influence both the local evaporator ($\beta$) and overall area ($\gamma$) coefficient factors. Accordingly, the local evaporator coefficients ($\beta$) for the first effect, second effect and the condenser of the TVC-MEE process are 1.252, 1.094 and 1.094, respectively. As for the FB-TVC-MEE, they are 1.252, 1.118 and 1.094, respectively. By applying these factors to Equation (6-21) and then Equation (6-22) the overall area coefficient factor ($\gamma$) are 1.16 for the TVC-MEE and 1.14 for the FB-TVC-MEE, and with which the total and specific capital costs can be evaluated according to Equations (6-23) and (6-24). Based on this procedure, and on account of the higher recovery of the flashed vapour waste stream in FB-TVC-MEE as compared to the TVC-MEE, the specific capital cost is around 11% less than that of TVC-MEE process.

![Figure 6-11 Gain Output Ratio (GOR) of the current MSF, TVC-MEE and FB-TVC-MEE processes](image)

Table 6-5 Degree of superheat in each effect

<table>
<thead>
<tr>
<th></th>
<th>First Effect</th>
<th>Second Effect</th>
<th>Condenser</th>
</tr>
</thead>
<tbody>
<tr>
<td>TVC-MEE</td>
<td>25.2 (°C)</td>
<td>9.4 (°C)</td>
<td>9.4 (°C)</td>
</tr>
<tr>
<td>FB-TVC-MEE</td>
<td>25.2 (°C)</td>
<td>11.8 (°C)</td>
<td>9.4 (°C)</td>
</tr>
</tbody>
</table>
6.5. Conclusion

A novel FB-TVC-MEE process has been developed for the exploitation of the available waste heat streams in an alumina refinery plant. This system is meant to be coupled to the available flashed vapour waste streams, so as to recover the load of a selected evaporation unit. Its performance has been benchmarked with an optimised conventional TVC-MEE system and also the current selected unit in terms of process condensate (wash water), gain output ratio, specific pumping powers and specific capital costs reduction. This novel process can save around 82% of the selected evaporation unit’s available live steam. Its thermal performance is 468% better than the current system and 98% better than the TVC-MEE process. With respect to pumping power consumption, it shows a 5% increment for the FB-TVC-MEE as compared to the TVC-MEE process. A capital cost analysis indicates an 11% decrement for the overall specific capital cost, for the novel FB-TVC-MEE process as compared to the TVC-MEE process. Importantly FB-TVC-MEE saves 26% more live steam than the optimised conventional TVC-MEE, which makes this novel process a superior option, in terms of significantly slashing the operating costs and emission of the plant.


Bibliography


[115] L. S. Lasdon, A. D. Waren, A. Jain, M. Ratner, and J. Rice, “Design and testing of
Bibliography


Bibliography


Bibliography


Appendices

Appendix A. Enthalpies

Appendix B. BPE and NEA

Appendix C. Heat Transfer Coefficients
Appendices
Appendices

**Appendix A. Enthalpies**

Latent heat of evaporation [177]:

\[
h_{fg}(T_{vs}) = 2499.5698 - 2.20464 \cdot T_{vs} - 1.596 \times 10^{-3} \cdot T_{vs}^2 \tag{A.1}
\]

Seawater enthalpy [117]

\[
h_{f,sw} = h_{f,w} - X(a_1 + a_2 \cdot X + a_3 \cdot X^2 + a_4 \cdot X^3 + a_5 \cdot T_B + a_6 \cdot T_B^2 + a_7 \cdot T_B^3 + a_8 \cdot X \cdot T_B + a_9 \cdot X^2 \cdot T_B + a_{10} \cdot X \cdot T_B^2) \tag{A.2}
\]

Which is valid for \(10 \leq T_B \leq 120 \degree C\) and \(0 \leq X \leq 0.12\) kg/kg range with an accuracy of ±0.5%. \(h_{f,w}\) is the saturated enthalpy of pure water:

\[
h_{f,w} = 141.355 + 4202.07 \cdot T_B - 0.535 \cdot T_B^2 + 0.004 \cdot T_B^3 \tag{A.3}
\]

Which is valid for \(5 \leq T_B \leq 200 \degree C\) and an accuracy of ±0.02%.

And the constants are as below.

\[
\begin{align*}
a_1 &= -2.348 \times 10^4 \\
a_2 &= 3.152 \times 10^5 \\
a_3 &= 2.803 \times 10^6 \\
a_4 &= -1.446 \times 10^7 \\
a_5 &= 7.826 \times 10^3 \\
a_6 &= -4.417 \times 10^1 \\
a_7 &= 2.139 \times 10^{-1} \\
a_8 &= -1.991 \times 10^4 \\
a_9 &= 2.778 \times 10^4 \\
a_{10} &= 9.728 \times 10^1
\end{align*}
\]
Appendices

**Appendix B. BPE and NEA**

Boiling Point Elevation (BPE) for seawater is calculated by [117]:

\[ BPE = \lambda \cdot X^2 + \sigma \cdot X \]  
\[ (B.1) \]

\[ \lambda = -4.584 \times 10^{-4} \cdot T_B^2 + 2.823 \times 10^{-1} \cdot T_B + 17.95 \]  
\[ (B’.1) \]

\[ \sigma = 1.536 \times 10^{-4} \cdot T_B^2 + 5.267 \times 10^{-2} \cdot T_B + 6.56 \]  
\[ (B''.1) \]

The validity of equation (B.1) is for \( 0 \leq T_B \leq 200 \, ^\circ \text{C} \); \( 0 \leq X \leq 0.12 \, \text{kg/kg} \) range with an accuracy of ±0.018 K.

