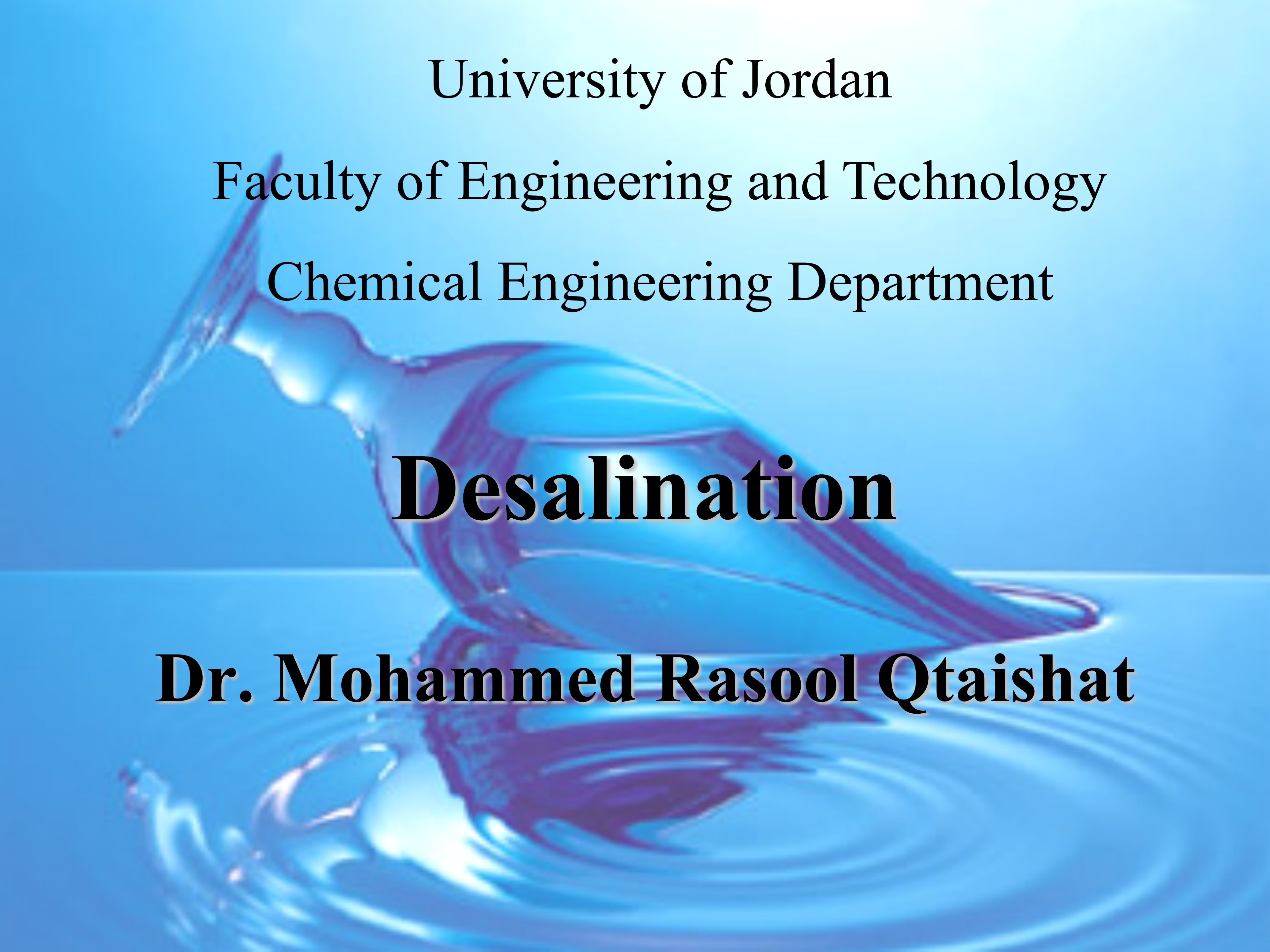


University of Jordan

Faculty of Engineering and Technology

Chemical Engineering Department

Desalination

Dr. Mohammed Rasool Qtaishat

Desalination Today

- 3% of earth water is fresh and available for human use; remaining 97% is sea water.
- More than 133 countries use desalination.
- 2/3 of the planet suffer from shortage of water.
- More than 17,000 desalination plants in the world.
- ~52% of world market is in the GCC States.

Desalination in the MENA Region

- Future water shortages warrant immediate action.
- Need for additional, reliable, and safe water supply for population growth and expansion in industry.
- Water supply is an issue of economic growth and national sustainability.
- Potable water quality.
- The cost of desalination is reducing.
- Water source for desalination: seawater and brackish water.

Desalination Processes

1. Membrane Processes

- Membrane desalination technologies.
- Operation and maintenance of RO plants.
- RO pretreatment (conventional, membrane, innovative) and post-treatment.
- Fouling and scaling in membrane processes.
- NF/RO modeling and process simulation.

Desalination Processes

2. Thermal Processes

- Thermal seawater desalination technologies.
- Operation and maintenance of thermal desalination plants and lifetime extension.
- Scaling, corrosion and material selection in thermal desalination plants.
- Design, construction and commissioning of thermal desalination plants.

Principles Salinity

Water Resources

Water consists of water molecules. The formula of water is well known namely H_2O.

Different water resources are available on earth:

- rainwater, snow;
- ice (glaciers, Antarctica);
- seawater;
- river water, lakes;
- ground water;
- domestic waste water;
- industrial waste water;

None of these water sources consists of pure water

Contaminants

All these waters may contain different types of contaminants e.g.

- natural contaminants;
- contaminants due human activities.

We may use other categories as well e.g.

- inorganic (e.g. salts)
- organic (e.g. natural organic matter originating from decay of e.g. trees, plants etc.)
- microbial e.g. bacteria, viruses, protozoan, algae etc.

Contaminants

Or we may distinguish in 3 groups according to their size:

Group	Size
dissolved	< 0.001µm or <1 nm
colloidal	0.001-1µm or 1-1000 nm
suspended	>1µm or >1000 nm

Salts

- Are solids
- Dissolve in water.
- Solubility, depends strongly on their nature and temperature.

What are salts?

- Salts consists of at least one cation (metal) and at least one anion.

NaCl	Sodium Chloride
$CaSO_4$	Calcium Sulphate
$CaCO_3$	Calcium Carbonate
$MgCl_2$	Magnesium Chloride
$NaNO_3$	Sodium Nitrate
CaF_2	Calcium Fluoride
$BaSO_4$	Barium Sulphate
$SrSO_4$	Strontium Sulphate

Solubility of salts in pure water in g/L

	Na	**Ca**	**Mg**	**K**	**Ba**	**Sr**
Cl	360	730	560	330	370	510
SO_4	170	2	350	110	0.002	0.11
NO_3	840	1220	740	300	90	70
CO_3	190	0.013	1	1080	0.02	0.01
F	45	0.016	0.076	930	1.6	0.1

Solubility of salts in pure water

Sodium salts e.g. Cl, SO_4, NO_3, CO_3 and F

- Are very well soluble.

Nitrates e.g. Na, Ca, Mg, K, Ba and Sr

- Are very well soluble.

Calcium and Strontium salts e.g. SO_4, CO_3 and F

- have very limited solubility.

Carbonates

- have very limited solubility for Ca, Mg, Ba and Sr.

Remark: Salts having low solubility may cause scaling in reverse osmosis plants due to precipitation.

Most important cations and anions

- Dissolved inorganic natural contaminants, namely salts, are responsible for the salinity. Salinity comes from cations and anions.

Cations	Anions
Sodium Na^+	Chloride Cl^-
Calcium Ca^{2+}	Sulphate SO_4^{-2}
Magnesium Mg^{2+}	Hydrogen Carbonate HCO_3^-
Potassium K^+	Nitrate NO_3^-

What are cations and anions ?

- **Since the solution is electrically neutral the total number of positive charges must be equal to the total number of negative charges.**
- **The number of charges carried by an ion, is equal to the valence of the atom.**
- **An ion may carry one valence e.g.**

  Na^+, K^+, Cl^- , HCO^-_3, F^-
- **Or two valences e.g.**

  Ca^{2+}, Mg^{2+}, Ba^{2+}, Sr^{2+} , SO_4^{-2}
- **Or three valences e.g.**

  PO_4^{3-}

How large are ions ?

- Ions are very small, some are even smaller than water molecules.
- Size of inorganic ions

H^+	0.053 nm	H_2O	0.33 nm
K^+	0.25	Cl^-	0.24
Na^+	0.37	NO_3^-	0.26
Ca^{2+}	0.62	HCO_3^-	0.42
Mg^{2+}	0.70	SO_4^{-2}	0.46

Electrical conductivity of water

- Pure water is a very poor conductor of electricity but when e.g. salts are dissolved in it, the solution conducts the electrical current.

Example: If you have the followings:

- A beaker glass with water and sodium(+) and chloride (-) ions.
- Two electrodes, cathode(-) anode (+)
- a battery
- lamp

Electrical conductivity of water

- An electrical current is flowing from the anode through the battery to the cathode.

- The electric current in the wire is carried by electrons.

- The electric current in the water is carried by the positively (Na^+) and negatively (Cl^-)charged ions.

- The presence of salts dissolved in water increases its conductivity (reduces its resistance), which varies depending on temperature.

Electrical conductivity of water

- In practice electrical conductivity is measured with a conductivity meter. The electrical conductivity is used as a surrogate parameter for salinity (Total Dissolved Salts) it is expressed as µS/cm.

- Rule of thumb: $TDS=K.EC_{25}$

Water type	EC_{25} µS/cm	Factor K
Permeate	1-10	0.5
	300-800	0.55
Sea water	45,000-60,000	0.7
Concentrate	65,000-85,000	0.75

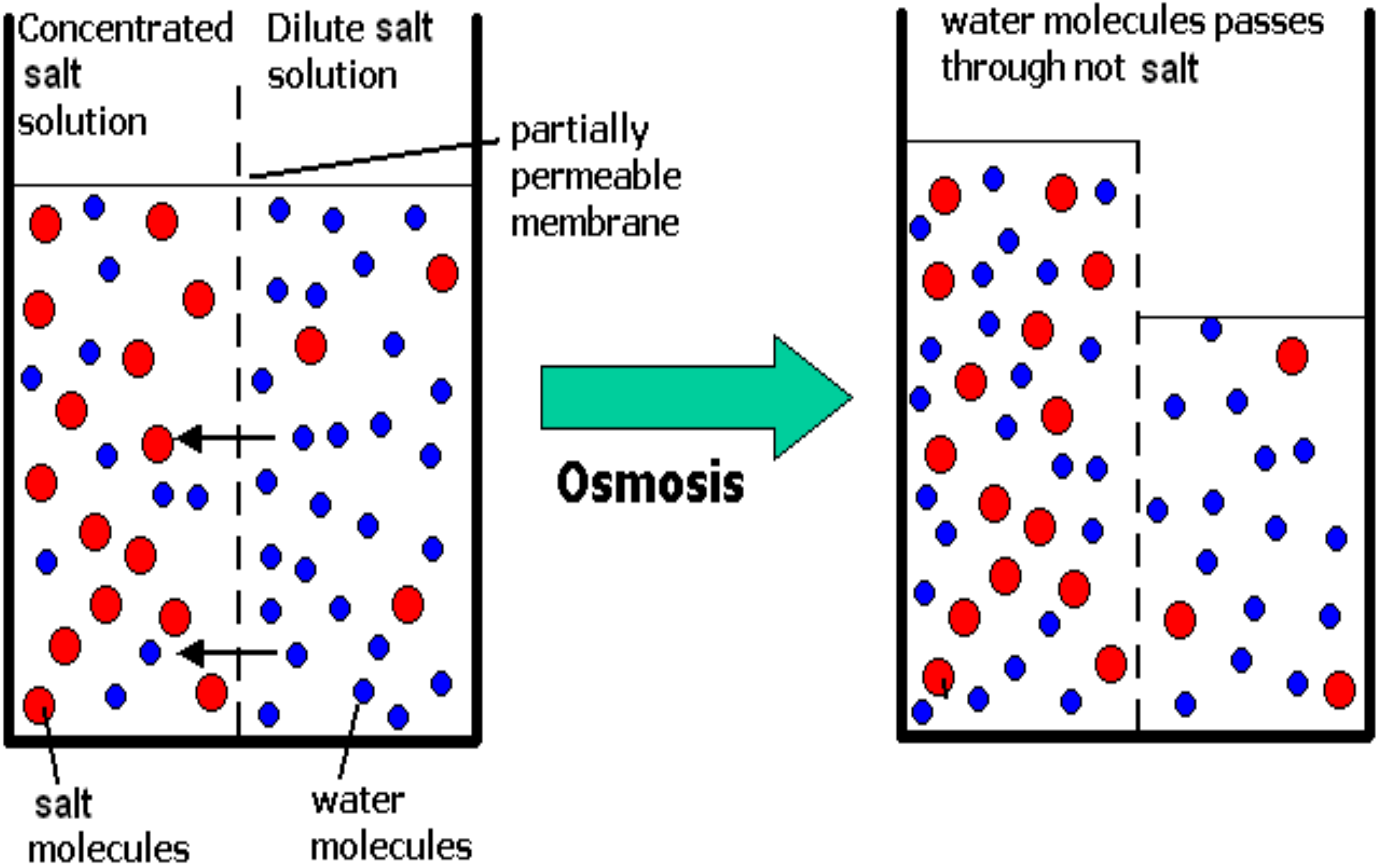
Osmosis and Osmotic pressure

- Osmotic is the diffusion of water through a semi permeable membrane (permeable to water and not salt) from a region of low concentration to a region of high concentration.

- Osmosis can be opposed by increasing the pressure in the region of high salt concentration with respect to that in the low concentration region.

- The pressure at which the water flow through the membrane is stopped is called osmotic pressure.

Osmosis


Concentrated salt solution
Dilute salt solution
partially permeable membrane
salt molecules
water molecules
Osmosis
water molecules passes through not salt

Osmotic pressure

- Several formulas are used for calculating osmotic pressure e.g.

$$\pi = \frac{C.(T+320)}{491000}\,bar \qquad \text{for C < 20,000mg/l}$$

$$\pi = \frac{(0.0117.\mathrm{C}) - 34}{14.23} \cdot \frac{T+320}{345}\,bar \qquad \text{for C > 20,000 mg/l}$$

Osmotic pressure

Or

$$\pi = R(T + 273)\sum m_i$$

$R = 0.082\ \ \mathrm{L.bar/mol.^{o}K}$

$\mathrm{T} = \mathrm{temperature}\ {}^{o}C$

$\sum\ = sum$

$m_i = molar$ concentration of all ions.

Questions

a) Calculate the osmotic pressure with formulas of a solution of 1000 mg NaCl /L;

b) Calculate the osmotic pressure with both formulas of sea water containing:

Chloride	18,890 mg/L	Sodium	10,556 mg/L
Sulphate	2,649	Magnesium	1272
Hydrogen		Calcium	400
Carbonate	140 mg/l	Potassium	380

Answers : 0.7bar and 0.8, and 25.4 bar and 26.2 bar

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Pressure Driven Membrane Processes

Dr.Mohammed Rasool Qtaishat

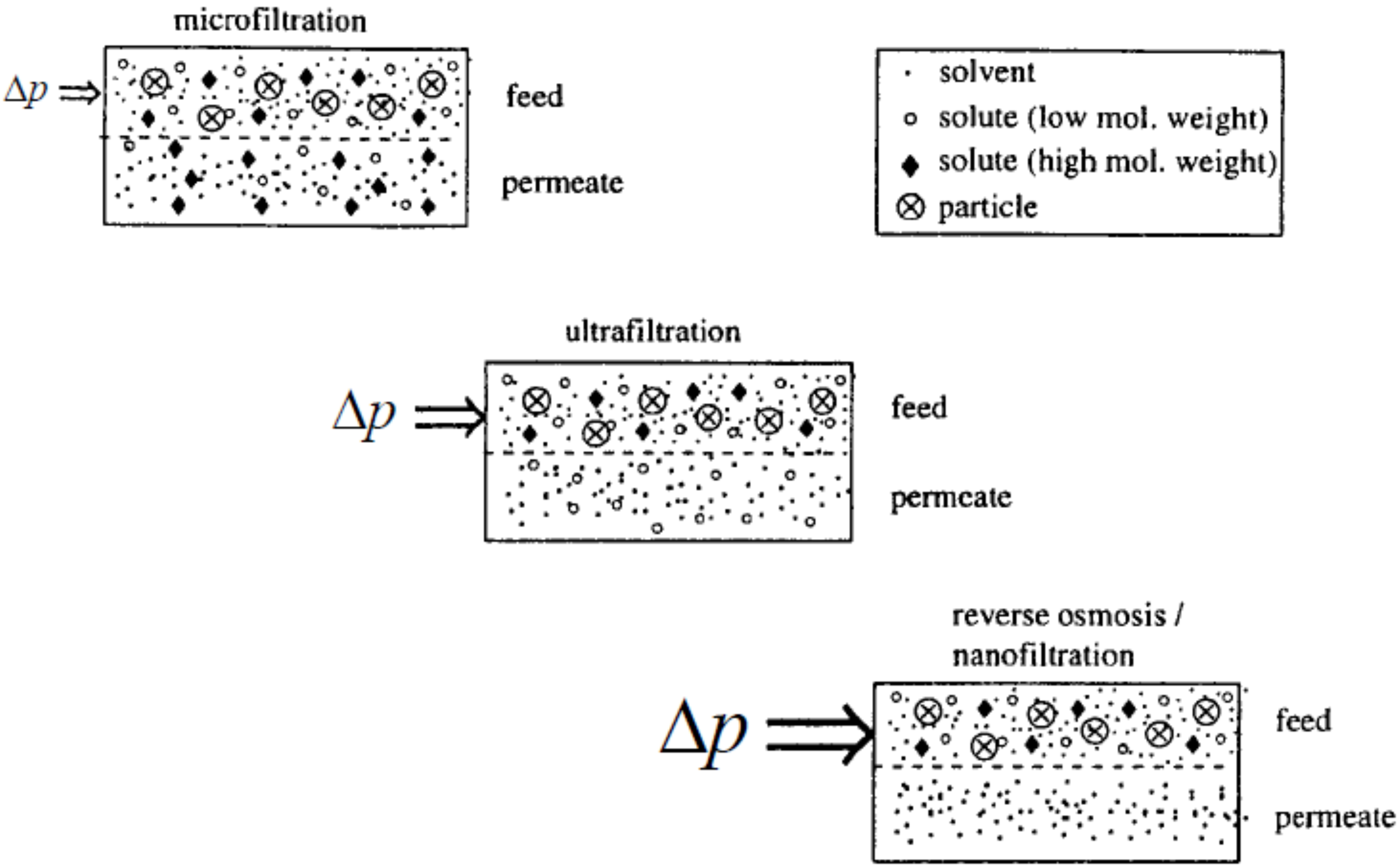
Introduction

- **Types and characteristics of pressure-driven separations**

⇒ The driving force is the **total pressure.**

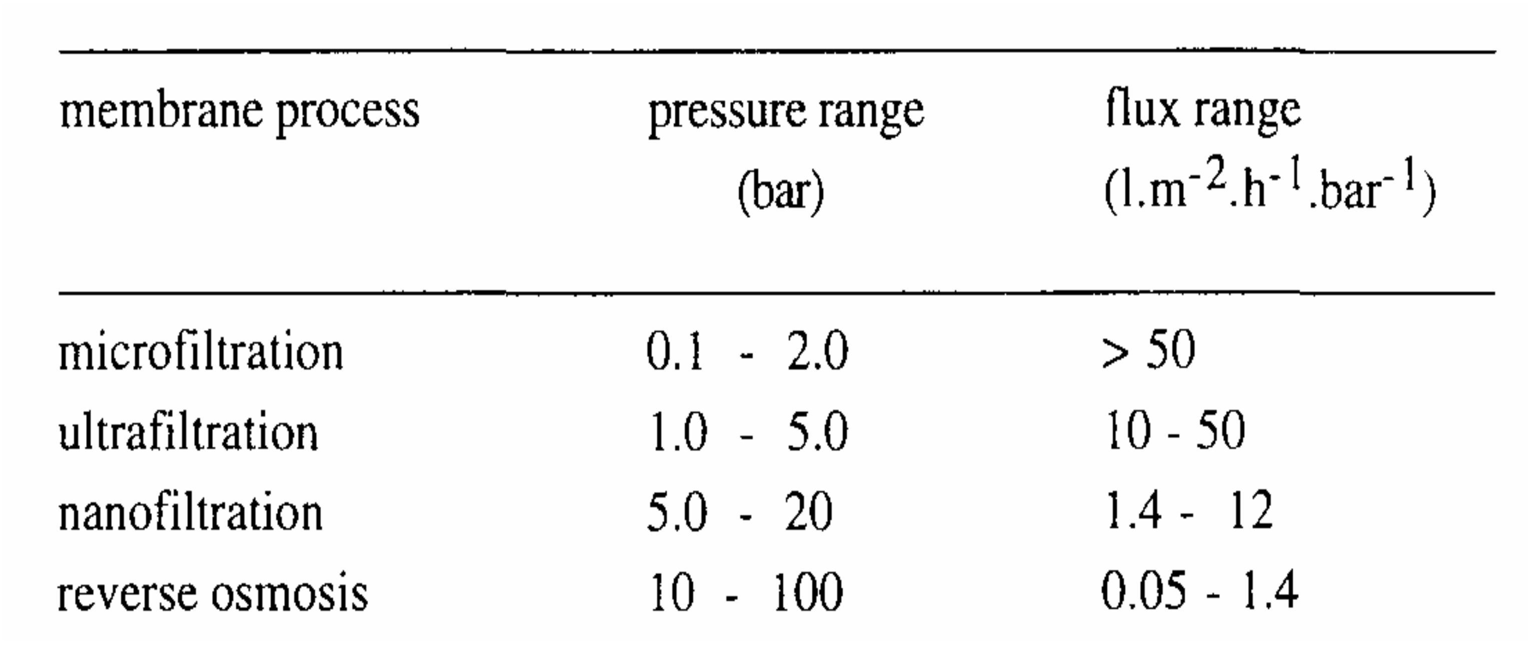
⇒ In pressure-driven separations, which include microfiltration, ultrafiltration, nanofiltration and reverse osmosis, solvent and small solutes pass through while large solutes are retained by membrane.

Types and characteristics of pressure-driven separations



Types and characteristics of pressure-driven separations

- Pressure-driven separations are defined by the **type of membrane they employ,** which determines the **size of solute they separate, their operating conditions and productivity.**



membrane process	pressure range (bar)	flux range ($l.m^{-2}.h^{-1}.bar^{-1}$)
microfiltration	0.1 - 2.0	> 50
ultrafiltration	1.0 - 5.0	10 - 50
nanofiltration	5.0 - 20	1.4 - 12
reverse osmosis	10 - 100	0.05 - 1.4

Types and characteristics of pressure-driven separations

- The separation in pressure-driven processes is characterized by the **retention (*R*)**.

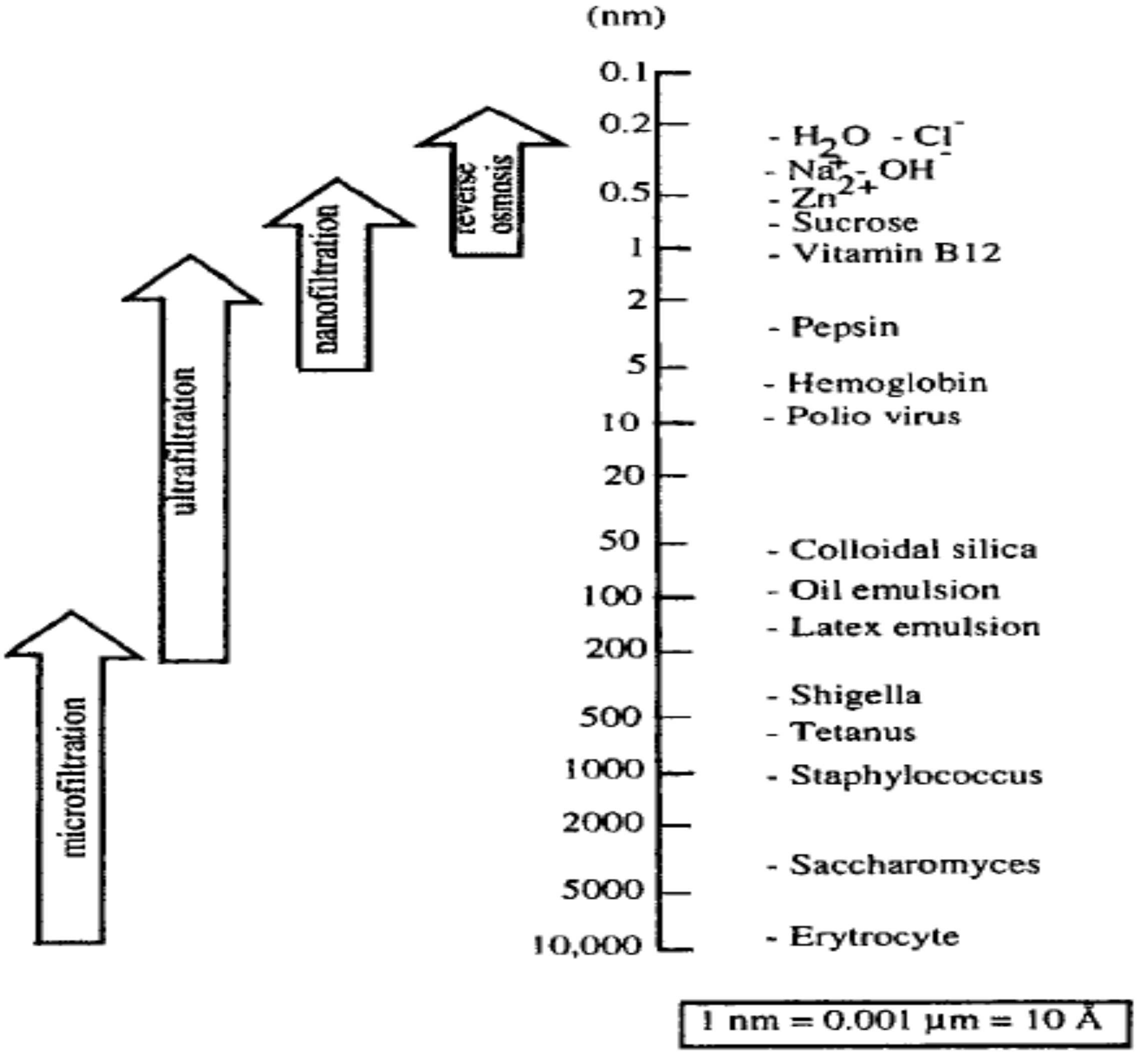
$$R = \frac{c_f - c_p}{c_f}$$

where: C_f and C_p are the solute concentrations in feed and permeate, respectively

- Typically, $C_p << C_f$ so that the flux of solute (J_s) is much less than the flux of solvent (J_w) and

$$J_{total} = J = J_s + J_w \approx J_w$$

Molecule size vs. recommended pressure driven process



Microfiltration and Ultrafiltration

- Microfiltration membranes have pores in the range of 0.05 – 10 µm porous while ultrafiltration membranes have pores in the range of 1 – 100 nm.
- Separation is based on a simple **sieving mechanism – which is analogous to classical filtration;**
- interactions of solute and solvent with membrane are negligible.
- The transport of solvent in microfiltration and ultrafiltration by appropriate equation for a steadystate **viscous flow.**

Microfiltration and Ultrafiltration

- Hagen-Poiseuille equation for cylindrical pores:

$$J = -\frac{\varepsilon r^2}{8\eta\tau}\frac{dp}{dx}$$

- Kozeny-Carman equation for the gaps between spheres:

$$J = -\frac{\varepsilon^3 d^2}{180\eta(1-\varepsilon)^2}\frac{dp}{dx}$$

- Where: ε is the porosity, r is the pore radius, τ is the tortousity, η is the viscosity of solvent, d is the diameter of spherical particle, dp/dx is a differential pressure drop.

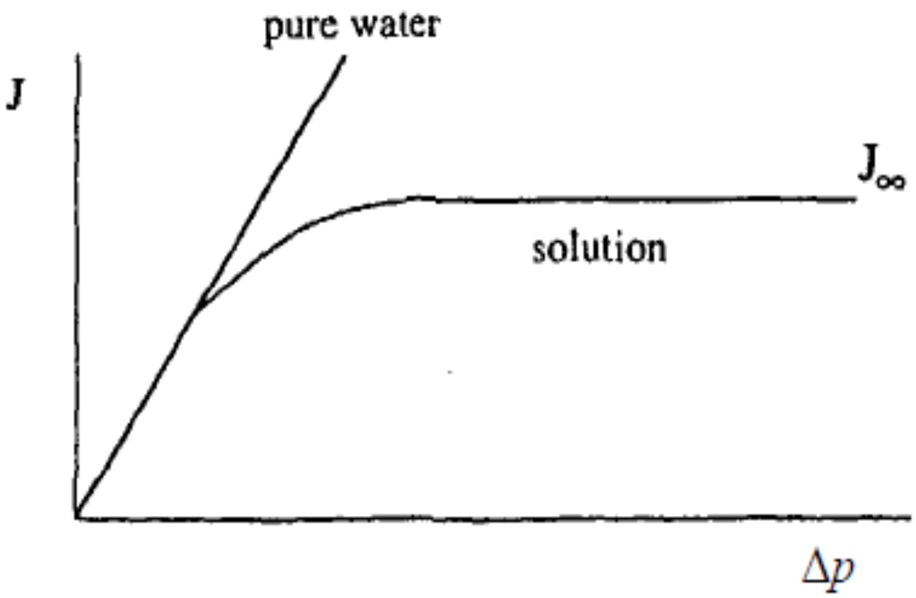
Microfiltration and Ultrafiltration

- At the same time the flux of solvent *J* may be described by **Darcy's law:** $J = A\Delta p$

where: *A* is the permeability constant and *Δp* is the pressure drop across the membrane.

- The **permeability constant,** *A* is inversely proportional to the viscosity of solvent and membrane thickness; it strongly depends on structural factors such as the porosity, pore size, pore distribution, and pore geometry.
- Darcy's law, which predicts direct proportionality between the flux and the applied pressure (pressure drop across the membrane), would be applicable if the feed did not contain solute. The presence of solute causes a deviation from Darcy's law.

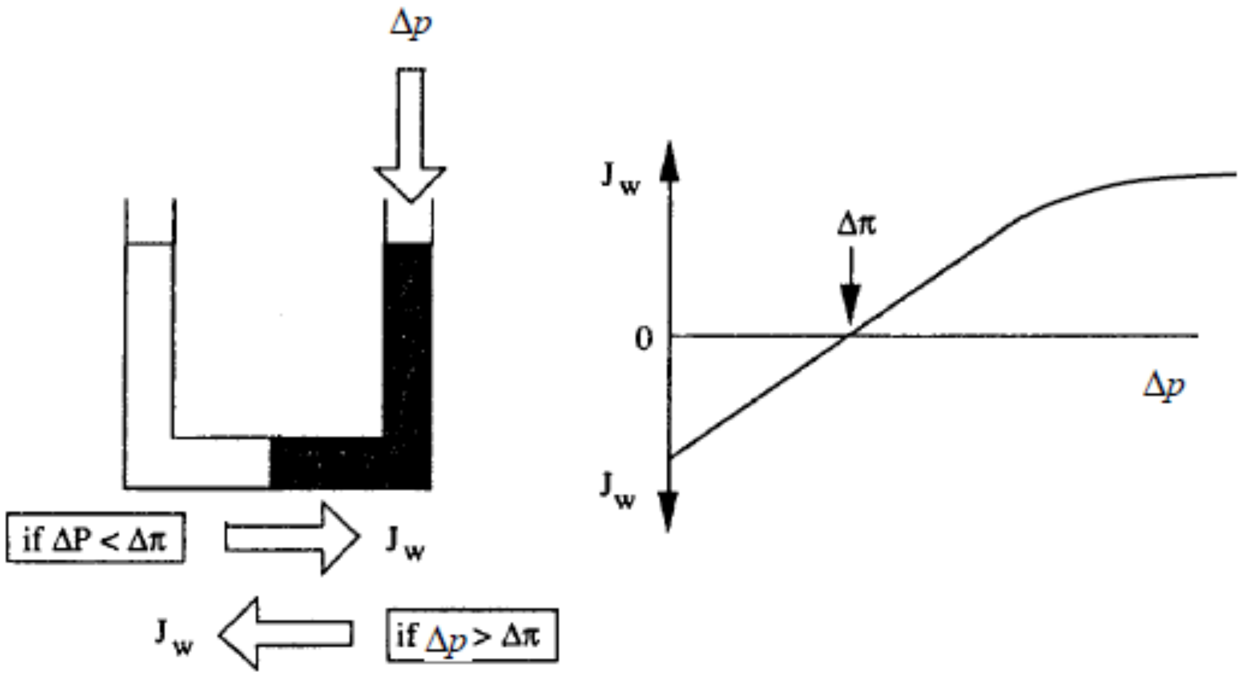
Microfiltration and Ultrafiltration



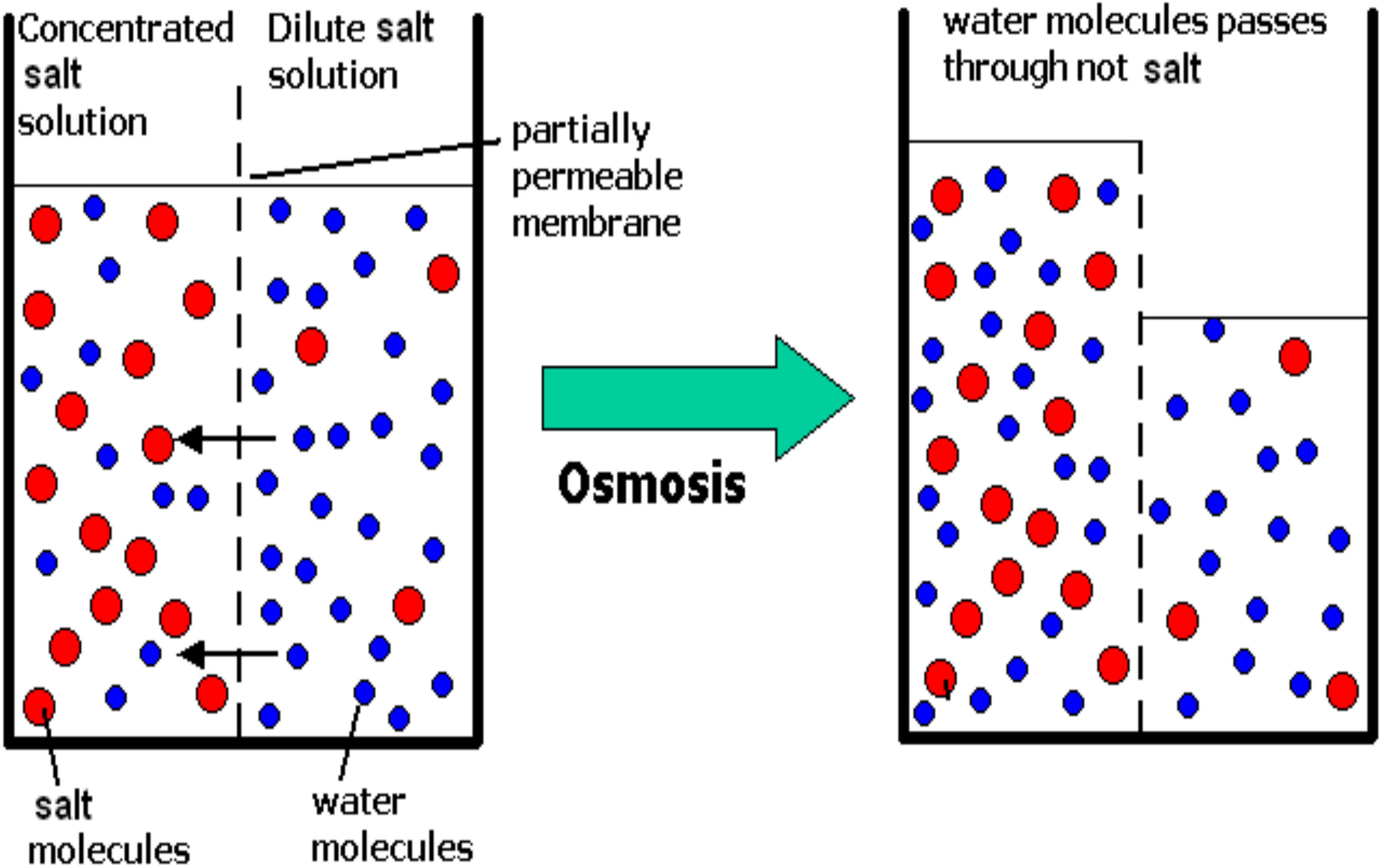
- The existence of a limiting flux (J_∞) as result of **boundary layer phenomena, such as concentration polarization and gel layer formation.**

Reverse nanofiltration and reverse osmosis

- Nanofiltration and reverse osmosis membranes have pore size less than 2 nm, but the transport through these membranes is usually described by the **solution-diffusion mechanism.**
- The driving force nanofiltration and reverse osmosis must be corrected for the **osmotic pressure.**



Osmosis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partially permeable membrane
salt molecules
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Osmosis
water molecules passes through not salt

Osmotic pressure

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Osmotic pressure

Or

$$\pi = R(T + 273)\sum m_i$$

$$R = 0.082 \ \ \mathrm{L.bar/mol.^{o}K}$$

$$\mathrm{T} = \mathrm{temperature}\ {}^{o}C$$

$$\sum \ \ = sum$$

$m_i = molar$ concentration of all ions.

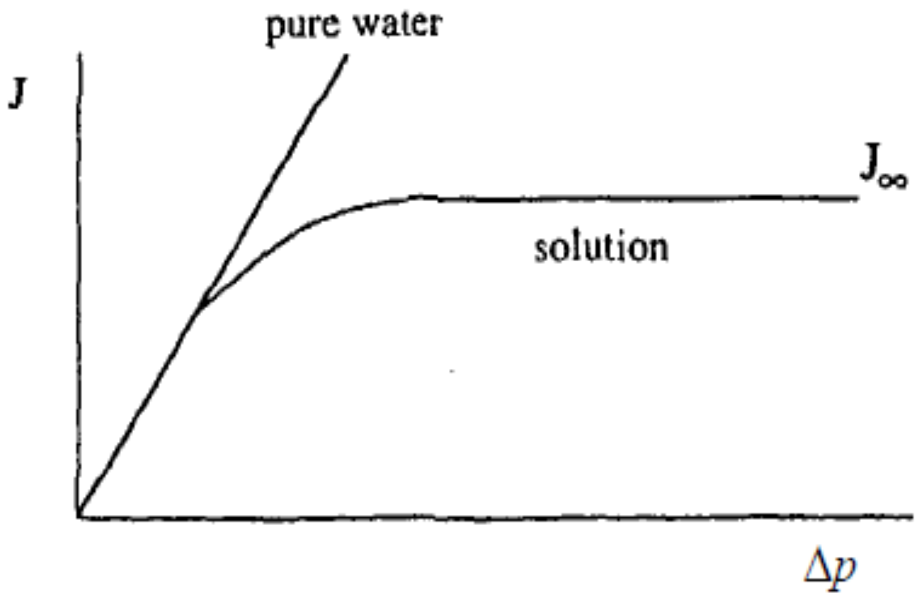
Experimental values of osmotic pressure

Sea Salt Solutions	
Wt% Salts	**Osmotic Pressure (atm)**
0	0
1.00	7.10
3.45*	25.02
7.5	58.43
10	82.12

* Value of standard seawater

Reverse nanofiltration and reverse osmosis

- Similarly to microfiltration and ultrafiltration, nanofiltration and reverse osmosis are also subject of boundary layer phenomena and above certain Δp, J_w is no longer directly proportional to Δp.



Reverse nanofiltration and reverse osmosis

- **Solvent flux in nanofiltration and reverse osmosis processes:**

Neglecting boundary layer phenomena, the flux of solvent is given by,

$$J_w = P_w(\Delta p - \Delta \pi)$$

where: P_w is the permeability coefficient of solvent and $\Delta\pi$ is the osmotic pressure.

Reverse nanofiltration and reverse osmosis

- Since the transport in nanofiltration and reverse osmosis membrane is described by the solution-diffusion mechanism, P_w (the permeability coefficient of solvent) is a property of the membrane material and is defined as,

$$P_w = \frac{D_w c_w V_w}{RTl_m}$$

where: D_w is the diffusion coefficient of solvent in membrane c_w is the solubility (distribution) coefficient of solvent in membrane, V_w is the molar volume of solvent, l_m is the membrane thickness, T is the absolute temperature, and R is gas constant.

Reverse nanofiltration and reverse osmosis

- In addition to solvent, solute may also be transported through the membrane and its flux (J_s) *is* given by,

$$J_s = P_s \Delta c_s = \frac{D_s K_s}{l_m}\left(c_f - c_p\right)$$

Where: P_s , D_s, and K_s are the permeability, diffusion and solubility (distribution) coefficients of solute in membrane, and c_f and c_p are the concentrations feed and permeate, respectively.

- Alternatively, the transport of solvent and solute in membrane may be described by different pore models, in case of which the physical interpretation of P_w and P_s becomes different.

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Reverse Osmosis

Dr.Mohammed Rasool Qtaishat

Reverse Osmosis

- In reverse osmosis, the solute rejection R is defined as the ratio concentration difference across the membrane divided by the bulk concentration on the feed or concentrate side (fraction of solute remaining in the feed stream).

$$R = \frac{c_f - c_p}{c_f} = 1 - \frac{c_p}{c_f}$$

Reverse Osmosis

$$R = \frac{B(\Delta P - \Delta \pi)}{1 + B(\Delta P - \Delta \pi)}$$

$$B = \frac{P_w}{D_s K_s c_{w2}} = \frac{A_w}{A_s c_{w2}}$$

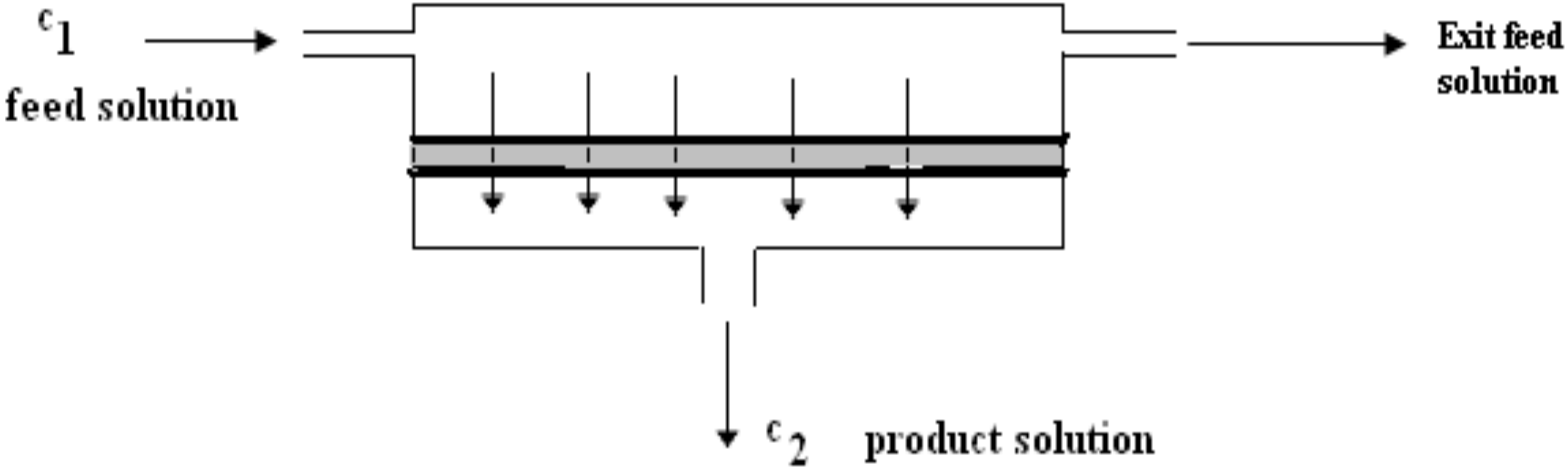
- Where B is in atm^{-1}. Note that B is composed of the various physical properties P_w, D_s, and K_s of the membrane and must be determined experimentally for each membrane.
- Usually, the product $D_s K_s$ is determined, not the value of D_s and K_s separately. Also, many of the data reported in the literature give values of (P_w/L_w) or A_w in kg solvent /s.m^2.atm and $(D_s K_s / l_m)$ or A_s in m/s and not separate values of l_m , P_w , and so on.

Reverse Osmosis

Example: ***Experimental Determination of Membrane Permeability.***

Experiment at 25 C^o were performed to determine the permeabilities of a cellulose acetate membrane. The laboratory test section shown in the figure has membrane area A=2x10^{-3} m^2. The inlet feed solution concentration of NaCl is c_1=10 kg NaCl/m^3 solution (10 g NaCl/L, ρ_1=1004 kg solution /m^3). The water recovery is assumed low so that the concentration c_1 in the entering feed solution flowing past the membrane and the concentration of the exit feed solution are essentially equal. The product solution contains c_2= 0.39 kg NaCl/m^3 solution (ρ_2=997 kg solution /m^3)and its measured flow rate is 1.92x10^{-8} m^3 solution /s. A pressure differential of 5514 kPa (54.42 atm) is used. Calculate the permeability constants of the membrane and the solute rejection R.

Reverse Osmosis



Reverse Osmosis

Solution: since c_2 is very low (dilute solution), the value c_{w2} can be assumed as the density of water or c_{w2}=997 kg soluvent/m^3. To convert the product flow rate to water flux, J_w, using an area of $2x10^{-3}m^2$.

$$J_w = (1.92\times10^{-8} m^3/s)(997 kg \text{ solvent /m}^3)/(2.00\times10^{-3} m^2)$$

$$= 9.57\times10^{-3} \text{ kg solvent /s.m}^2$$

$$J_s = \frac{J_w c_2}{c_{w2}} = \frac{(9.57\times10^{-3})(0.39)}{997}$$

$$= 3.744\times10^{-6} kg\ solute \text{ NaCl/s.m}^2$$

Reverse Osmosis

- To determine the osmotic pressure from a reference table. The concentration are converted as follows. For c_1, 10 kg NaCl is in 1004 kg solution /m3 (ρ_1=1004). Then 1004-10=994 kg H_2O in 1 m^3 solution. Hence, in feed solution where the molecular weight of NaCl =58.45 , (10x1000)/(994x58.45)=0.1721g mol NaCl /kg H_2O. From a reference table the predicted π_1=7.8 atm by linear interpolation. Substituting the predicted π_1=8.39 atm, which is higher than the experimental value. For the product solution, 997-0.39=996.6 kg H_2O. Hence, (0.39x1000)/(996.6x58.45)=0.00670 g mol/kg H_2O. From the same table π_2=0.32 atm.

Then $\Delta\pi = \pi_1 - \pi_2$=7.8-0.32=7.48atm and ΔP=54.42 atm.

Reverse Osmosis

$$J_w = 9.57\times10^{-3} = \frac{P_w}{L_m}(\Delta p - \Delta\pi) = \frac{P_w}{L_m}(54.42 - 7.48)$$

$$\text{solving } \left({}^{P_w}\!/_{L_m}\right) = A_w = 2.039\times10^{-4}\, kg \text{ solvent/s.m}^2 .atm$$

$$N_s = 3.744\times10^{-6} = \frac{\mathrm{D_s}K_s}{L_m}(c_1 - c_2) = \frac{D_s K_s}{L_m}(10.00 - 0.39)$$

$$solving\ (\mathrm{D_s}K_s / L_m) = A_s = 3.896\times10^{-7}\, m/s.$$

Reverse Osmosis

$$R = \frac{c_1 - c_2}{c_1} = \frac{10 - 0.93}{10} = 0.961$$

$$B = \frac{P_w / L_w}{(D_s K_s / L_m) c_{w2}} = \frac{2.039 \times 10^{-4}}{(3.896 \times 10^{-7}) 997} = 0.5249 atm^{-1}$$

$$R = \frac{B(\Delta P - \Delta \pi)}{1 + B(\Delta P - \Delta \pi)} = \frac{0.5249(54.42 - 7.48)}{1 + 0.5249(54.42 - 7.48)} = 0.961$$

Applications, Equipment, and Models for Reverse Osmosis

- **Effects of operating variables:**

- In many commercial units operating pressures in reverse osmosis range from about 1035 up to 10 350 kPa (150 up to 1500 psi).
- As the feed pressure is increased, solvent or water flow through the membrane increases and the solute flow remains constant, giving lower solute concentration in the product solution.
- At constant applied pressure, increasing the feed solute concentration increases product solute concentration. Also the amount of solute present in the product solution increases because of the higher feed concentration.

Applications, Equipment, and Models for Reverse Osmosis

- If a reverse osmosis unit has a large membrane area and the path between the feed inlet and outlet is long, the outlet feed concentration can be considerably higher than the inlet feed.

- Many manufacturers use the feed solute or salt concentration average between inlet and outlet to calculate the solut or salt rejection R.

Applications, Equipment, and Models for Reverse Osmosis

Example: *Prediction of performance in a reverse osmosis unit.*

A reverse osmosis membrane to be used at 25°C for a NaCl feed solution containing 2.5g NaCl /L (2.5Kg NaCl/m^3, ρ=999kg/m^3) has a water permeability constant $A_w=4.81\times10^{-4}$ kg/s.m^2.atm and a solute (NaCl) permeability constant $A_s=4.42\times10^{-7}$m/s. Calculate the water flux and solute flux through the membrane using a ΔP =27.20 atm and the solute rejection R. Also calculate c_2 of the product solution.

Applications, Equipment, and Models for Reverse Osmosis

Solution:

In the feed solution, c_1=2.5 kg NaCl/m^3 and ρ=999kg solution/m^3 . Hence, for the feed, 999-2.5 =996.5 kg H_2O in 1.0 m^3 solution; also for the feed, (2.50x1000)/(996.5x58.45)=0.04292 g mol NaCl/kg H_2O. π_1=1.97atm. Since the product solution c_2 is unknown, a value of c_2=0.1 kg NaCl/m^3 and C_{w2} =997 kg solvent/m^3.