Boiling Point Elevation (BPE) of aqueous solutions can be found from the below chart [1]:

![Boiling Point Elevation (BPE) of aqueous solutions](image)

Non-Equilibrium Allowance (NEA) is calculated by [125]:

\[ NEA_{10} = (0.9784)^{T_v} \cdot (15.7378)^{W} \cdot (1.3777)^{W^{-3600\times10^{-6}}} \]  
\[ (B.2) \]

\[ W = 105.6 + 18.06 \cdot D_{FC} \]  
\[ (B.3) \]

(For the flashing chamber section: \( D_{FC} = \sum_{i=1}^{j} m_{v,FC,i} \))
Appendices

\[ NEA = \left( 0.5 \Delta T_{st} + NEA_{10} \right) \cdot \left[ \frac{NEA_{10}}{0.5 \cdot \Delta T_{st} + NEA_{10}} \right]^{0.3281} \cdot L \]  

(B.4)
Appendices

**Appendix C. Overall Heat Transfer Coefficient**

The calculation of the overall heat transfer coefficients across the first and the second effects and the condenser of the Flash Boosted Thermal Vapour Compression Multi-Effect Evaporation (FB-TVC-MEE) process are explained as follow.

Overall heat transfer coefficient across the condenser can be calculated from the equation below as a function of the saturated temperature of the inlet vapour [70], which is applicable to both TVC-MEE and FB-TVC-MEE processes:

\[ U_{\text{cond.}} = 1.7194 + 3.2063 \times 10^{-2} \times T_{\text{sat,cond.}} - 1.5971 \times 10^{-5} \times (T_{\text{sat,cond.}})^2 + 1.991 \times 10^{-7} \times (T_{\text{sat,cond.}})^3 \]  

(C.1)

The general equation for overall heat transfer coefficient of the falling film evaporators (first and second effects) is written as:

\[ \frac{1}{U} = (\frac{1}{\alpha_{\text{inn}}} + R_{F,\text{inn}}) \cdot \frac{r_o}{r_{\text{inn}}} + \frac{r_o}{\kappa_{\text{tube}}} \cdot \ln(\frac{r_o}{r_{\text{inn}}}) + \frac{1}{\alpha_o} + R_{F,o} \]  

(C.2)

In our case the overall heat transfer coefficient with a reasonable approximation can be considered as a function of the film coefficients (\(\alpha_i\) and \(\alpha_o\)) since the conduction resistance of the wall and fouling resistances (namely \(R_{F,i}\) and \(R_{F,o}\)) are considerably smaller. \(\frac{r_o}{r_i}\) has been considered to be 1.2 in our calculations which is based on typical diameter range for schedule 40 tube evaporators.

According to [178], \(\alpha_i\) is calculated based on (C.3).

\[ \alpha_i = \alpha^+ \cdot (\frac{\mu L^2}{\rho L^2 \cdot k L^3 \cdot g})^{-1/3} \]  

(C.3)

where:

\[ \alpha^+ = 1.6636 \cdot Re_L^{-0.2648} \cdot Pr_L^{0.1592}; 15 < Re_L < 3000; 2.5 < Pr_L < 200 \]  

(C.4)

The relevant heat transfer coefficient for the steam condensation on the outer tube can be calculated from Equations (C.5) or (C.6) [179], [180].

\[ \alpha_o = \frac{Re_L \cdot k_L}{1.08 \cdot Re_L^{1.22} - 5.2} \cdot \left(\frac{a}{\nu L^2}\right)^{1/3}; 30 < Re_L < 1800 \]  

(C.5)
Appendices

\[ \alpha_o = \frac{Re_L \cdot k_L}{8750 + 58 \cdot Pr_L^{-0.5} \cdot (Re_L^{0.75} - 253)^{1/3}} \cdot \left(\frac{g}{g_L^2}\right)^{1/3}; \quad Re_L > 1000 \quad (C.6) \]

The calculated heat transfer coefficients are shown in table C.1.

<table>
<thead>
<tr>
<th>MEE-TVC</th>
<th>FB-MEE-TVC</th>
</tr>
</thead>
<tbody>
<tr>
<td>1st Effect</td>
<td>2nd Effect</td>
</tr>
<tr>
<td>( \alpha_i )</td>
<td>3.8</td>
</tr>
<tr>
<td>( \alpha_o )</td>
<td>4.7</td>
</tr>
<tr>
<td>( U )</td>
<td>1.9</td>
</tr>
</tbody>
</table>