Then for the product solution, (0.1x1000) /(996.9x58.45) =0.00172 g mol NaCl/kg H_2O and π_2=0.08atm. Also, $\Delta\pi = \pi_1 - \pi_2$=1.97-0.08=1.89atm.

Applications, Equipment, and Models for Reverse Osmosis

$$J_w = A_w(\Delta P - \Delta\pi) = 4.81\times10^{-4}(27.20-1.89)$$

$$=1.217\times10^{-2}\, kgH_2O/s.m^2$$

For calculating R :

$$\mathrm{B} = \frac{\mathrm{A_w}}{\mathrm{A_s} c_{w2}} = \frac{4.81\times10^{-4}}{4.42\times10^{-7}\times997} = 1.092 atm^{-1}$$

$$R = \frac{B(\Delta P - \Delta\pi)}{1+B(\Delta P-\Delta\pi)} = \frac{1.092(27.20-1.89)}{1+1.092(27.20-1.89)} = 0.965$$

$$R = 0.965 = \frac{c_1 - c_2}{c_1} = \frac{2.50 - c_2}{2.50}, \qquad c_2 = 0.0875 kg\ \mathrm{NaCl/m^3}$$

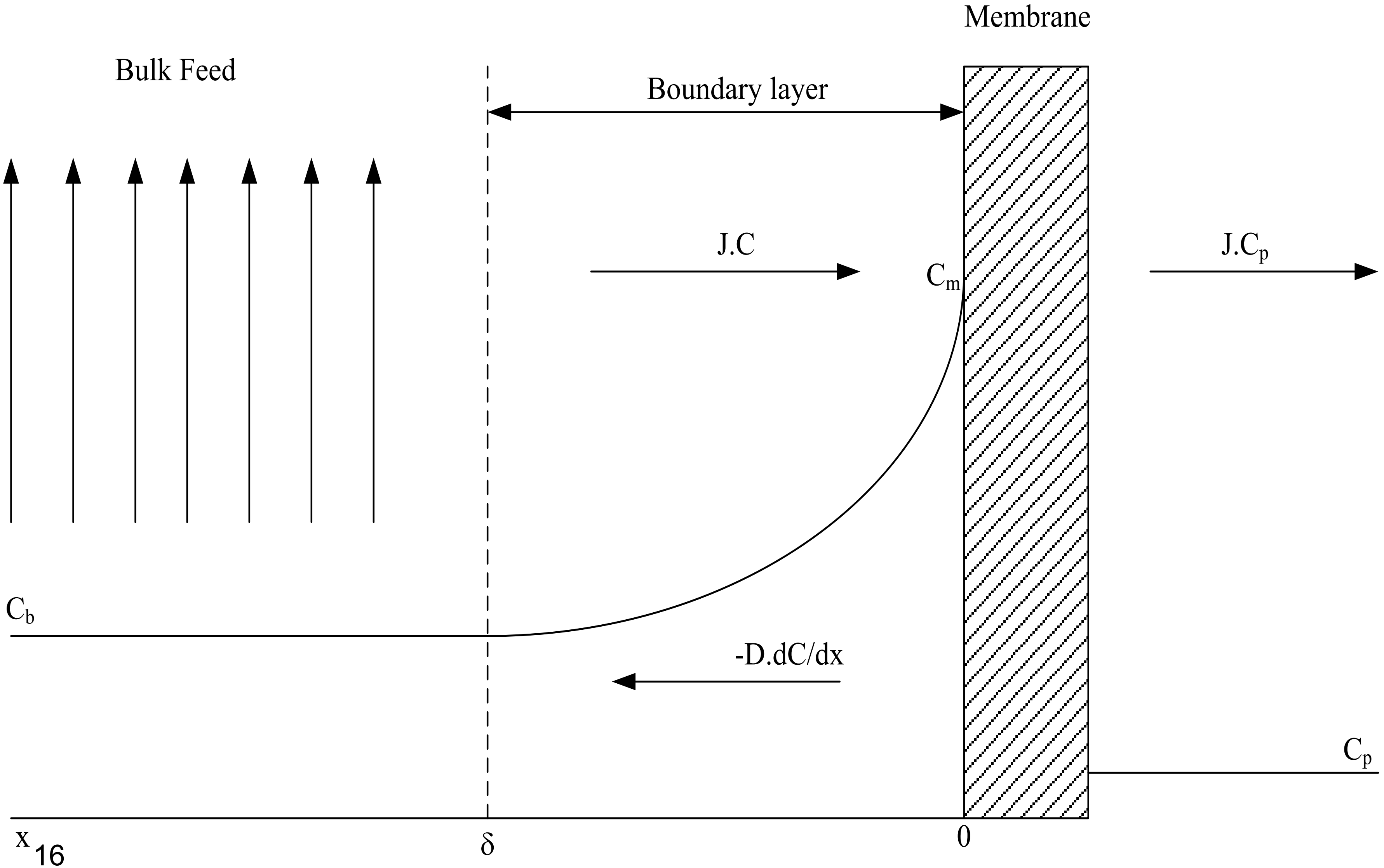
$$N_s = A_s(c_1 - c_2) = 4.42\times10^{-7}(2.5-0.0875)$$

$$=1.066\times10^{-6} kg\ \mathrm{NaCl/s.m^2}$$

Concentration polarization in reverse osmosis

- In desalination localized concentrations of solute build up at the point where the solvent leaves the solution and enters the membrane. The solute accumulates in a relatively stable boundary layer next to the membrane. Concentration polarization,β,is defined as the ratio of the salt concentration at the membrane surface to the salt concentration in the bulk feed stream c_1.
- Concentration polarization causes the water flux to decrease since the osmotic pressure π_1 increases as the boundary layer concentration increases and the overall driving force ($\Delta P-\Delta\pi$) decreases. Hence,often the ΔP must be increased to compensate which gives higher power costs.

Concentration polarization in reverse osmosis



Concentration polarization in reverse osmosis

- The effect of the concentration polarization β can be included approximately by modifying the value of Δπ in the equation $N_w = \frac{P_w}{L_m}(\Delta P - \Delta\pi)$ as follows:

$$\Delta\pi = \beta\pi_1 - \pi_2$$

- It is assumed that the osmotic pressure π_1 is directly proportional to the concentration, which is approximately correct so equation $N_s = \frac{D_s K_s}{L_m}(c_1 - c_2)$ can be modified as:

$$N_s = A_s(\beta c_1 - c_2)$$

Concentration polarization in reverse osmosis

- The usual concentration polarization ratio is 1.2 to 2.0,i.e., the concentration in the boundary layer is 1.2 to 2.0 times c_1 in the bulk feed solution. This ratio is often difficult to predict. In desalination of sea water using values of about 1000 psia= ΔP , π_1 can be large. Increasing this π_1 by a factor of 1.2 to 2.0 can appreciably reduce the solvent flux. For brackish waters containing 2 to 10 g/L and using ΔP values of 17 to 55 atm abs, the value of π_1 is low and concentration polarization is not important.
- The boundary layer can be reduced by increasing the turbulence using higher feed solution velocities. However, this extra flow results in a smaller ratio product solution to the feed. Also, screens can be put in the flow path to introduce turbulence.

Permeability Constants of Reverse Osmosis Membranes

- Permeability constants for membranes must be determined experimentally for the particular type of membrane to be used. For cellulose acetate membranes, typical water permeability constant A_w range from about 1×10^{-4} to 5×10^{-4} kg solvent/s.m^2.atm. Values for other types of membranes can differ widely.

- Generally, the water permeability constant for a particular membrane does not depend upon the solute present. For the solute permeability constants A_s of cellulose acetate membranes, some relative typical values are as follows:

Permeability Constants of Reverse Osmosis Membranes

Solutes in cellulose acetate membranes	A_s (m/s)
NaCl	4.0×10^{-7}
$BaCl_2$	1.6×10^{-7}
$MgCl_2$	2.2×10^{-7}
$CaCl_2$	2.4×10^{-7}
Na_2SO_4	4×10^{-7}
KCl	6×10^{-7}
NH_4Cl	6×10^{-7}

Types of Equipment for Reverse Osmosis

- **Types of Equipment for membrane processes:**

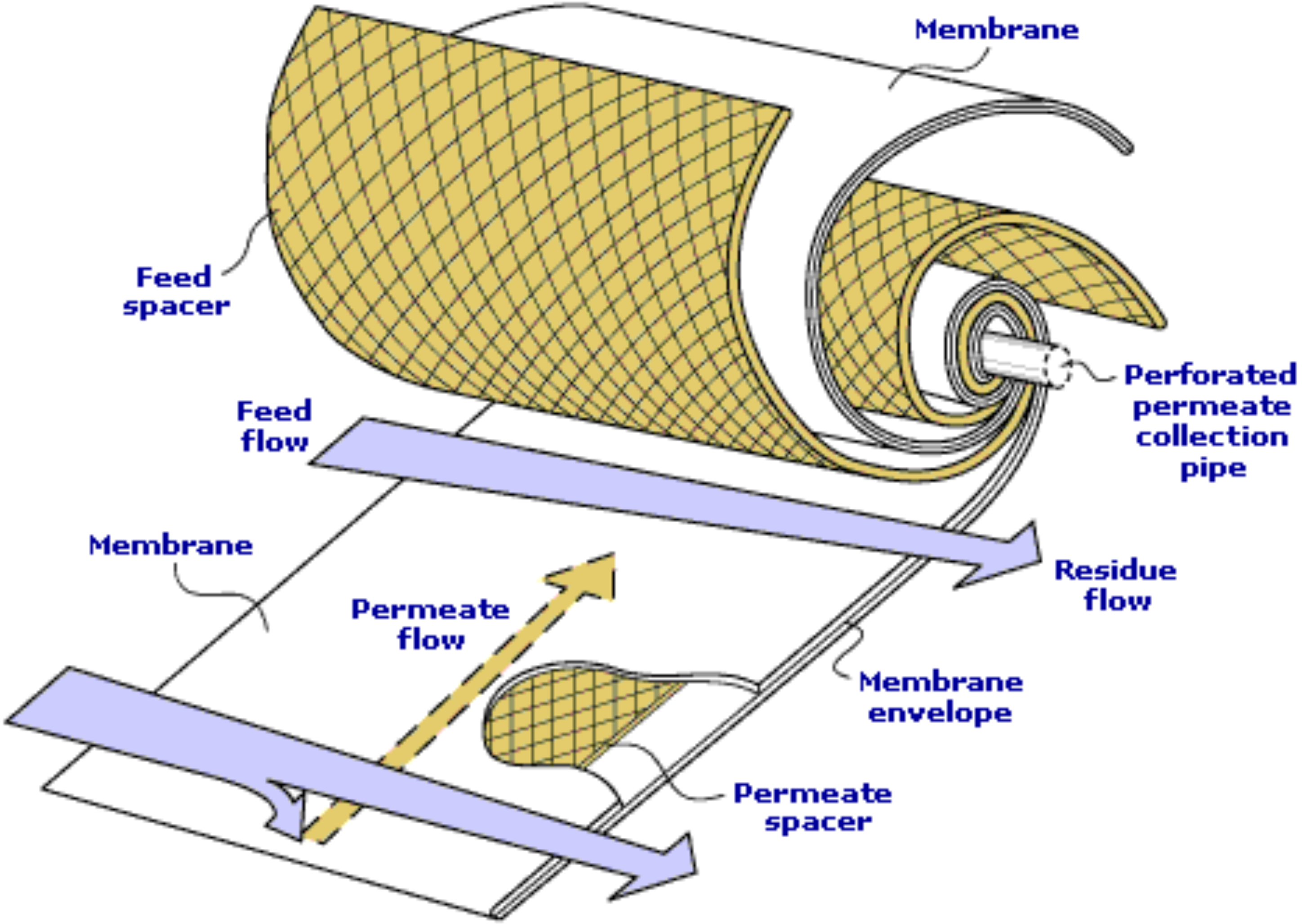
1. ***Flat membranes:*** Flat membranes are mainly used for experimental use to characterize the permeability of the membrane. The modules are easy to fabricate and use and the areas of the membranes are well defined.In some cases modules are stacked togther like a multilayer sandwich or plate-and –frame filter press. The major drawback of this type is the very small membrane area per unit separator volume. Small commercial flat membranes are used for producing oxygen-enriched air for individual medical applications.

Types of Equipment for Reverse Osmosis

2. ***Spiral-wound membranes:***

This configuration maintains the simplicity of fabricating flat membranes while increasing markedly the membrane area per unit separator volume up to 100 ft^2/ft^3 (328 m^2/m^3)while decreasing pressure drops. The assembly consists of a sandwich of four sheets wrapped around a central core of a perforated collecting tube. The four sheets consists of a top sheet of an open separator grid for the feed channel, a membrane, a porous felt backing for the permeate channel, and another membrane as shown in the figure:

Types of Equipment for Reverse Osmosis



Types of Equipment for Reverse Osmosis

- The spiral wound element is 100 mm to 200 mm in diameter and is about 1 to 1.5 m by about 2 to 2.5 m. The space between the membranes (open grid for feed) is about 1 mm and the thickness of the porous backing (for permeate) is about 0.2 mm.
- The whole spiral wound element is located inside a metal shell. The feed gas enters at the left and of end of the shell, enters the feed channel, and flows through this channel axial direction of the spiral to the right end of the assembly. Then the exit residue gas leaves the shell at this point. The feed stream, which is in the feed channel, permeates perpendicularly through the membrane.

Types of Equipment for Reverse Osmosis

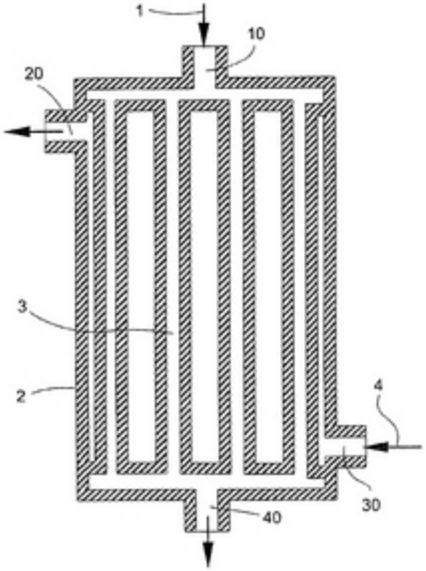
- This permeate then flows through the permeate channel in a direction perpendicular to the feed stream toward the perforated collecting tube, where it leaves the apparatus at one end.

3. ***Hollow fiber membranes:***

The membranes are in the shape of very small diameter hollow fibers. The inside diameter of the fibers is in the range 100 to 500 μm and the outside 200 to 1000 μm with the length up to 3 to 5 m. The module resembles a shell and tube heat exchanger. Thousands of fine tubes are bound together at each end into a tube sheet is surrounded by a metal shell having a diameter of 0.1 to 0.2 m, so that the membrane area per unit volume is up to 10000m^2/m^3.

Types of Equipment for Reverse Osmosis

- Typically, the high pressure feed enters into the shell side at one end and leaves at the other end. The hollow fibers are closed at one end of the tube bundles. The permeate gas inside the fibers flows counter currently to the shell-side flow and is collected in a chamber where the open ends of the fibers terminate. Then the permeate exits the device.



University of Jordan

Faculty of Engineering and Technology

Chemical Engineering Department

Desalination

Dr. Mohammed Rasool Qtaishat

Brakish Water Composition

- Typical analysis of some brakish waters

Constituent (ppm)	Roswell New Mexico	Webster South Dakota	Bahrain
TDS	15,467	1,541	2,196
Cl '	8040	7	1,020
SO_4 ''	1490	765	192
HCO_3 '	205	298	-
F '	-	0.68	-
CO_3 ''	-	-	-
Na^{++}	4970	102	458
Mg^{++}	161	74	67

Brakish Water Composition

Constituent (ppm)	Roswell New Mexico	Webster South Dakota	Bahrain
Ca^{++}	588	193	217
K^{+}	-	9	-
Fe^{++}	-	1.4	-
Mn	-	0.92	-
SiO_2	13	-	-
pH	-	7.62	7.5

Sea Water Alkalinity

- pH of sea water 7.8 - 8.3
- Alkalinity is due to the presence of HCO_3' ,CO_3'',OH', H_2BO_3'.
- HCO_3' and CO3 ' ' (as well as free CO_2 in sea water) are related to the equilibrium of sea water with the CO_2 in the atmosphere.
- A measure of alkalinity is the amount of a strong acid needed to meutralize a sample of sea water.
- Total Alkalinity (TA) =millimole /L may be expressed (or titration alkalinity)

$$TA = \{HCO_3^{-}\} + 2\{CO_3''\} + \{H_2BO_3''\} + \{OH' - H^{+}\}$$

- Normally $\{H_2BO_3'\}$ is very small.

Sea Water Alkalinity

- ✓ **Thus for example**

If A ml acid normally N is consumed to bring down the pH of a saline sample to 8.3.

And

B ml is further consumed to bring the pH of the sample down to 4.5

-Then

$$\text{P alkalinity} = \frac{\text{A} \times \text{N} \times 50 \times 1000}{Sample\ \text{volume}}\ \ ppm\ \text{CaCO}_3\ equivalents$$

and

$$M\ alkalinity = \frac{(A + B) \times N \times 50 \times 1000}{Sample\ \text{volume}}\ \ \text{ppm CaCO}_3\ \text{equivalents}$$

Desalination Terminology

- **Salinity:** gramms of dissolved inorganic matter per kg of sea water, all bromides and iodides being replaced by an equivalent amount of chloride and all carbonates converted to oxides

- **Brakish Water:** when dissolved solids lie between that of potable and sea water (1000 to 10000 ppm)

- **Brine**

- A saline solution with TDS> that of sea water .

- Or a reject streams from branish water operation

Desalination Terminology

- **$CaCO_3$ Equivalents:**
  - To covert from ppm of a given constituent to ppm as $CaCO_3$ multiply by:

$$\frac{50}{\textit{equivalent}\ \text{weight of constituent}}$$

- **Bicarbonate Alkalinity:**
  - HCO_3' ion concentration expressed as ppm $CaCO_3$

$$= ppm\ \mathrm{HCO_3}' \times \frac{50}{61}$$

Desalination Terminology

- **Carbonate Alkalinity:**

$$ppm\ \mathrm{CO_3}'' \times \frac{50}{30}$$

- **Total Alkalinity (T.A):**

- Summation of all alkalinity components expressed as ppm $CaCO_3$

$$= \mathrm{ppm\ CO_3}'' \times 1.67 + ppm\ \mathrm{HCO_3}' \times 0.82 + ppm\ \mathrm{OH}' \times 2.99$$

- **Total Hardness:**

- Sum of Ca, Mg ion concentration expressed as ppm $CaCO_3$

Desalination Terminology

- **Carbonate Hardness (Temporary Hardness):**

- Hardness attributed to bicarbonates of calcium and/or magnesium expressed as ppm $CaCO_3$

- **Noncarbonate Hardness (Permanent Hardness):**

- Hardness due to sulphates, chlorides and nitrates of Ca and Mg (expressed as ppm $CaCO_3$)
- Exists only if total hardness > M alkalinity.

Desalination Terminology

- **Chloranity:**

- Grammes Cl' per kg sea water after all bromides, iodides replaced by an equivalent amount of chloride.
- Practical determination of chlorinity involves titration with $AgNO_3$ using $k_2Cr_2O_7$ as an indicator.
- Normally for sea water

Salinity % = 0.03 + 1.805 x chlorinity %

- **Chlorosity:**

- Grammes Cl /l sea water at 20 C° after all bromides and iodides replaced by chloride

$$pH=- \log(H^{+})$$

Desalination Terminology

- **Saturation pH:**

  pH at which sea water is saturated in a particular scale forming component i.e. when its ionic produce = solubility product

- **Saturation Index:**

  Equals actual pH- saturation pH

  positive saturation index : means supersaturated solution.

  negative saturation index : under saturated solution.

Scale Formation

1. **Definition of Scales:**

- Deposits of minerals which form on solid surfaces of industrial equipment.
- In desalination: deposits on evaporator and condenser tubes, membrane surfaces…etc.
- Soft scales (sludges) are not as harmful as hard scales.

2. **Scaling Problems:**

- **In distillation:**
  - ☐ loss in thermal efficiency
  - ☐ Operational difficulties leading to a reduction in production or even to shut down.

Scale Formation

2. Scaling Problems:

- **In RO and ED processes:**
  - increase in electric or hydraulic power consumption
  - Operational difficulties.

3. Main Scale Components:

- Scale component resulting from sea water are mainly:
  - CaCO3
  - Mg (OH)2
  - $CaSO_4$, $CaSO_4 .\ ½\ H_2O$, $CaSO_4 .2H_2O$

- Sometimes scales are accompanied by iron oxides resulting from corrosion.

Scale Formation

4. Why Do Scales Form:

- Any salt may precipitate when its ionic product exceeds its solubility limit at the prevailing environmental conditions.
- Scale forming compounds are generally sparingly soluble in water and can easily attain their solubility limits under operational desalination conditions (concentration, alkalinity, pH, temperature….etc).
- The main scale forming compounds have inverted solubility i.e. their solubility decreases with temperature, as compared to compounds like NaCl, $MgCl_2$ which have normal solubility, and are extremely soluble.

Scale Formation

4. Why Do Scales Form:

- The solubility limit of any salt is its ionic product at saturation, i.e. if salt concentration is increased so its ionic product exceeds the solubility product, then precipitation occurs.

- Take the possibility of $Mg(OH)_2$ precipitation from sea wate 19% Cl^-, pH= 8, Mg^{++}=1290 ppm as an example. Solubility product in sea water at 25 C° (termed k'_{sp})=5 x 10^{-11}

$$\mathit{ionic}\ \text{product}\ \mathrm{Mg(OH)_2} = \{Mg^{++}\}\{OH^-\}^2 \quad \text{(moles / l)}$$

$$= \left\{\frac{1290}{24\times1000}\right\}\{10^{-6}\}^2$$

$$= 5.4\times10^{-14}$$

precipitation can not occur

Scale Formation

4. Why Do Scales Form:

- The solubility limits of $CaCO_3$, $Mg(OH)_2$ and $CaSO_4$ is greater in sea water than in distilled water, (ion) interaction.
- The solubility limit of $CaCO_3$ and $Mg(OH)_2$ in sea water is a function of salinity (or chlorinity) besides other factors (e.g. temperature, pH, …etc)
- Taking K_{sp} =solubility limit in distilled water at 25 C^o

K'_{sp} = solubility limit in 19 % Cl sea water, pH=8.2, 25 C^o

- Note:

	K_{sp}	K'_{sp}
$CaCO_3$	$0.5 \ X \ 10^{-8}$	$50 \ X \ 10^{-8}$
$Mg(OH)_2$	$1 \ X \ 10^{-11}$	$5 \ X \ 10^{-11}$

Scale Formation

4. Why Do Scales Form:

- ☐ Take an example on the possibility of $CaCO_3$ precipitation at 25 C° from a sea water sample pH= 8, HCO_3'=140ppm and Ca^{++}= 400ppm

- $\{CO_3''\}$ determination is difficult, But we could estimate it from $\{HCO_3'\}$ knowing that $\{HCO_3'\}$ / $\{CO_3''\}$ at pH= 8 is 230

$$\{HCO_3'\} = \frac{140}{61\times1000} = 2.3\times10^{-3}\,g\text{ ion }/l \quad ,\{\mathrm{CO_3}'\} = \frac{2.3\times10^{-3}}{230} = 1\times10^{-5}\text{ g ion/l}$$

$$\text{ionic product} = \frac{400}{40\times1000}\times1\times10^{-5}$$

$$= 10\times10^{-8}(g\text{ ion/l})^2\,\mathit{then}\text{ under these conditions }\mathrm{CaCO_3}\text{ cannot precipitate}$$

Scale Formation

4. Why Do Scales Form:

- Even when the ionic product exceeds the solubility limit i.e. even when supersaturation occurs, precipitation may not occur.
- Scale formation thus requires:
- ionic concentration i.e. ionic product > solubility limit
- provision of a nucleation centre:

e.g. A heating surface
A solid crystal
A rough metal site …etc

- A reasonable contact or residence time.

Scale Formation

5. Factors affecting scaling in general:

☐ Composition of sea water feed

☐ Plant operating conditions
- concentration ratio at various locations.
- temperature and its distribution
- pH
- alkalinity
- existence of nucleation sites: rough or smooth surfaces.
- residence or retention times in specific locations.

Scale Formation

6. Principal scale forming reactions:

$$2HCO_3' \xrightarrow{heat} CO_3'' + H_2O + CO_2$$

$$CO_3'' + Ca^{++} \rightarrow CaCO_3\,ppt$$

$$CO_3'' + H_2O \xrightarrow{heat} 2OH' + CO_2$$

$$Mg^{++} + 2(OH)^- \rightarrow Mg(OH)_2\,ppt$$

$$Ca^{++} + SO_4'' \rightarrow CaSO_4\,ppt$$

Scale Formation

7. Stability diagrams:

- ☐ These are equilibrium solubility charts for scale forming compounds
- ☐ The solubility of $CaCO_3$ is affected by:
- temperature
- pH.
- alkalinity

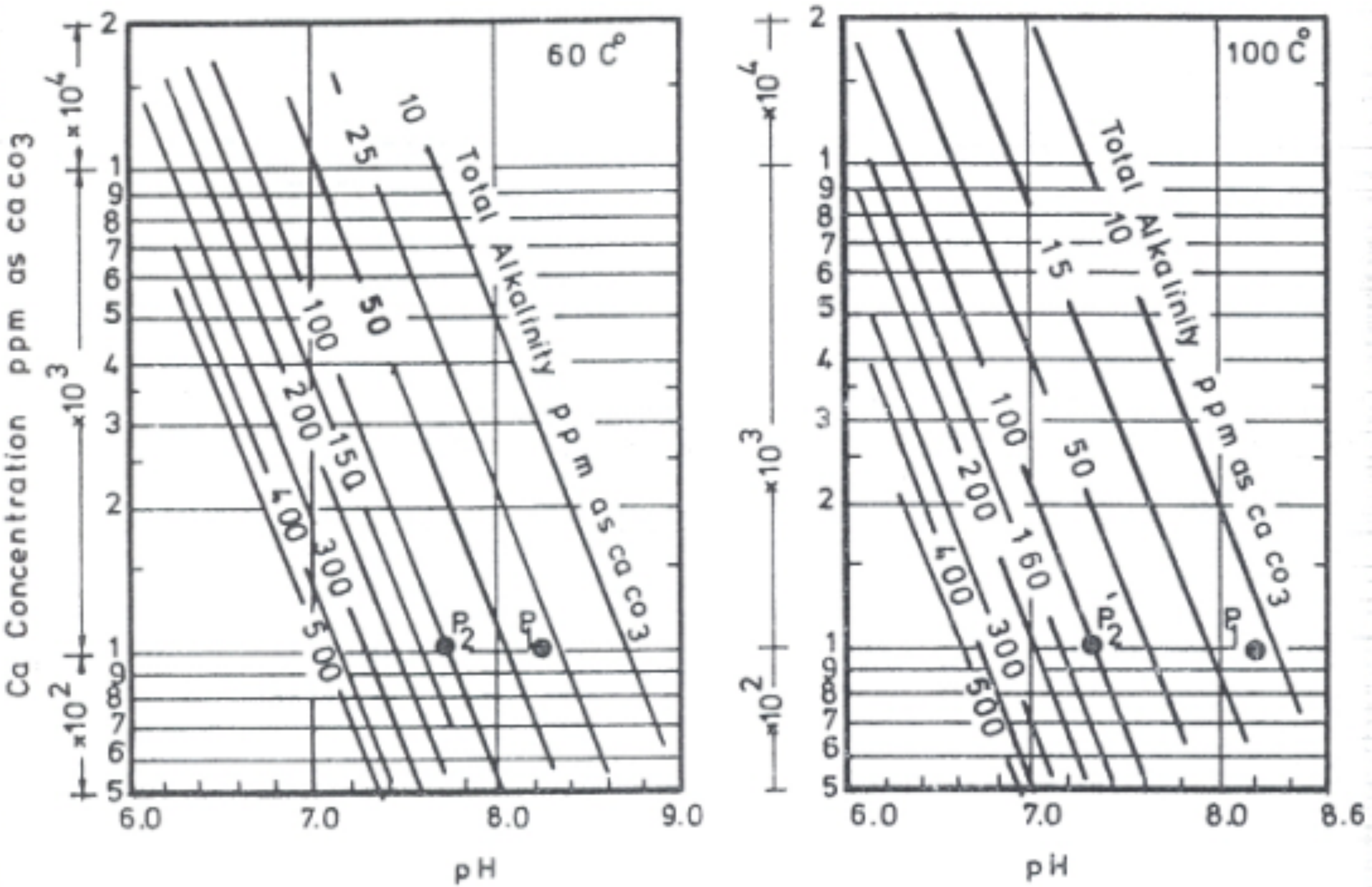
e.g. for sea water solution of a certain alkalinity supersaturation can be obtained by increasing :

Concentration, temperature, or pH.

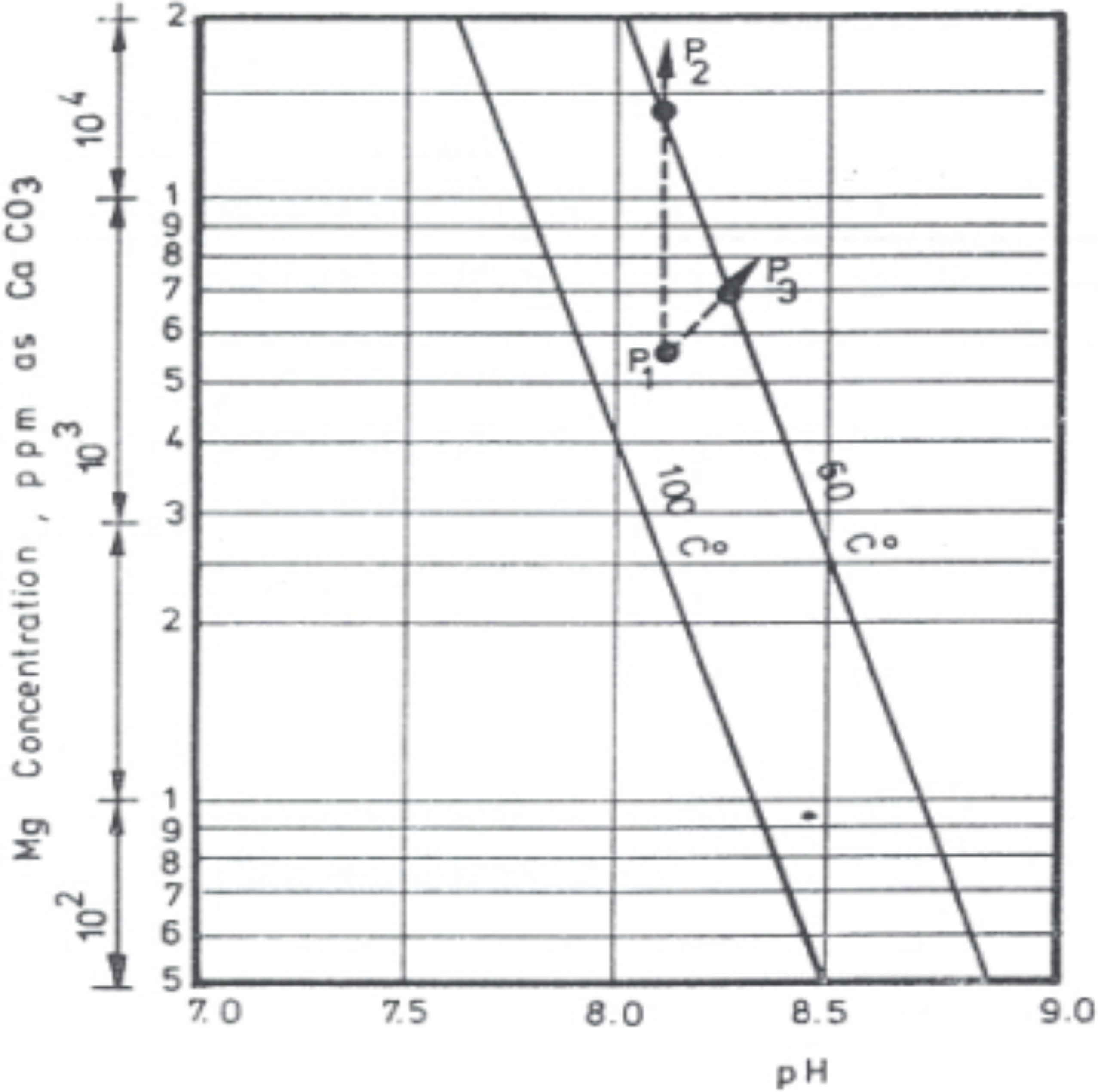
Scale Formation

7. Stability diagrams:

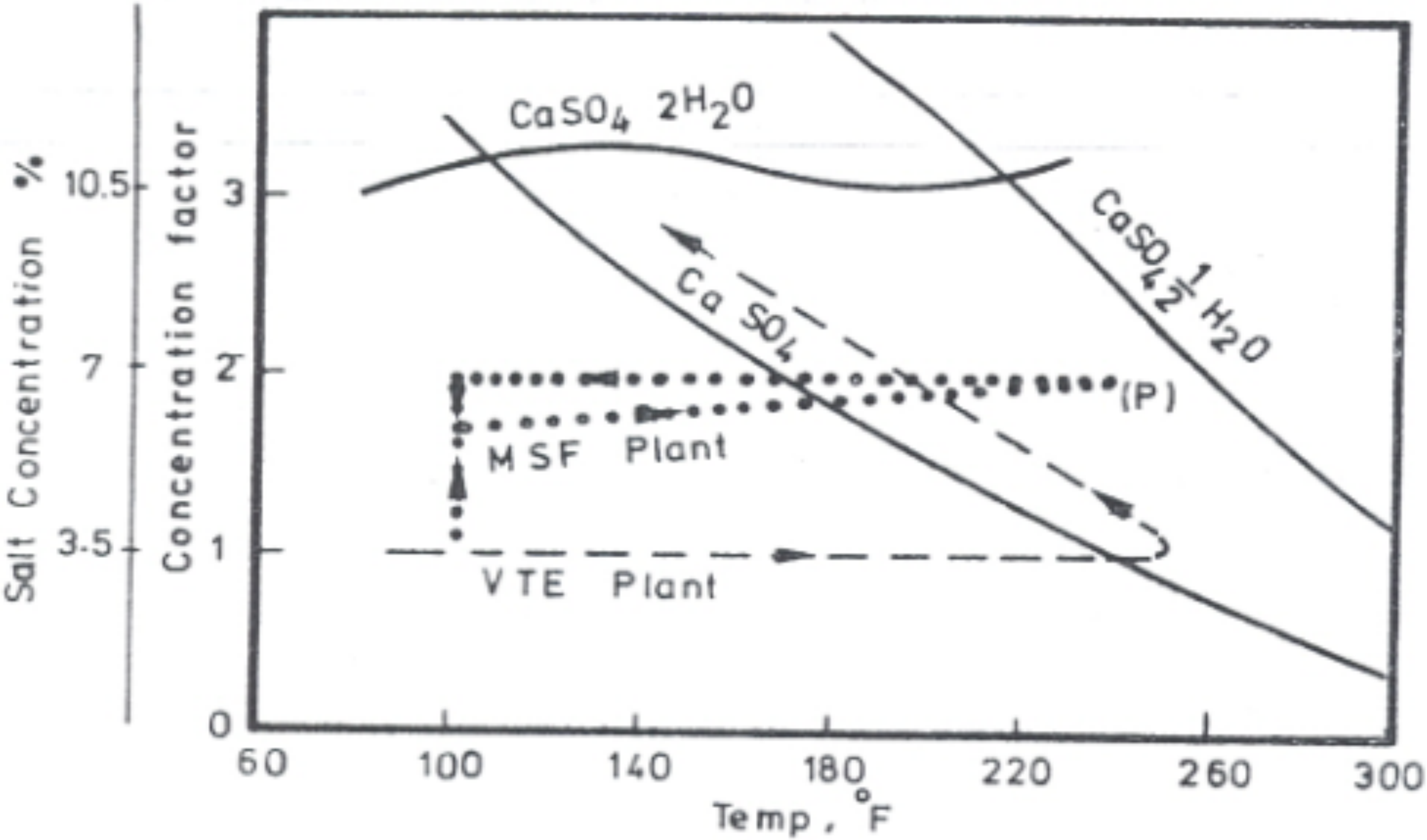
- The solubility of $Mg(OH)_2$ is affected by temperature and pH.
- The solubility of $CaSO_4$ is affected by temperature
- Saturation pH (pH_s) is that at which water is saturated in a certain scale forming compound.
- Saturation index = Actual pH - pH_s



STABILITY DIAGRAM OF CALCIUM CARBONATE IN CONCENTRATED SEA WATER



STABILITY DIAGRAM OF MAGNESIUM HYDROXIDE IN CONCENTRATED SEAWATER



SOLUBILITY LIMITS OF CALCIUM SULPHATE IN SEAWATER

Scale Formation

8. Calcium Sulphate scales:

- Three types - Anhydrite $CaSO_4$
  - Hemihydrate $CaSO_4.1/2\ H_2O$
  - Gypsum $CaSO_4.2\ H_2O$
- On concentrating a sample of sea water
  - ionic product Ca^{++} x SO_4^{-2} increases
  - solubility product also increases with concentration (ionic interaction)
  - super saturation of $CaSO_4$ occurs above a concentration factor of ~ 1.5
  - so scaling with $CaSO_4$ is possible between a concentration factor (C.F.) of 1.5 to 3.
  - above a .C.F. of 3, $CaSO_4.1/2\ H_2O$ scaling is possible.

Scale Formation

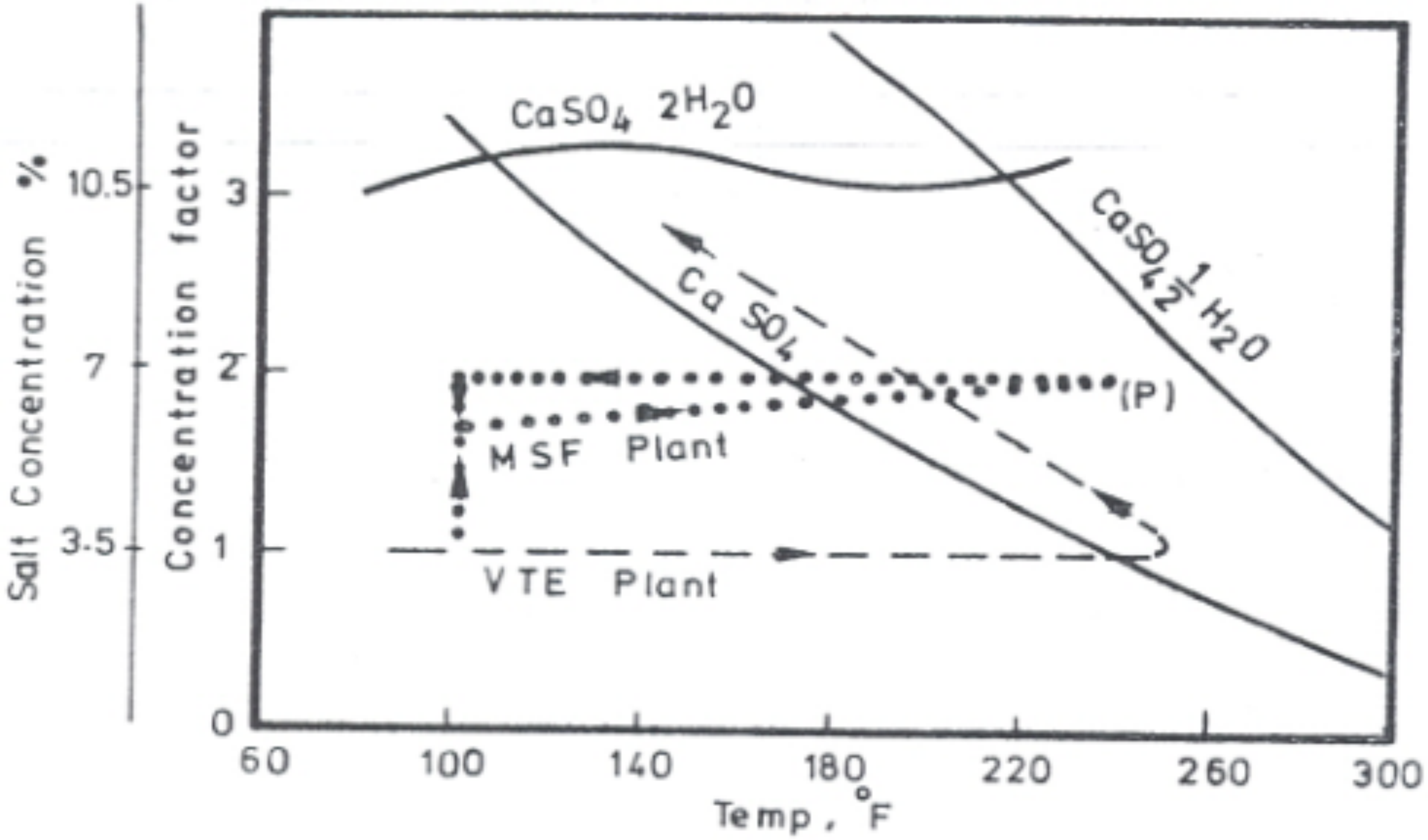
8. Calcium Sulphate scales:

- $CaSO_4$ scaling may not occur spontaneously on supersaturation (due to concentration or temperature): needs considerable residence time.
- $CaSO_4.1/2\ H_2O$ scaling needs shorter residence time.
- Consider calcium sulphate scaling as a function of temperature and concentration:
  - an actual process can operate at some degree of $CaSO_4$ supersaturation.
  - at temperatures and concentrations to the right of $CaSO_4.1/2\ H_2O$ curve, scaling rapidly occur.
  - Gypsum may form at low temperature, and high concentration factors.

Scale Formation

8. Calcium Sulphate scales:

- $CaSO_4$ + $CaSO_4.1/2\ H_2O$: form in high temperature effects
- $CaSO_4.2\ H_2O$ may form in low temperature effects when concentration factor is around 3.
- sulphate are relatively hard and adherent scales.



SOLUBILITY LIMITS OF CALCIUM SULPHATE IN SEAWATER

Scale Formation

9. Alkaline Scales:

- ☐ These are $CaCO_3$ and $Mg(OH)_2$
- ☐ $CaCO_3$ is a soft scale
- ☐ Both are sparingly soluble in sea water.
- ☐ Both scales dissolve in acids
  - Solubility of $CaCO_3$ increases <u>ten fold</u> on changing pH from 8 to 7.
  - increasing CO_2 overpressure increases $CaCO_3$ solubility.
- ☐ $CaCO_3$ starts to precipitate at 77 ^oC (thermal decomposition)

Scale Formation

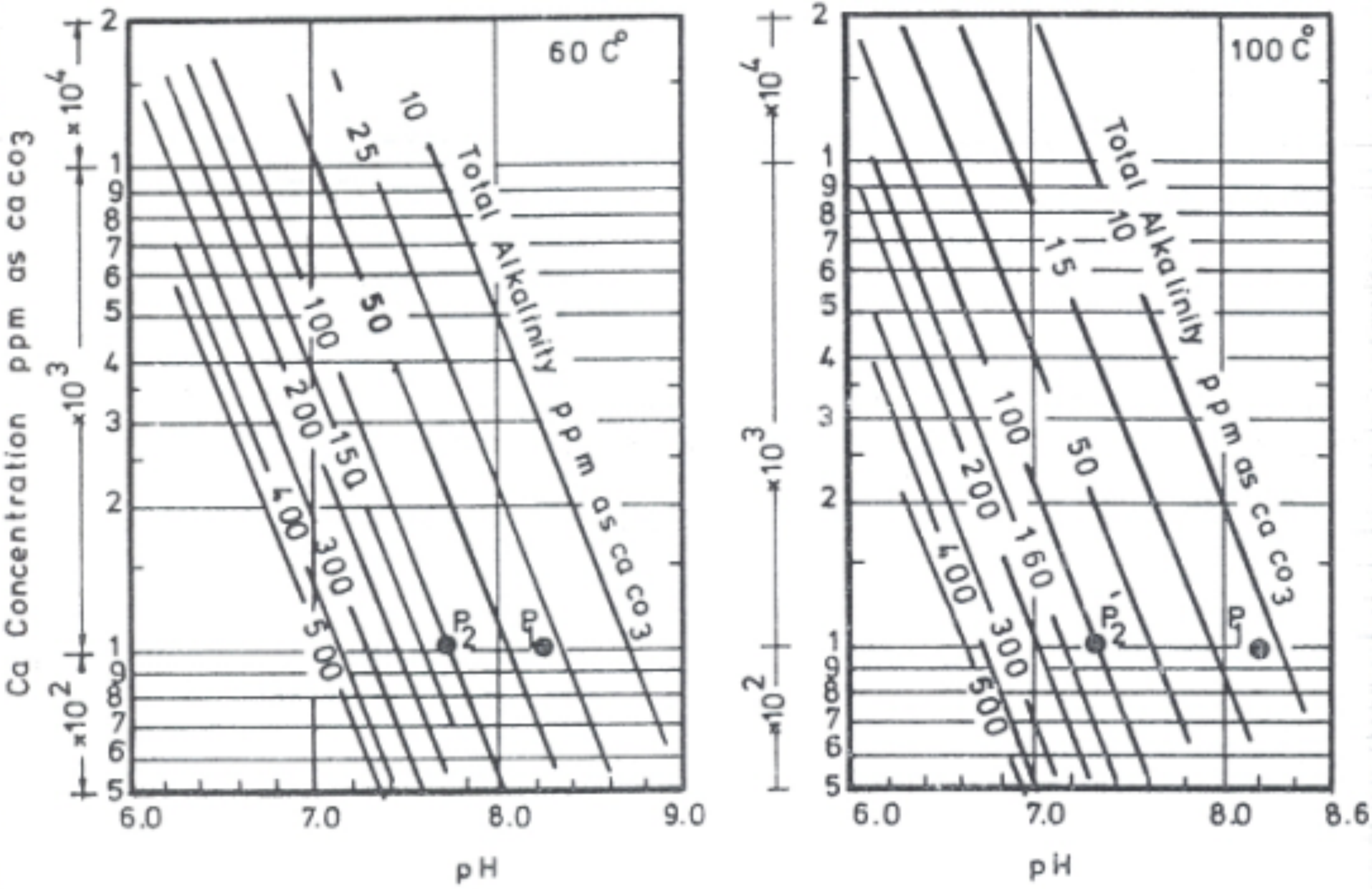
9. Alkaline Scales:

- Given higher temperature and enough time

$$CO_3'' + H_2O \rightarrow 2OH' + 2CO_2$$

- then $Mg(OH)_2$ precipitate above 82 C^o
- However, the quantity of $Mg(OH)_2$ remains small compared to $CaCO_3$ up to about 105 C^o

- Relative rates of formation of $CaCO_3$ and $Mg(OH)_2$ depend on: temperature, residence time, rate of CO_2 release.



STABILITY DIAGRAM OF CALCIUM CARBONATE IN CONCENTRATED SEA WATER

Scale Formation

9. Alkaline Scales:

- ☐ Both temperature and pH encourage formation of $Mg(OH)_2$ scales (stability diagrams).
- ☐ Alkalinity, temperature and pH affects the precipitation of $CaCO_3$ scale (stability diagrams).
  - Precipitation of $CaCO_3$ is related to its solubility product, & $\{Ca^{++}\}$ $\{CO_3''\}$
  - CO_3'' ion concentration is not easy to determine due to the reaction:

$$H^+ + CO_3'' \leftrightarrow HCO_3'$$

  - This reaction would go ← as pH is increased.

Scale Formation

9. Alkaline Scales:

- the equil constant of this reaction

$$K = \frac{\{HCO_3{}'\}}{\{CO_3{}''\}\{H^+\}} = 2.3\times10^{10}$$

$$\text{the ratio } \frac{\{\mathrm{HCO_3}{}'\}}{\{CO_3{}''\}} = 230 \text{ at pH} = 8$$

- then at a certain temperature CO_3'' ion concentration is a function of HCO_3 '' ion concentration and pH.

☐ Estimation of saturation pH and saturation index for $CaCO_3$ in sea water: (stability diagram).

Scale Formation

9. Alkaline Scales:

Assume:

$$Ca^{++} = 2000 \text{ as ppm equivalent CaCO}_3 \left\{400 \times \frac{50}{20} \times 2\right\} (at \text{ C.F} = 2)$$

$$\text{Total alkanity} = 100 \text{ as ppm CaCO}_3$$

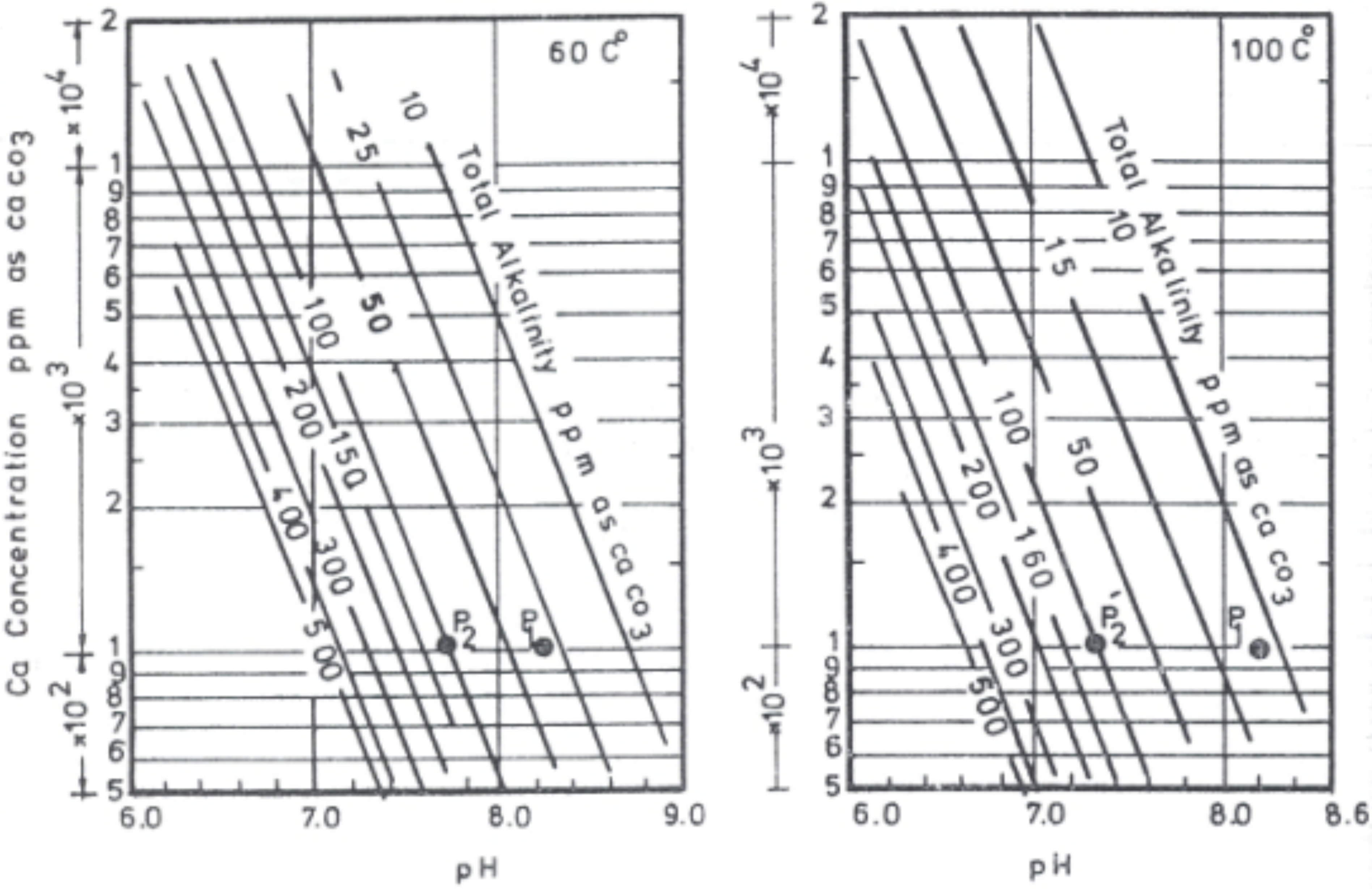
$$pH = 8.5$$

$$Temperature = 100C^o$$

$$Then \text{ from the stability diagram}$$

$$\text{pH}_\text{S} = 7$$

$$saturation \text{ index} = 8.5 - 7 = 1.5$$



STABILITY DIAGRAM OF
CALCIUM CARBONATE IN
CONCENTRATED SEA WATER

Scale Formation

10. Concluding Remarks:

- $CaCO_3$ can deposit at low temperature starting from 60 C^o
- $Mg(OH)_2$ forms at higher temperatures and concentrations.
- $CaSO_4$ can form at 100 C^o when concentration ratios approach 1.5.
- Gypsum can form at lower temperatures and concentrations ratios approaching 3
- In sea water stills it would thus be expected that:

Up to 60 C^o : only $CaCO_3$ scale

60-100 C^o: $CaCO_3$ and $Mg(OH)_2$ the latter increases with temperature and time

Above 100 C^o: $CaCO_3$ and $Mg(OH)_2$ and appreciable amounts of calcium sulphates which would increase with temperature.

Choice of Scale Prevention Method

The Choice of Scale Prevention technique depends on:

- Nature of raw water (composition, pH, alkalinity, etc.)

- Operating conditions:
  - Temperature (lowest temperature would be best, but would require larger size of equipment and other limitations).
  - Circulation velocity: The higher the better.
  - Retention time (should be minimum , and avoiding stagnation pockets).
  - Surface finish of heat transfer bodies: Smooth surface best.

- Economic Considerations
  - These are related to the unit size, location, availability of additives, operating and capital cost….etc.

Scale Prevention

Scale formation may be minimized by applying one or more of the following principles

- Process and unit design to minimize scaling
- Removal of scale formers (softening)
- Modifying the environmental conditions to avoid supersaturation (addition of acids that broaden the temp. range in which the water is stable)
- Modification of formation and growth mechanism of scale (Soft + Sludgy *not* hard and adherent) this can be achieved by adding polyphosphates, which helps in precipitating soft compounds which could be easily removed.

University of Jordan

Chemical Engineering Department

Membrane Distillation

Dr. Mohammed Rasool Qtaishat

Membrane Distillation Principle

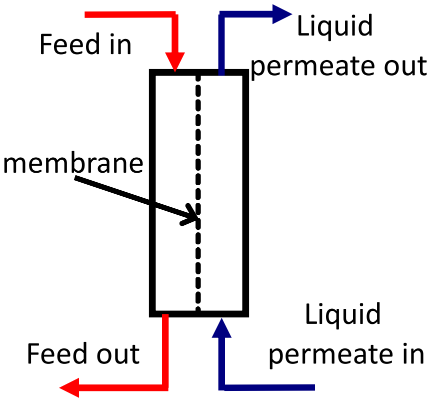
Membrane distillation (MD) is a thermally driven process in which a microporous membrane acts as a physical support separating a warm solution from a cooler chamber, which contains either a liquid or a gas, depending on the used configuration

MD Configurations

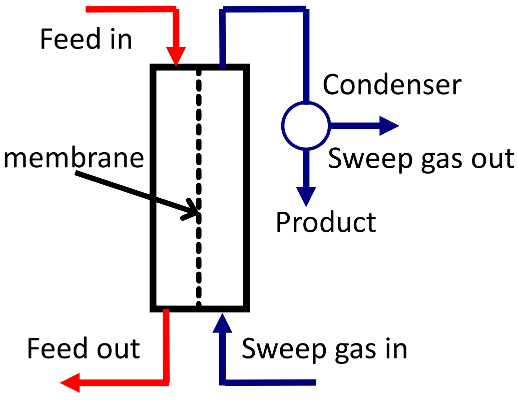
- **Direct contact membrane distillation (DCMD)**
- **Air gap membrane distillation (AGMD)**
- **Vacuum membrane distillation (VMD)**
- **Sweeping gas membrane distillation (SGMD)**

MD Configuration and Research Growth

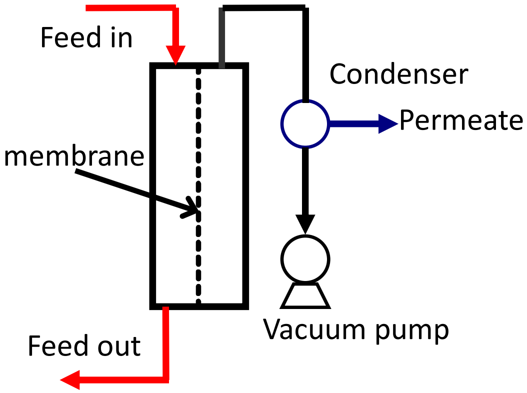
Traditional MD configurations



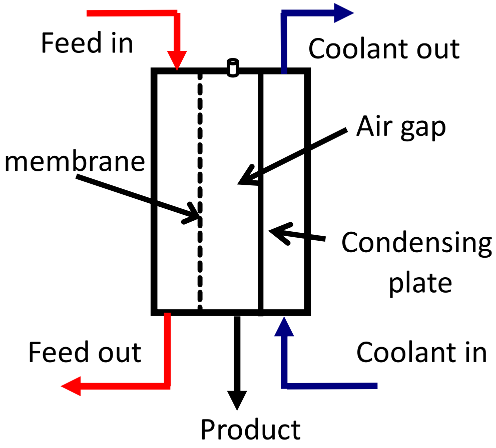
DCMD



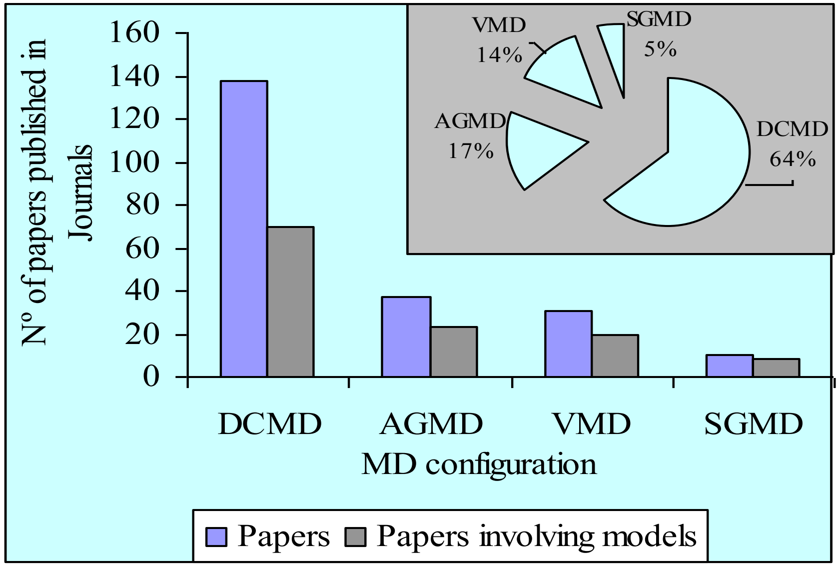
SGMD

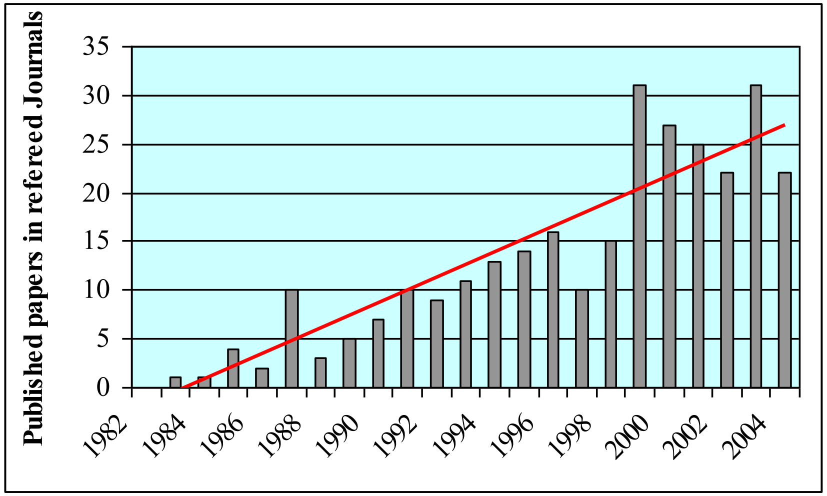


VMD



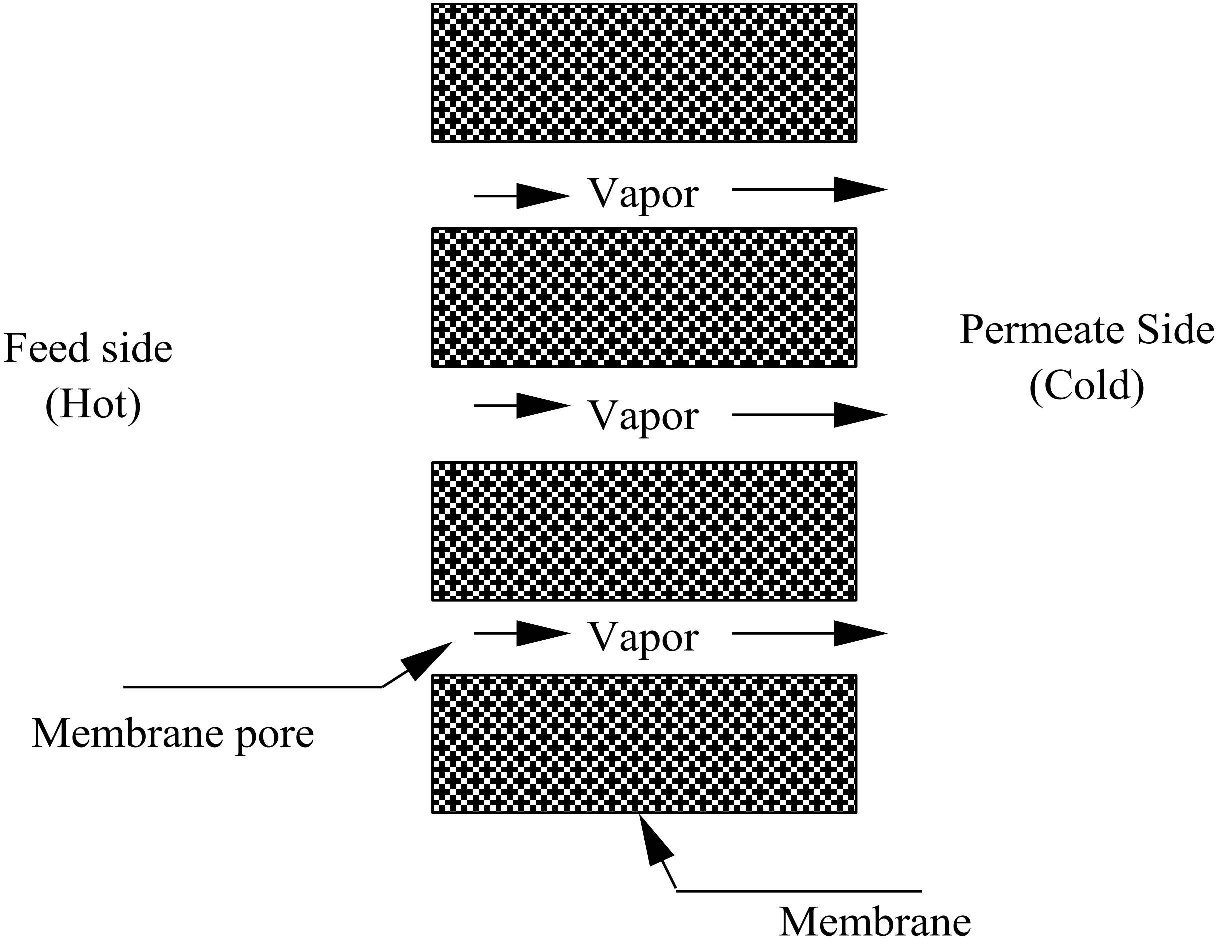
AGMD





Growth of MD research activity

MD Principle



MD Advantages

- **100% (theoretical) rejection of ions, macromolecules, colloids, cells and other non-volatiles.**
- **Lower operating temperatures than conventional distillation.**
- **Lower operating pressures than conventional pressure-driven membrane separation processes.**
- **Reduced vapor spaces compared to conventional distillation processes**

MD Disadvantages

- ***The main disadvantage of MD process is the danger of membrane wetting, the wettability of the microporous membranes is function of three main factors: the surface tension of the process solution, membrane material and the membrane structure***

- ***Laplace Equation*** $\Delta P = -\frac{2\gamma}{r_p}\cos\theta$ ***Where,***

- ***ΔP: is the pressure across the membrane (Pa)***
- ***γ: is the liquid surface tension (N/m)***
- ***r_p: is the pore radius (m)***

MD Disadvantages

➢ **Other Disadvantages that include:**

- ***Lower permeate flux*** **(Lower Water Production rate per unit area of the membrane) compared to conventional desalination process, (i.e. MSF, RO, etc.)**

- ***High cost*** **of membranes used in MD.**

Possible Applications of MD

- **Breaking azeotropic mixtures such as propionic acid-water mixture (*Thermodynamics*).**

- **Removal of volatile matters such as benzene (*Environmental*).**

- **Concentrating of non-volatile solutes (*Desalination*).**

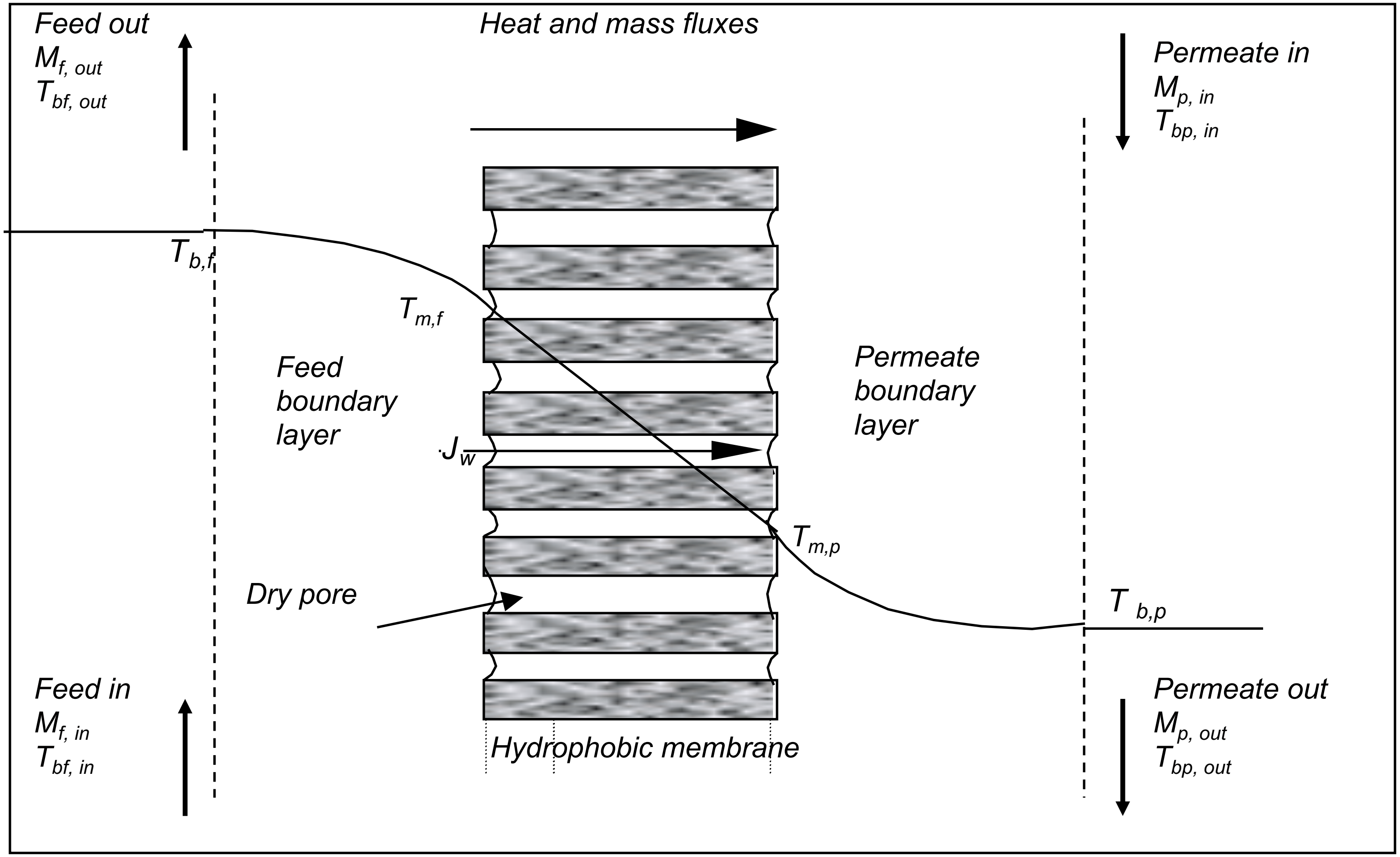
MD Membrane Requirements

- ***Porous***
- ***Entirely or partly hydrophobic***
- ***Good thermal stability***
- ***Excellent chemical resistance to feed solutions***
- ***High liquid entry pressure into the pores:***

⇒ high hydrophobicity & small maximum pore size

- ***High permeability: ⇒ high porosity, large pore size, low pore tortuosity and thickness.***
- ***Low thermal Conductivity.***

MD Theoretical Background



Heat Transfer in Membrane Distillation

The following heat transfer flux is involved in the DCMD system

1. Through the feed solution boundary layer:

$$Q_f = h_f (T_{b,f} - T_{m,f})$$

2. Through the hydrophobic top layer:

$$Q_m = h_m (T_{mf} - T_{mp}) + J_w \Delta H_v$$

3. Through the Permeate solution boundary layer:

$$Q_p = h_p (T_{mp} - T_{bp})$$

Heat Transfer in Membrane Distillation

- At steady state, the overall heat transfer flux through the whole DCMD system, *Q*, is given by

$$Q_f = Q_m = Q_p = Q$$

- Combining equations , the heat flux can be written as follows:

$$Q = \left(\frac{1}{h_f} + \frac{1}{h_m + \frac{J_w \Delta H_v}{T_{mf} - T_{mp}}} + \frac{1}{h_p} \right)^{-1} (T_{bf} - T_{bp})$$

Heat Transfer in Membrane Distillation

- As a result, the overall heat transfer coefficient (*U*) for the DCMD process may be written as:

$$U = \left(\frac{1}{h_f} + \frac{1}{h_m + \dfrac{J_w \Delta H_v}{T_{mf} - T_{mp}}} + \frac{1}{h_p} \right)^{-1}$$

Mass Transfer in Membrane Distillation

- In MD process, the mass transport is usually described by assuming a linear relationship between the mass flux (J_w) and the water vapour pressure difference through the membrane distillation coefficient (B_m):

$$J_w = B_m(p_{mf} - p_{mp})$$

- Where p_{mf} and p_{mp} are the partial pressures of water at the feed and permeate sides evaluated by using Antoine equation at the temperatures T_{mf} and T_{mp}, respectively; such as the following:

$$P^v = \exp\left(23.328 - \frac{3841}{T - 45}\right)$$

- Where P^v is the water vapour pressure in Pascal and T is the corresponding temperature in Kelvin.

Mass Transfer in Membrane Distillation

- Various types of mechanisms have been proposed for transport of gases or vapours through porous membranes:

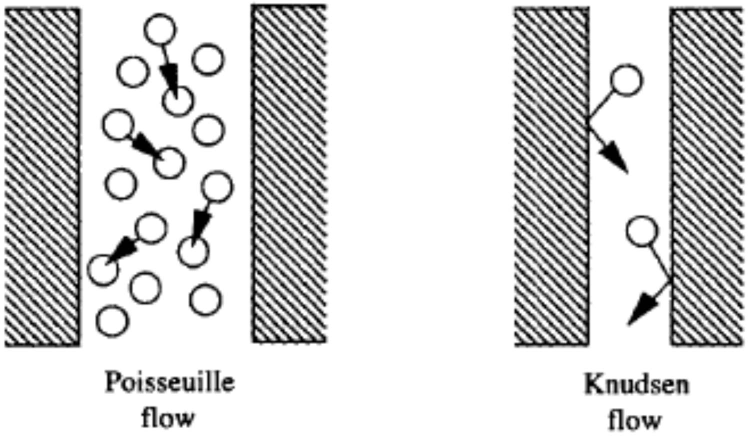
1. Knudsen model,
2. viscous model,
3. ordinary-diffusion model,
4. and/or the combination thereof.

- The governing quantity that provides a guideline in determining which mechanism is operative under a given experimental condition is the Knudsen number, K_n, defined as the ratio of the mean free path (λ) of the transported molecules to the pore size (diameter, d) of the membrane; i.e. $K_n = \lambda/d$.

Mass Transfer in Membrane Distillation

- **Knudsen and Poiseuille (viscous) flows**

⇒ Knudsen (molecular) is one of the mechanisms for gas separation in porous membranes.



⇒ The type of flow is determined by the ratio (r/λ) *of the mean free path of gas molecules* (λ) *and* the pore radius (r)*; alternatively by a Knudsen number* ($KN = \lambda /2r$)

Mass Transfer in Membrane Distillation

⇒ From the kinetic theory of gases:

$$\lambda = \frac{3\eta}{2p}\left(\frac{\pi RT}{2M}\right)^{1/2}$$

where M = molecular weight, p = pressure, η = viscosity of the gas, R = universal gas constant, T = absolute temperature

- For $r/\lambda << 1$ (large KN) ⇒ Knudsen flow
- For $r/\lambda >> 1$ (small KN) ⇒ Poiseulle flow
- For $r/\lambda \approx 1$ ($KN \approx 1$) ⇒ transition region

Mass Transfer in Membrane Distillation

- Due to the fact that in DCMD process both the hot feed and the cold permeate water are brought into contact with the membrane under atmospheric pressure, the total pressure is constant at ≈ 1 atm resulting in negligible viscous flow
- As a result the governing equations that describes the mass transfer are:

1. Knudsen region,
2. continuum region (or ordinary-diffusion region)
3. transition region (or combined Knudsen/ordinary-diffusion region).

Mass Transfer in Membrane Distillation

1. Knudsen Model:

if $Kn > 1$ (i.e. $r < 0.5\ \lambda$) the net DCMD membrane permeability can be expressed as follows

$$B_m^K = \frac{2}{3}\frac{\varepsilon r}{\tau\delta}\left(\frac{8M}{\pi RT}\right)^{1/2}$$

- Where ε, τ, r, δ are the porosity, pore tortuosity, pore radius and thickness of the hydrophobic membrane, respectively; M is the molecular weight of water, R is the gas constant and T is the absolute temperature.

Mass Transfer in Membrane Distillation

2. Continuum region (or ordinary-diffusion region)

if $Kn < 0.01$ (i.e. $r > 50\ \lambda$), molecular diffusion is used to describe the mass transport in continuum region caused by the virtually stagnant air trapped within each membrane pore due to the low solubility of air in water. In this case the following relationship can be used for the net DCMD membrane permeability

$$B_m^D = \frac{\varepsilon}{\tau\delta}\frac{PD}{P_a}\frac{M}{RT}$$

Mass Transfer in Membrane Distillation

- Where P_a is the air pressure, P is the total pressure inside the pore assumed constant and equal to the sum of the partial pressures of air and water liquid, and D is the water diffusion coefficient. The value of PD (Pa m^2/s) for water-air was calculated from the following expression

$$PD = 1.895\ 10^{-5} T^{2.072}$$

Mass Transfer in Membrane Distillation

3. The transition region,

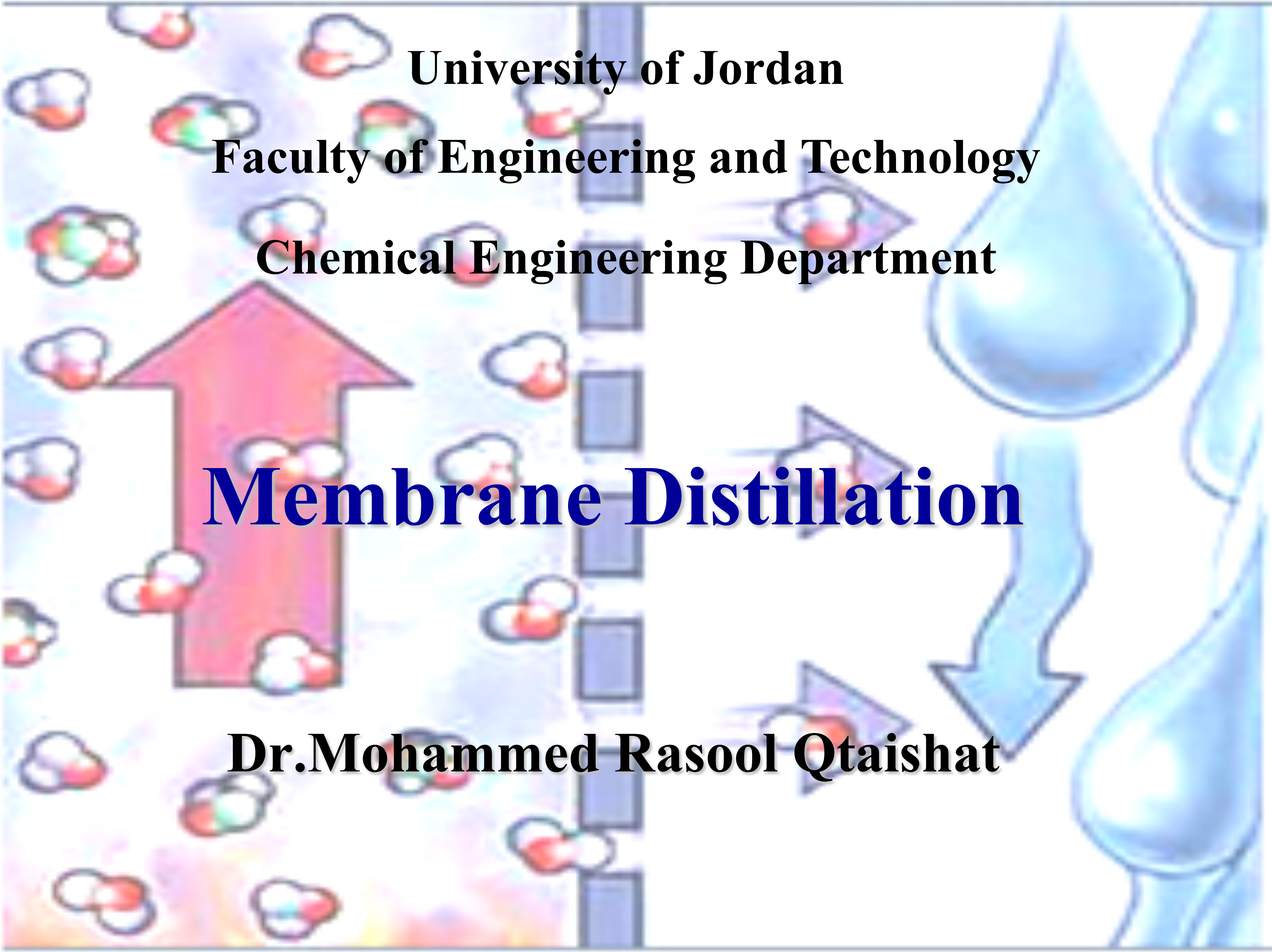
If $0.01 < Kn < 1$ (i.e. $0.5\lambda < r < 50\lambda$), the molecules of water liquid collide with each other and diffuse among the air molecules. In this case, the mass transport takes place via the combined Knudsen/ordinary-diffusion mechanism and the following equation is used to determine the water liquid permeability

$$B_m^C = \left[\frac{3}{2} \frac{\tau\delta}{\varepsilon r} \left(\frac{\pi RT}{8M} \right)^{1/2} + \frac{\tau\delta}{\varepsilon} \frac{P_a}{PD} \frac{RT}{M} \right]^{-1}$$

Empirical correlation [6]	**Flow Regime**
$Nu = 1.86(\mathrm{Re}\,\mathrm{Pr})^{1/3}$	**Laminar**
$Nu = 3.66$	**Laminar**
$Nu = 4.36$	**Laminar**
$Nu = 0.097\,\mathrm{Re}^{0.73}\,\mathrm{Pr}^{0.13}$	**Laminar**
$Nu = 1.95(\mathrm{Re}\,\mathrm{Pr})^{1/3}$	**Laminar**
$Nu = 0.13\,\mathrm{Re}^{0.64}\,\mathrm{Pr}^{1/3}$	**Laminar**
$Nu = 0.023\,\mathrm{Re}^{0.8}\,\mathrm{Pr}^{1/3}$	**Turbulent**
$Nu = 0.036\,\mathrm{Re}^{0.8}\,\mathrm{Pr}^{1/3}$	**Turbulent**
$Nu = 0.027\,\mathrm{Re}^{0.8}\,\mathrm{Pr}^{c}\left(\frac{\mu_{bf}}{\mu_{mf}}\right)^{0.14}$	**Turbulent**
$Nu^{b} = \frac{(f/8)\,\mathrm{Re}\,\mathrm{Pr}}{1.07 + 12.7(f/8)^{1/2}\left(\mathrm{Pr}^{2/3} - 1\right)}$	**Turbulent**
$Nu^{b} = \frac{(f/8)(\mathrm{Re} - 1000)\,\mathrm{Pr}}{1 + 12.7(f/8)^{1/2}\left(\mathrm{Pr}^{2/3} - 1\right)}$	**Turbulent**

The friction factor, *f*, in these correlation was estimated by:

$$f = (0.79\ln(\mathrm{Re}) - 1.64)^{-2}$$



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Membrane Distillation

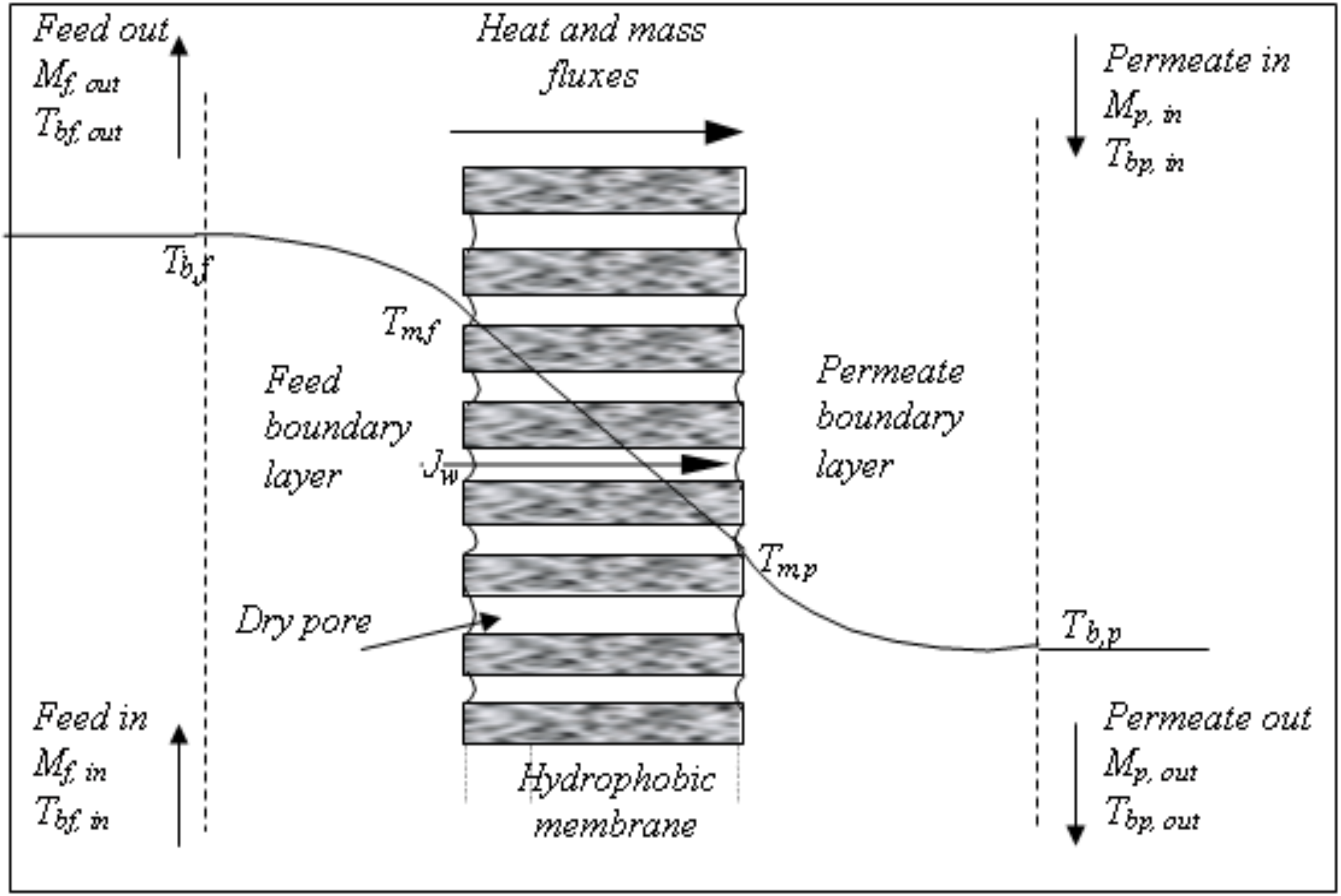
Dr.Mohammed Rasool Qtaishat

Temperature polarization

- The driving force for water vapour migration through the membrane pores is the temperature difference between the feed/membrane interface temperature (T_{mf}) and the permeate/membrane interface temperature (T_{mp}).

- This temperature difference leads to a decrease from the theoretical driving force, which is defined as the difference between the bulk feed temperature (T_{bf}) and the bulk permeate temperature (T_{bp}). This phenomenon is known as temperature polarization.

Temperature polarization



Temperature polarization

- The temperature polarization coefficient (TPC) is defined as the ratio between the actual driving force and the theoretical driving force. As a result the temperature polarization coefficient is expressed mathematically as following:

$$TPC = \frac{T_{mf} - T_{mp}}{T_{bf} - T_{bp}}$$

Temperature polarization

- It is impossible to measure the membrane/interface temperatures experimentally; usually these temperatures are evaluated by performing a heat balance that relates them to the bulk temperatures.

- In order to solve this heat balance for membrane interface temperatures, the heat transfer coefficients in the adjoining liquid boundary layers to the membrane should be evaluated.

Temperature polarization

- Generally, the boundary layers heat transfer coefficients are evaluated using empirical correlations for the determination of Nusselt number and a wide variety of these correlations are available in the literature.

- Sometimes it is difficult to select the proper empirical correlation that best describes the heat transport in the used MD module.

Evaporation Efficiency (EE)

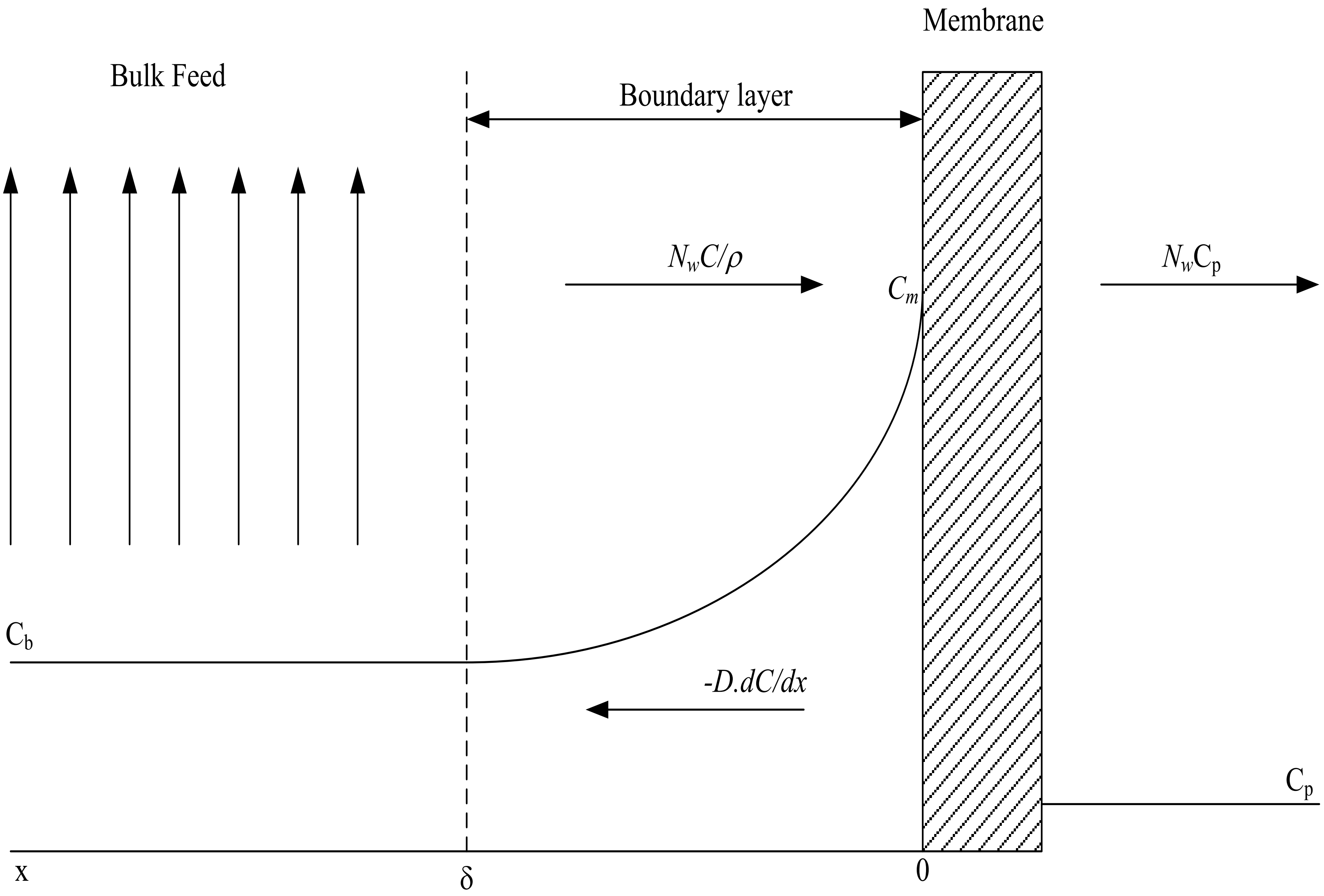
- The evaporation efficiency, *EE*, is defined as the ratio between the heat transferred because of water vapour migration through the membrane pores and the total heat transferred through the membrane. Mathematically, the evaporation efficiency is expressed by

$$EE = \frac{Q_{m,M.T}}{Q_{m,M.T} + Q_{m,cond}} = \frac{J_w H_v}{J_w H_v + h_m (T_{mf} - T_{mp})}$$

$Q_{m,M.T}$:heat transferred because of water vapour migration through the membrane pores.

$Q_{m,cond}$:conductive heat transfer through the membrane.

Concentration polarization



Concentration polarization

- In desalination, localized concentration of solute build up at the point where the solvent leaves the solution and enters the membrane. The solute accumulates in a relatively stable boundary layer next to the membrane. Concentration polarization,ζ or CPF (Concentration Polarization Factor), is defined as the ratio of the salt concentration at the membrane surface C_m to the salt concentration in the bulk feed stream C_b.

$$\zeta = C_m/C_b$$

Concentration polarization

- At steady state the convective flux equals the diffusion flux

$$\frac{N_w C}{\rho} = -D\frac{dC}{dx}$$

$N_w C/\rho$ =[kg solvent/(s.m^2)](kg solute/m^3)/(kg solvent/m^3) = kg solute/s.m^2

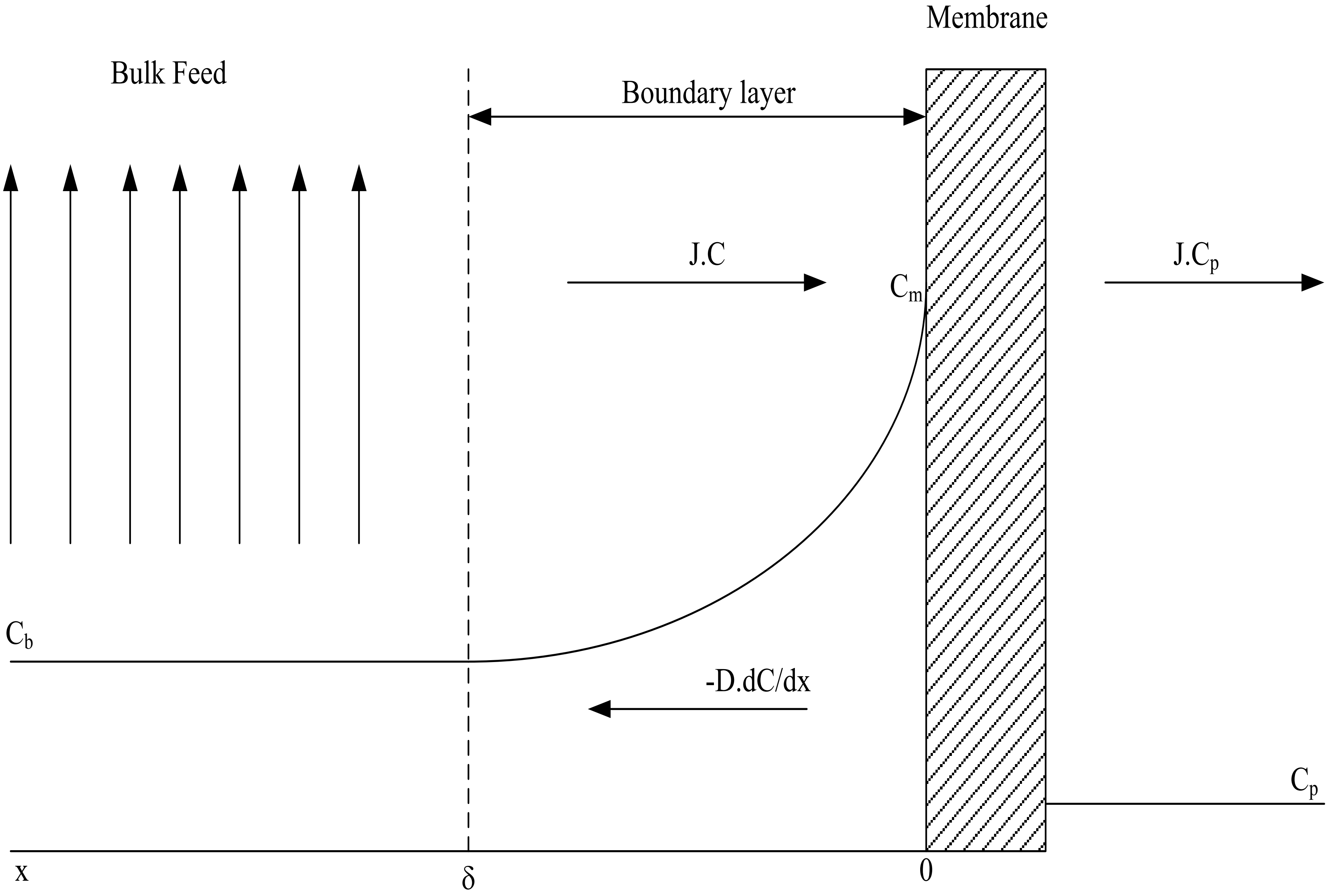
D: is the diffusivity of solute in solvent, m^2/s; and x is the distance, m. Integrating this equation between the limits of x=0 and $C=C_m$ and $x=\delta$ and $C=C_b$

Concentration polarization

$$\frac{N_w}{\rho} = \left(\frac{D}{\delta}\right)\ln\left(\frac{C_m}{C_b}\right) = k_c \ln\left(\frac{C_m}{C_b}\right)$$

where k_c is the mass transfer coefficient, m/s.

Concentration polarization



Concentration polarization

We can define ζ as:

$$\zeta = \frac{\exp(J/k)}{R_{\text{int}} + (1 - R_{\text{int}})\exp(J/k)}$$

- Where R_{int} is the intrinsic retention factor defined as

$$R_{\text{int}} = 1 - \frac{C_p}{C_m}$$

- J is the flux (m/s) and k is the solute mass transfer coefficient (m/s). C_b, C_m, and C_p, are the bulk, membrane and permeate concentrations (kg/m^3), respectively.

MULTIPLE EFFECT EVAPORATOR

INTRODUCTION

- The multiple effect evaporation system is formed a sequence of single effect evaporators, where the vapor formed in one effect is used in the next effect.
- The vapor reuse in the multiple effect system:

a) Allows reduction of the brine and the temperature to low values.
b) Prevent rejection of large amount of energy to the surrounding,

- The main bulk of the multiple effect evaporation processes is found in the food, pulp and paper, petroleum, and petrochemical industries

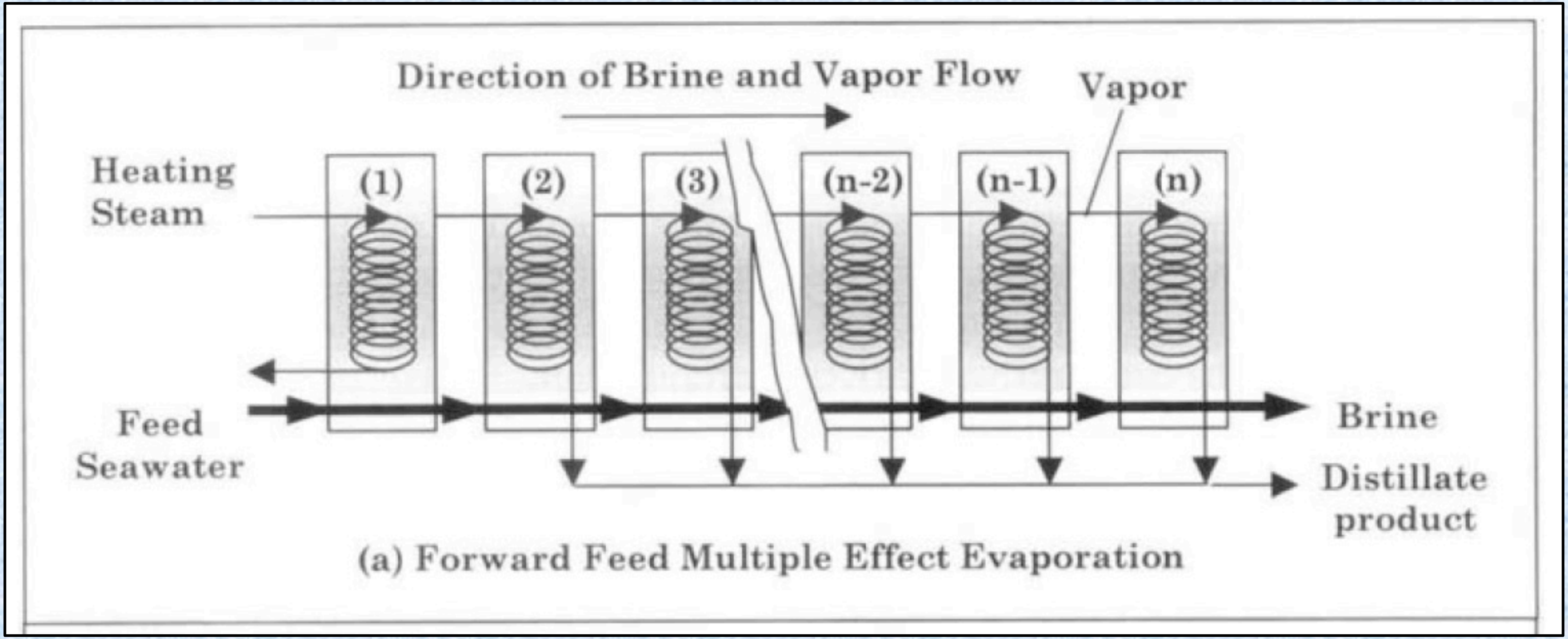
INTRODUCTION

- The multiple effect evaporation process has three configurations:

1. Forward feed Multiple Effect Evaporation (MEE-FF).
2. Backward feed Multiple Effect Evaporation (MEE-BF)
3. Parallel or Cross Flow flwo Multiple Effect Evaporation (MEE-P/C).

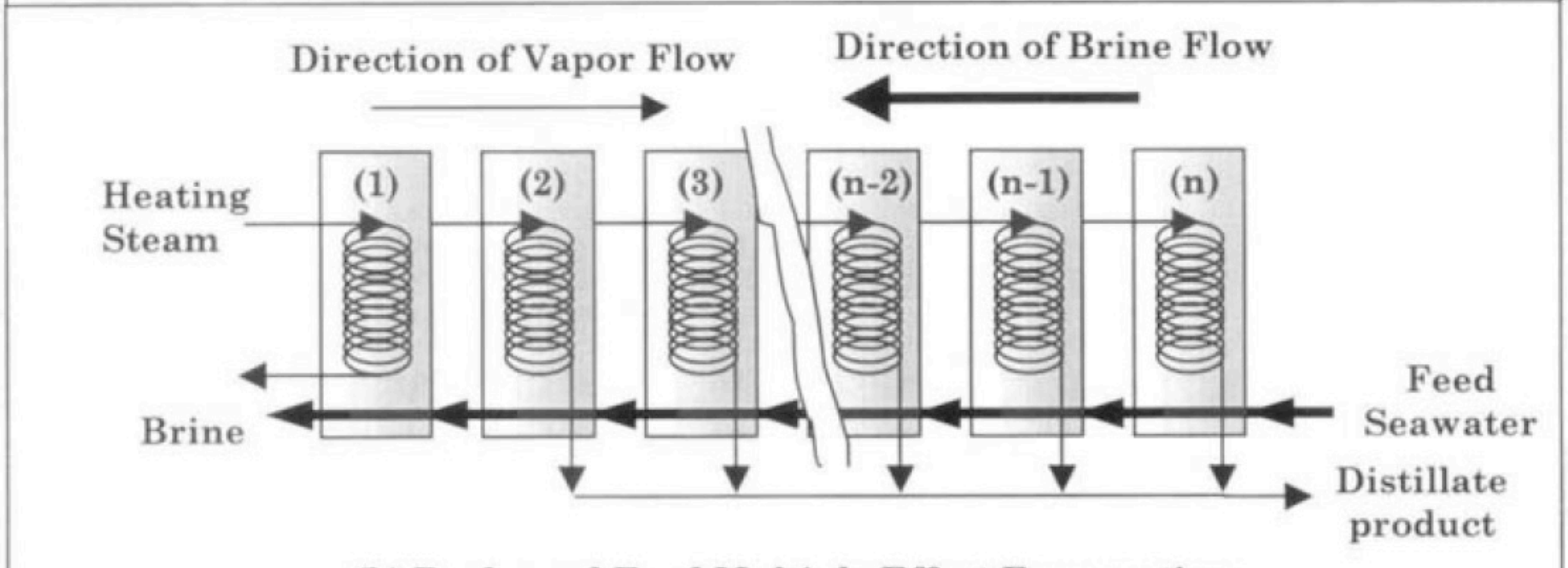
- The three configurations differ in the flow directions of the heating steam and the evaporating brine.

MEE-FF



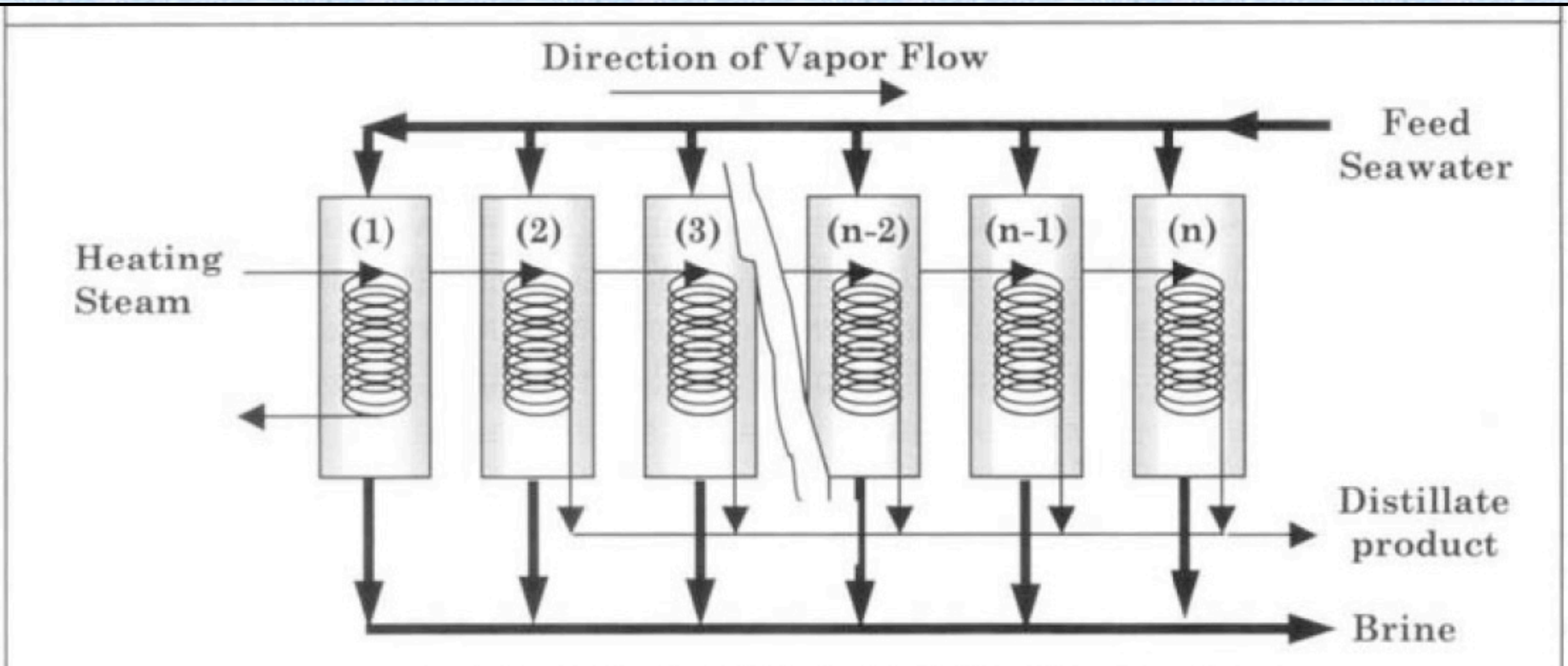
(a) Forward Feed Multiple Effect Evaporation

MEE-BF



(b) Backward Feed Multiple Effect Evaporation

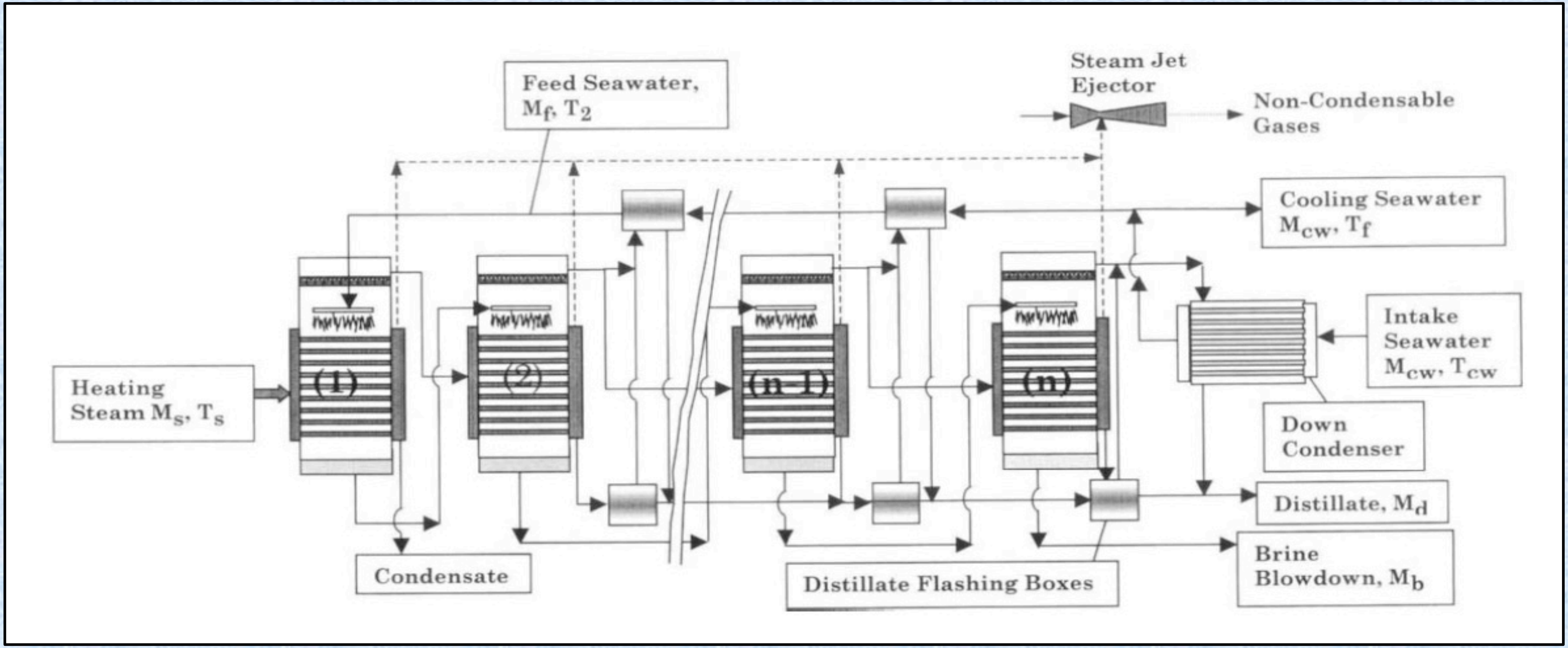
MEE-P/C



(c) Parallel Feed Multiple Effect Evaporation

FORWARD FEED MULTIPLE EFFECT EVAPORATION

Figure 1 shows a schematic diagram for the forward-feed Multiple-Effect Evaporation (MEE-FF) seawater desalination process.



- The system includes:

1. The evaporators, equal to n.
2. A series of feed water preheaters, equal to n-2
3. A train of flashing boxes, equal to n-1
4. A down condenser, and a venting system

- The direction of **heat flow** as well as the **flow direction** of the brine and vapor is from **left to right**, i.e., from effect 1 to effect n.

- The pressure in the effects decreases in the flow direction. $P_1>P_2>P_3$

- Each effect (evaporator) contains:

a) Heat exchange tubes b) vapor space c) brine spray nozzles
d) mist eliminator e) brine collecting box

Horizontal falling film evaporator:

The most widely used in the MEE desalination processes.

- Advantages:

1. Efficient water distribution and tube wetting,
2. High heat-transfer rates,
3. Absence of dry patches,
4. Low scale formation and tube damage,
5. Efficient disengagement of vapors and non-condensable gases,
6. Proper venting of the non-condensable gases, and
7. Simple monitoring of scaling and fouling.

- Disadvantages:

Scale and fouling on the outer surface of the tubes.

Such system proved to reduce internal scaling and fouling by 50% of the design value

PROCESS DESCRIPTION

- The intake seawater flows into the condenser of the last effect at a flow rate of M_{cw} $+M_f$. This stream absorbs the latent heat of vapors formed in the last effect and flashing box.
- The seawater stream is heated from the intake temperature, T_{cw} to a higher temperature, T_f.
- The function of the cooling seawater, M_{cw} is to remove the excess heat added to the system in the first effect by the motive steam.
- On the other hand, the feed seawater, M_f, is heated by the flashed off vapors formed in the last effect and the associated water flash box.
- The cooling seawater, $M_{cw}$is rejected back to the sea
- The temperature of the feed water increases from T_f to t_2 as it flows inside the tubes of the preheaters.

PROCESS DESCRIPTION

- Heating of the feed seawater is made by condensing the flashed off vapors from the effects, d_j, and the flash boxes, $\acute{d}_j$

- The feed water, M_f, leaves the last preheater (associated with the second effect) and is sprayed inside the first effect.

***It is interesting to note that the preheater of the first effect is integrated in the heat exchanger of the effect. This is because there is no flash box in the first effect or flashed off vapors within the effect.

- The brine spray forms a thin film around the succeeding rows of horizontal tubes. Its temperature rises to the boiling temperature, T_1, which corresponds to the pressure of the vapor space.

PROCESS DESCRIPTION

- The saturation temperature of the formed vapor, T_{v1} is less than the brine boiling temperature by the boiling point elevation, $(BPE)_1$.

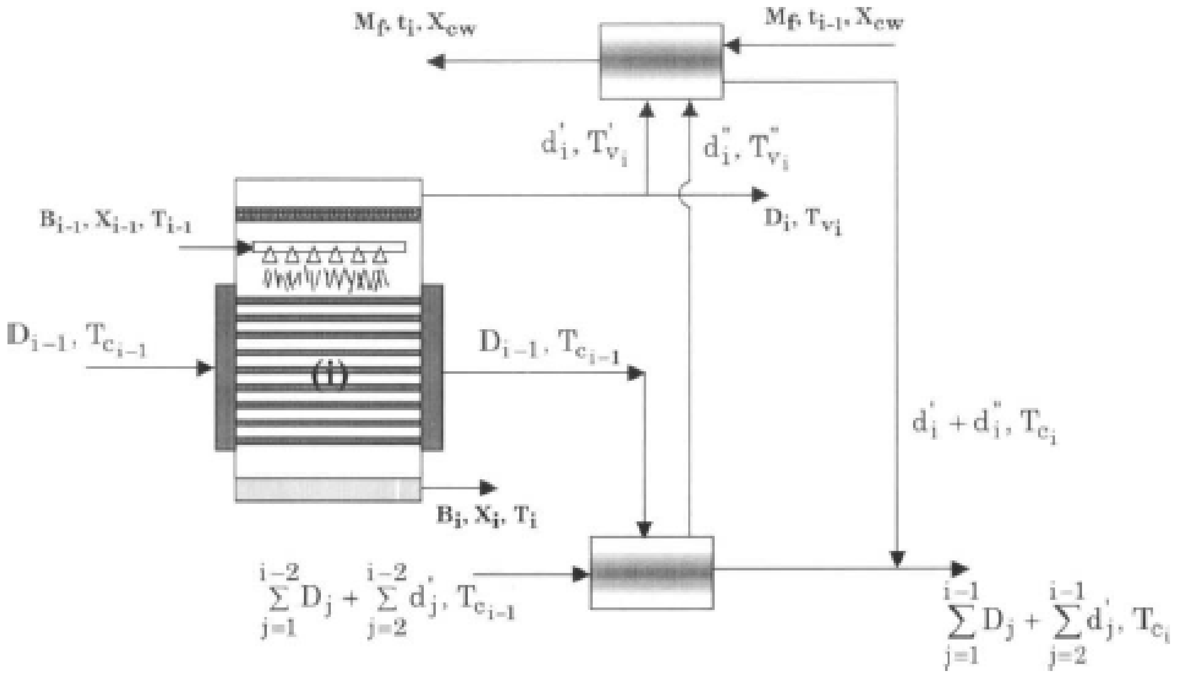
- A small portion of vapor, D_1, is formed by boiling in the first effect. The remaining brine, $M_f - D_1$, flows into the second effect, which operates at a lower temperature and pressure.

- Vapor is formed in effects 2 to n by two different mechanisms, boiling and flashing. The amount vapor formed by boiling is D_j and the amount formed by flashing is d_j

- Flashing occurs in effects 2 to n because the brine temperature flowing from the previous effect, T_{j-1} is higher than the saturation temperature of the next effect, T_{vj} Therefore, vapor flashing is dictated by the effect equilibrium.
- In effects 2 to n, the temperature of the vapor formed by flashing, T'_{vj} is lower than the effect boiling temperature, T_j by the boiling point elevation $(BPE)_j$ and the non-equilibrium allowance $(NEA)_j$.

PROCESS DESCRIPTION

- In the flash boxes, a small quantity of flashing vapors, $\bar{d}_j$ is formed with a temperature equal to T''_{vj}. This temperature is lower than the vapor condensation temperature in effect j , T_{cj}, by the non-equilibrium allowance $(NEA)'_j$.



- Motive steam, M_s, extracted from an external boiler drives vapor formation in the first effect. The vapor formed by boiling in the first effect, D_1, is used to drive the second effect, which operates at a lower saturation temperature, T_2.

- Reduction in the vapor temperature is caused by:

1. Boiling point elevation,
2. Non-equilibrium allowance (NEA)
3. Losses caused by depression in the vapor saturation pressure by frictional losses in the demister, transmission lines, and during condensation.

As show in the last figure:

1. the brine leaving the effect decreases by the amount of vapor formed by boiling, D_j, and by flashing, d_j.
2. The distillate flow rate leaving the flash box increases by the amount of condensing vapors from the previous effect, D_{j-1} and d_{j-1}
3. The brine concentration increases from $X_{j-1}$to X_j upon vapor formation
4. The effect and flash box temperatures decrease from $T_{j-1}$to T_j and from $T_{j-1}^{`}$ to $\grave{T}_j$ respectively.

END OF CHAPTER 8

SINGLE EFFECT EVAPORATION VAPOR COMPRESSION

INTRODUCTION

- Single effect evaporation system combined with various types of heat pumps, includes:

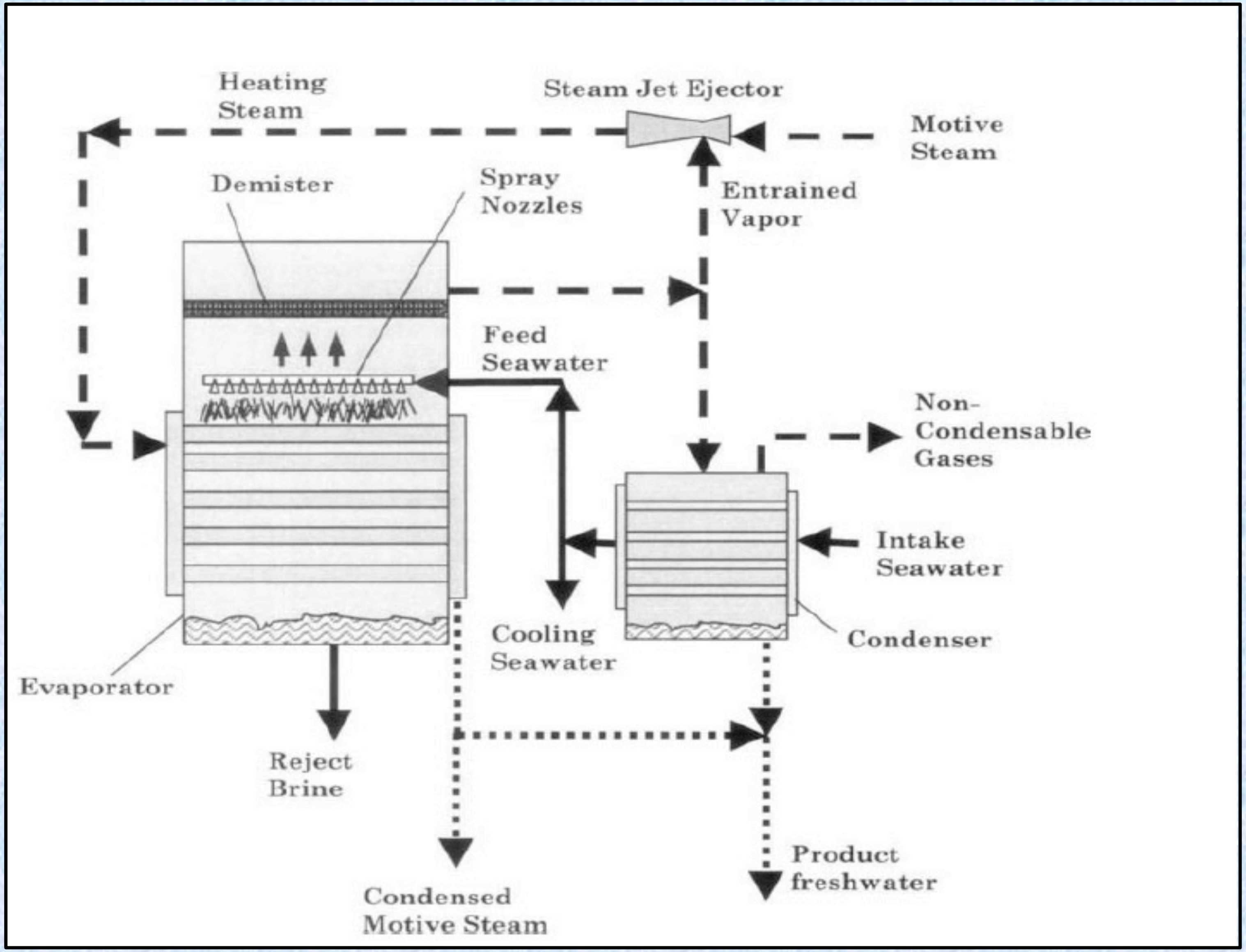
1. Thermal Vapor Compression TVC
2. Mechanical Vapor Compression MVC
3. Absorption Vapor Compression ABVC
4. Adsorption Vapor Compression ADVC

This chapter focuses on evaluation of the single effect evaporation system combined with various types o f heat pumps. The evaluation in the following:

- Process description.
- Model development.
- Performance evaluation

SINGLE EFFECT THERMAL VAPOR COMPRESSION TVC

Figure 1 shows a schematic diagram for the single-effect thermal vapor-compression



The main components of the unit are:

- The evaporator.
- the steam jet ejector.
- the feed condenser or heater .

PROCESS DESCRIPTION

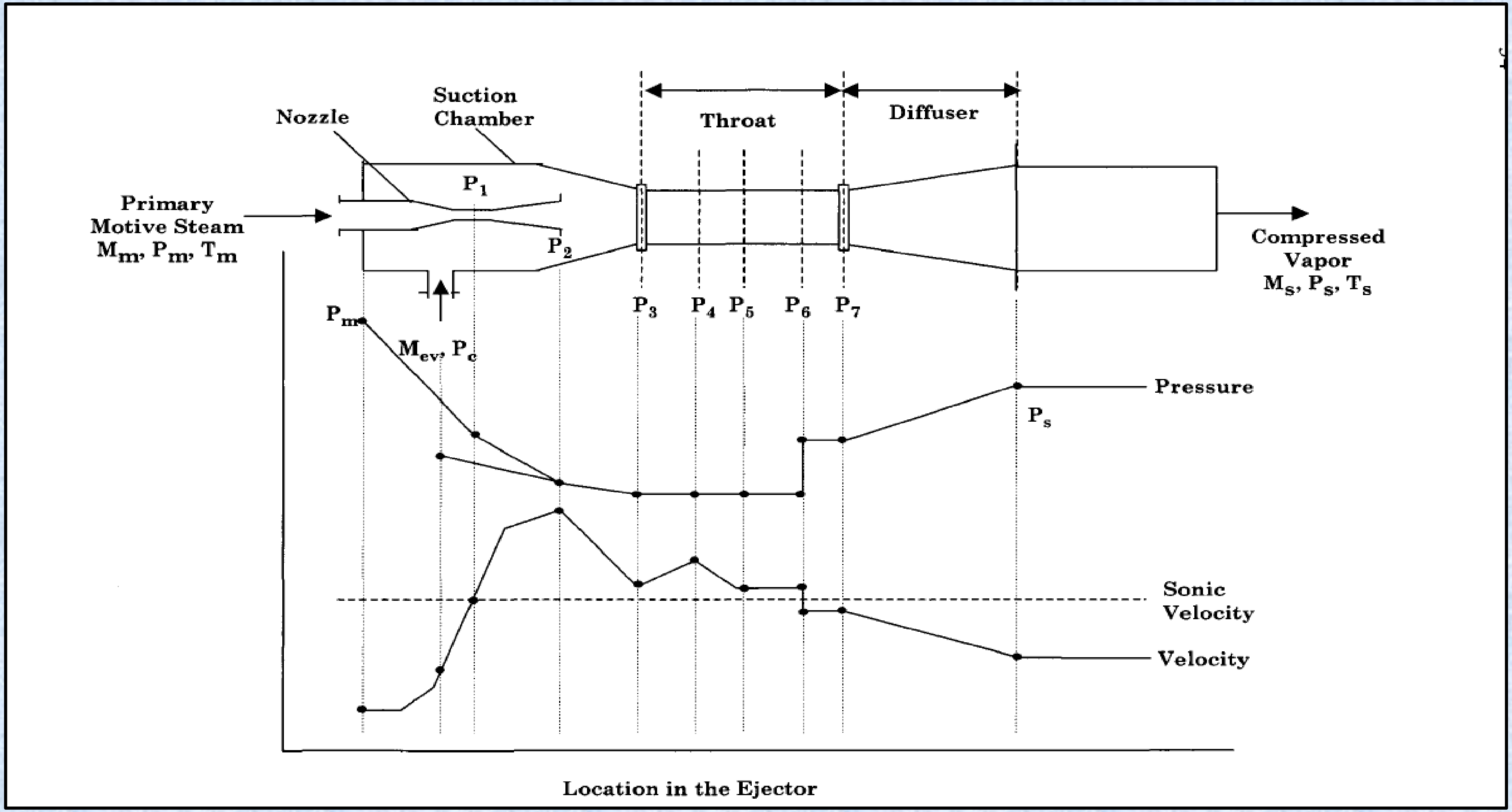
- The evaporator consists of :

1. Evaporator/condenser heat exchange tubes
2. Vapor space
3. Un-evaporated water pool
4. Line for removal of non-condensable gases
5. Water distribution system,
6. Mist eliminator

- The feed heater or the heat sink unit is usually a counter-current surface condenser in which the non-condensable gases leave at a temperature approaching the temperature of the feed water.

STEAM JET EJECTOR

- The steam jet ejector is composed of a steam nozzle, a suction chamber, a mixing nozzle, and a diffuser.



STEAM JET EJECTOR

- The ejector is used to increase the pressure of the entrained vapor M_{ev} from pressure P_{ev} to a higher pressure P_s.
- As the motive steam at flow rate of M_m expands in the nozzle from state 1 to state 2, its static pressure energy is converted to kinetic energy.
- The nozzle is a converging/diverging shape to expand the steam to velocities greater than the speed of sound (supersonic).
- The suction chamber is used to keep the nozzle properly positioned with respect to the diffuser and to direct the entrained vapor
- The entrained vapor M_{ev} enters the suction chamber at pressure P_{ev} where it mixes with the motive steam.

STEAM JET EJECTOR

- The two streams mix together as they pass through the converging section of the venturi diffuser. The mixture enters the throat section of the diffuser, completely mixed, at the sonic velocity of the mixture.
- The mixture leaves the ejector at a pressure P_s that is intermediate to the motive (P_m) and suction (P_{ev}) pressures.

PROCESS DESCRIPTION

- The intake seawater ($M_{cw}+M_f$) @ $T=T_{cw}$ and salt conc. X_f Introduced into the tube side of the preheater where its Temperature increases to (T_f).
- The cooling water (M_{cw})is dumped back to the sea. Its function is to remove of the excess heat added to the system in the form of motive steam necessary to drive the steam jet ejector
- The heating of the feed seawater (M_f) in the condenser tubes from (T_{cw}) to (T_f) is essential to increase the thermal performance of the process
- The heat needed to warm the seawater inside the condenser is supplied by condensing a controlled portion of vapor formed by boiling in the evaporator M_C.

PROCESS DESCRIPTION

- The vapor condensation temperature and consequently the pressure in the vapor space for both the evaporator and the condenser is controlled by:

1. The cooling water flow rate, M_{cw}
2. The feed water temperature, T_{cw}
3. The available heat transfer area in the condenser, A_c
4. The overall heat transfer coefficient between the condensing vapor and the circulating seawater, U_c.

PROCESS DESCRIPTION

- Accordingly, the condenser has three functions:

1. Removes the excess heat from the system.
2. Improves the process performance ratio.
3. Adjusts the boiling temperature inside the evaporator.

PROCESS DESCRIPTION (STEPS)

- The feed seawater (Mf) is chemically treated and deaerated before being pumped to the evaporator.
- Condensation of the saturated heating steam and release of its latent heat provides the required sensible and latent for water evaporation from the feed seawater.
- As a result, the feed water temperature (T_f) is raised to the boiling temperature (T_b).
- The vapor formed by boiling with a rate of (M_d) is free of salts.
- The temperature of the generated vapor T_v is less than the boiling temperature T_b by the boiling point elevation (BPE).

PROCESS MODELING

The model for the single-effect evaporation system is divided into six parts:

1. Performance parameter.
2. Material balances.
3. Evaporator and condenser energy balances.
4. Boiling point elevation and thermodynamic losses.
5. Evaporator and condenser heat transfer area.
6. Steam jet ejector design equations

1. Performance parameter.

Performance of the TVC is determined in terms of the following variables:

a) The amount of product fresh water per unit mass of motive steam, or the performance ratio, PR

$$\boxed{PR = \frac{M_d}{M_m}}$$

b) The specific heat transfer surface area, sA.

$$\boxed{sA = \frac{A_e + A_c}{M_d}}$$

c) The specific cooling water flow rate, sMcw

$$\boxed{sM_{cw} = \frac{M_{cw}}{M_d}}$$

2. Material Balances (assume that the distillate water is salt free)

$$M_f = M_d + M_b$$

$$M_f X_f = M_b X_b$$

$$M_b = M_d \left(X_f / (X_b - X_f)\right)$$

$$M_f = M_d \left(X_b / (X_b - X_f)\right)$$

3.1 Evaporator Energy Balances

- In the evaporator, the dry saturated steam flowing from the steam jet ejector and admitted into the evaporator (M_m + M_{ev}) is used in to raise the temperature of the feed seawater M_f from the inlet temperature T_f to the boiling temperature T_b.
- In addition, it supplies the latent heat required to evaporate the specified mass of vapor, Md, or:

$$\boxed{Q_e = M_f\, C_p\, (T_b - T_f) + M_d\, \lambda_v = (M_m + M_{ev})\, \lambda_s} \quad \textbf{Eq (1)}$$

3.2 Condenser Energy Balances

- The **CONDENSER** operates on the vapor formed in the evaporator, (M_c) which is not entrained by the steam jet ejector
- The latent heat of condensation is transferred to feed seawater with a mass flow rate of (M_f+M_{cw}).
- The feed seawater (M_f) is introduced into the evaporator; while the remaining part (M_{cw}), which is known as the cooling water, is rejected.

The heat load of the condenser is given by

$$\boxed{Q_c = (M_f + M_{cw})\,C_p\,(T_f - T_{cw}) = M_c \lambda_c} \quad \textbf{Eq (2)}$$

4 .Boiling Point Elevation and Thermodynamic Losses

- The generated vapor is at the saturation temperature, T_v, which corresponds to the pressure in the evaporator vapor space. This temperature is less than the boiling temperature T_v by the boiling point elevation BPE, where,

$$\boxed{T_b = T_v + BPE}$$

Eq (3)

- The condensation temperature of vapor outside the tube bundle of the condenser T_c is less than the boiling temperature in the evaporator T_b by the boiling point elevation (BPE) and the saturation temperature depression associated with pressure losses in the demister (ΔT_p) and inside the condenser horizontal tubes (ΔT_c) .Thus

$$\boxed{T_c = T_b - (BPE + \Delta T_p + \Delta T_c)}$$

Eq (4)

5.1 Evaporator Heat Transfer Area

The dimensions of the required heat transfer surface area in the evaporator $\mathbf{A_e}$ are obtained from:

- ✓ The amount of the heat to be transferred $\mathbf{Q_e}$.
- ✓ The overall heat transfer coefficient $\mathbf{U_e}$.
- ✓ The difference between the condensation temperature of the steam, $\mathbf{T_s}$
- ✓ The boiling temperature of the seawater, $\mathbf{T_b}$.

This relation is given by

$$\boxed{A_e = Q_e/(U_e\,(T_s - T_b))} \quad \textbf{Eq (5)}$$

5.2 Condenser Heat Transfer Area

The heat transfer between the condensing vapor and the feed water in the condenser can be written in terms of an overall heat transfer coefficient (U_c), condenser heat transfer area (A_c), and the logarithmic mean temperature difference $(LMTD)_c$, thus:

$$\boxed{A_c = \frac{Q_c}{U_c (LMTD)_c}}$$

Eq (6)

The (LMTD)c is defined as:

$$\boxed{(LMTD)_c = \frac{T_f - T_{cw}}{\ln \dfrac{T_c - T_{cw}}{T_c - T_f}}}$$

Eq (7)

6. Steam Jet Ejector

- The most important and critical step in modeling the TVC desalination system is the evaluation of the performance of the steam jet ejector

- The main data required from analyzing the steam jet ejector is the **determination of the mass of motive steam required per unit mass of the entrained vapor (Ra),** given the pressure of the motive steam (P_m), discharge pressure (P_s) and the suction pressure (P_{ev}).

  - The following relationships is used to evaluate the performance of the steam jet ejector. The entertainment ratio is defined by:

$$\boxed{\mathrm{Ra} = 0.296 \frac{(\mathrm{P_s})^{1.19}}{(\mathrm{P_{ev}})^{1.04}} \left(\frac{\mathrm{P_m}}{\mathrm{P_{ev}}}\right)^{0.015} \left(\frac{\mathrm{PCF}}{\mathrm{TCF}}\right)}$$

Eq (8)

$$\boxed{\mathbf{Ra} = \frac{M_m}{M_{ev}}}$$

Where:

- **Ra :** The entrainment ratio and defined as the mass of motive steam per unit mass of entrained vapor.
- P_m , P_s and P_{ev} are the pressures of the **motive steam**, **discharge mixture** and **entrained vapor** respectively
- **PCF** : Motive steam pressure correction factor
- **TCF** : The entrained vapor temperature correction factor.

$$\boxed{PCF = 3x10^{-7}\,(P_m)^2 - 0.0009\,(P_m) + 1.6101}$$

Eq (9)

$$\boxed{TCF = 2x10^{-8}\,(T_{ev})^2 - 0.0006\,(T_{ev}) + 1.0047}$$

Eq (10)

P_m is in kPa and T_{ev} is in °C.

These equations are valid in the following ranges:

- $Ra < 4$
- $500 > T_{ev} > 10$ °C
- $3500 > Pm > 100$ kPa
- $\frac{P_s}{P_{ev}} \geq 1.81$

$$\boxed{\text{Compression ratio, Cr} = \frac{P_s}{P_{ev}}}$$

EXAMPLE 1:

A single-effect thermal vapor-compression system is designed at the following operating conditions:

- Boiling temperature, T_b, of 75 ^oC.
- Compression ratio, Cr, of 2.5.
- Motive steam pressure, $P_m$of 750 kPa.
- $P_v = 37.1kPa$
- ΔP_p The demister pressure drop is negligible
- Brine reject concentration, $X_b = 70000$ ppm
- Intake seawater salinity, $X_f = 42000$ ppm
- Intake seawater temperature, $T_{cw} = 25\ ^oC$
- System capacity, $M_d = 1$ kg/s
- Feed seawater temperature, $T_f = (T_b - 5) = 70\ ^oC$
- Condenser efficiency, η= 0.9.

Evaluation of the TVC system.

This includes a design problem to determine:

1. The specific heat transfer area.
2. The flow rate of the cooling water
3. The performance ratio.

Solution:

Substituting for X_f = 42000 ppm, X_b = 70000 ppm, and M_d = 1 kg/s in MB equation:

$$\boxed{M_f = X_b/(X_b - X_f) = 70000/(70000 - 42000) = 2.5 \text{ kg/s}}$$

$$\boxed{M_b = M_f - M_d = 2.5 - 1 = 1.5 \text{ kg/s}}$$

The boiling point elevation, BPE, is calculated from the correlation **@ T_b and X_b**

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

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

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

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

$$BPE = (0.0825431+0.0001883\,(75)+0.00000402\,(75)^2)\,(7) + (-0.0007625+0.0000902\,(75)-0.00000052\,(75)^2)\,(7)^2 + (0.0001522-0.000003\,(75)-0.00000003\,(75)^2)\,(7)^3 = 0.903\ ^oC$$

$$T_v = T_b - BPE = 75 - 0.903 = 74.097\ ^oC$$

$$Pev = Pv - \Delta Pp = 37.1 - 0 = 37.1\ kPa$$

Compression ratio, $Cr = \frac{P_s}{P_{ev}}$ ⇒ $P_s = (Cr)\,(P_{ev}) = (2.5)\,(37.1) = 92.75\ kPa$

The correlation for the saturation temperature of water vapor is given by:

$$T = \left(42.6776 - \frac{3892.7}{(\ln(P/1000) - 9.48654)}\right) - 273.15$$

$$T_s = \left(42.6776 - \frac{3892.7}{(\ln(P_s/1000) - 9.48654)}\right) - 273.15$$

$$= \left(42.6776 - \frac{3892.7}{(\ln(92.75/1000) - 9.48654)}\right) - 273.15$$

$$= 97.6\ ^oC$$

$U_e = 1.9695+1.2057xl0^{-2}Tb-8.5989xl0-5(Tb)^2+2.5651xl0^{-7}(Tb)^3$

Chapter 5

$$U_e = \begin{pmatrix} 1969.5 + 12.057\,(T_b) - 0.85989x10^{-1}\,(T_b)^2 \\ + 0.25651x10^{-3}\,(T_b)^3 \end{pmatrix} x10^{-3}$$

$$= \begin{pmatrix} 1969.5\ \ + 12.057\,(75) - 0.85989x10^{-1}\,(75)^2 \\ + 0.25651x10^{-3}\,(75)^3 \end{pmatrix} x10^{-3}$$

$$= 2.62\ kW/m^2\ {}^oC$$

$U_c = 1.7194+3.2063xl0^{-3}T_c+1.5971xl0^{-5}(T_c)^2-1.9918xl0^{-7}(T_c)^3$

Chapter 5

$$U_c = \begin{pmatrix} 1719.4 + 3.2063\,(T_c) + 1.5971x10^{-2}\,(T_c)^2 \\ -1.9918x10^{-4}\,(T_c)^3 \end{pmatrix} x10^{-3}$$

$$= \begin{pmatrix} 1719.4 + 3.2063\,(74.097) + 1.5971x10^{-2}\,(74.097)^2 \\ -1.9918x10^{-4}\,(74.097)^3 \end{pmatrix} x10^{-3}$$

$$= 1.96\ kW/m^2\ {}^oC$$

$$PCF = 3x10^{-7}\,(P_m)^2 - 0.0009\,(P_m) + 1.6101$$

$$TCF = 2x10^{-8}\,(T_{ev})^2 - 0.0006\,(T_{ev}) + 1.0047$$

$$\begin{aligned} PCF &= 3x10^{-7}\,(P_m)^2 - 0.0009\,(P_m) + 1.6101 \\ &= 3x10^{-7}\,(750)^2 - 0.0009\,(750) + 1.6101 \\ &= 1.104 \end{aligned}$$

$$\begin{aligned} TCF &= 2x10^{-8}\,(T_c)^2 - 0.0006\,(T_c) + 1.0047 \\ &= 2x10^{-8}\,(74.097)^2 - 0.0006\,(74.097) + 1.0047 = 0.96 \end{aligned}$$

$$\Downarrow$$

$$\begin{aligned} Ra &= 0.296\,\frac{(P_s)^{1.19}}{(P_{ev})^{1.04}}\left(\frac{P_m}{P_{ev}}\right)^{0.015}\left(\frac{PCF}{TCF}\right) \\ &= 0.296\,\frac{(92.75)^{1.19}}{(37.1)^{1.04}}\left(\frac{750}{37.1}\right)^{0.015}\left(\frac{1.104}{0.96}\right) = 1.82 \end{aligned}$$

Using Equation 1

$$Q_e = M_f\, C_p\, (T_b - T_f) + M_d\, \lambda_v = (M_m + M_{ev})\, \lambda_s$$

$$\lambda_v = 2501.897149 - 2.407064037\, T_v + 1.192217\text{x}10^{-3}\, T^2_v - 1.5863\text{x}10^{-5}\, T^3_v$$

$$\begin{aligned} &= 2501.897149 - 2.407064037(74.097) \\ &\qquad + 1.192217\text{x}10^{-3}\,(74.097)^2 - 1.5863\text{x}10^{-5}\,(74.097)^3 \\ &= 2323.6 \text{ kJ/kg} \end{aligned}$$

$$\lambda_s = 2501.897149 - 2.407064037\, T_s + 1.192217\text{x}10^{-3}\, T^2_s - 1.5863\text{x}10^{-5}\, T^3_s$$

$$\begin{aligned} &= 2501.897149 - 2.407064037(97.6) \\ &\qquad + 1.192217\text{x}10^{-3}\,(97.6)^2 - 1.5863\text{x}10^{-5}\,(97.6)^3 \\ &= 2263.6 \text{ kJ/kg} \end{aligned}$$

$$(2.5)(3.86)\,(75-70) + (1)(2326.37) = (M_m + M_{ev})(2266.76) \rightsquigarrow$$

$$\mathbf{Ra} = \mathbf{1.82} = \frac{M_m}{M_{ev}}$$

$M_m = 0.67$ kg/s.

$M_{ev} = 0.37$ kg/s

The cooling water flow rate is obtained from the condenser balance:

$$(M_f + M_{cw})\,(Cp)\,(T_f - T_{cw}) = (\eta)\,(M_d - M_{ev})\,(\lambda_c)$$
$$(2.5 + M_{cw})\,(3.97)\,(70-25) = (0.9)\,(1-0.37)\,(2323.6)$$

$\Rightarrow$ $M_{cw} = 4.83$ kg/s.

The evaporator and condenser loads are obtained from evaporator and condenser balance respectively

$\Rightarrow$

$$Q_e = (M_m + M_{ev})(\lambda_s) = (0.67 + 0.37)(2263.6) = 2354.1\ \text{kW}$$

$$Q_c = \eta\,(M_c)(\lambda_c) = (0.9)(0.63)(2323.6) = 1317.5\ \text{kW}$$

The evaporator A_e and condenser areas A_c are then calculated from Eq 5, Eq6 and Eq7, results in:

$$A_e = \frac{Q_e}{U_e(T_s - T_b)} = \frac{2354.1}{(2.62)(97.6-75)} = 39.8\ \text{m}^2$$

$$(\text{LMTD})_c = \frac{(T_f - T_{cw})}{\ln\dfrac{T_c - T_{cw}}{T_c - T_f}} = \frac{70-25}{\ln\dfrac{73.01-25}{73.01-70}} = 16.25\ ^{o}\text{C}$$

$$A_c = \frac{Q_c}{(U_c)(\text{LMTD})_c} = \frac{1317.5}{(1.96)(16.25)} = 41.18\ \text{m}^2$$

- The performance ratio is calculated from: $\boxed{PR = M_d/M_m = 1/0.675 = 1.48}$

- The specific cooling water flow rate sM_{cw} $\boxed{sM_{cw} = M_{cw}/M_d = 4.83}$

- The specific area sA $sA = (Ae+Ac)/M_d = 44.47 + 41.18 = 85.65\ m^2/(kg/s)$.

***See Example (2) In the text book page 67

PERFORMANCE CHARTS

System performance is presented in terms of variations in the system **design parameters** as a function of:

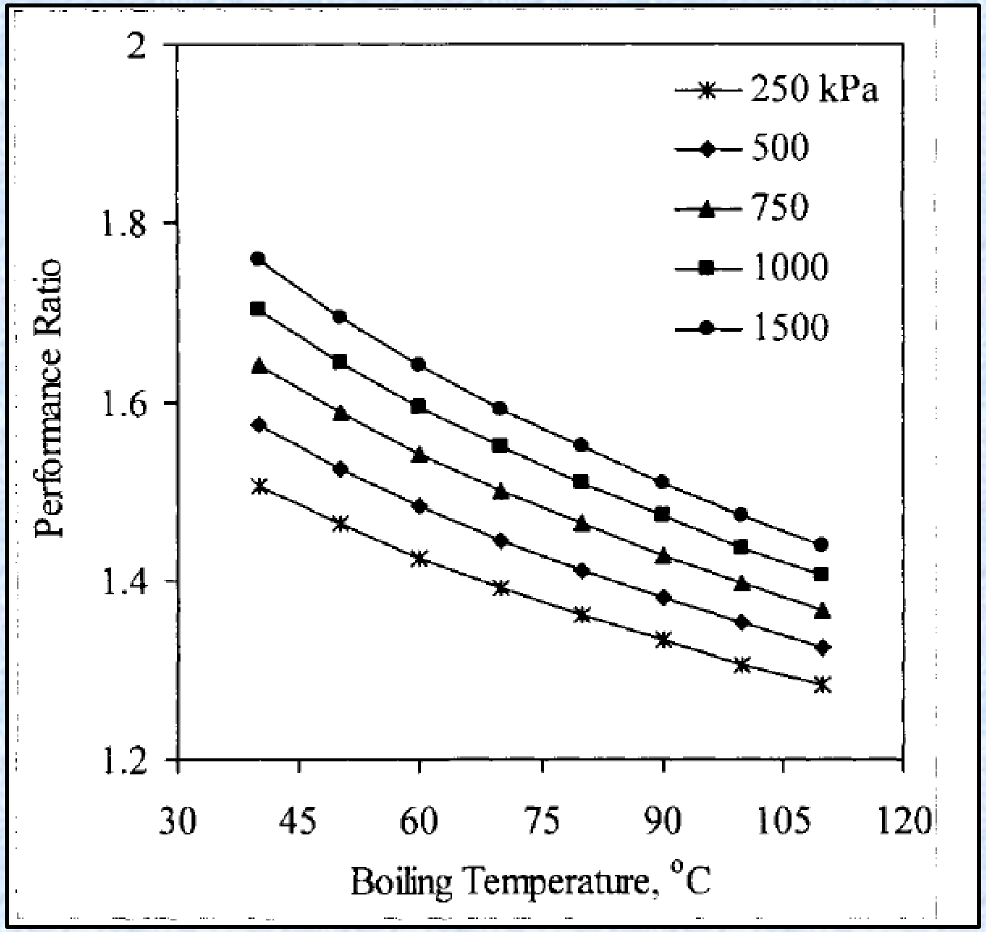
1. The boiling temperature, T_b
2. The compression ratio, Cr
3. Pressure of the motive steam, P_m

The system parameters include variations in :

1. The performance ratio, PR
2. The specific heat transfer area, sA
3. The specific cooling water flow rate, sM_{cw}

1. Variation in the system performance ratio as a function of boiling point T_b and the motive steam pressure P_m

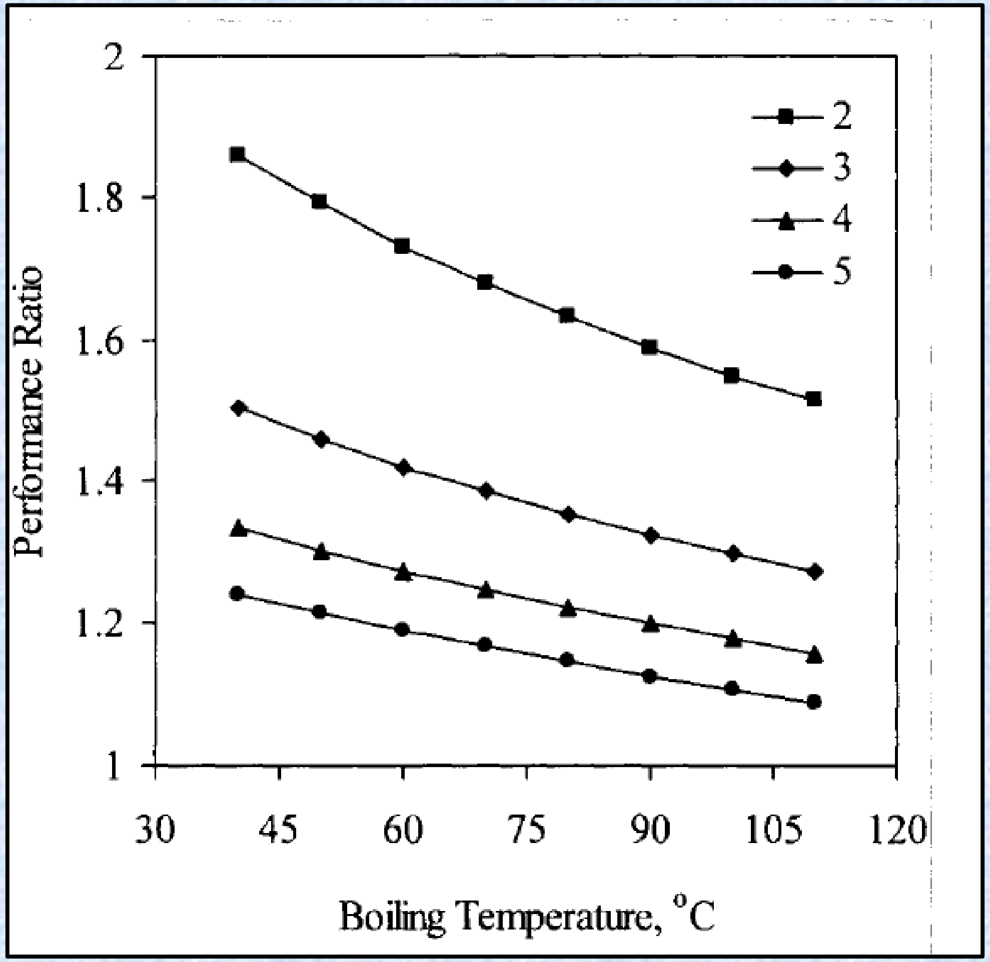
- The higher performance ratio values are obtained at :
  a) Low boiling temperatures
  b) High motive steam pressures
- At low boiling temperatures, the amount of motive steam consumed to compress the entrained vapor is low. This is because of the small increase in the vapor pressure at low temperatures.
- For example, the vapor pressure between 55 and 60 °C increases from 15.8 to 19.9 kPa is 26.5%. On the other hand, the vapor pressure increases from 70.14 to 84.55 kPa as the temperature increases from 90 to 95 °C



2. Variation in the system performance ratio as a function of boiling point T_b and the compression ratio Cr

- The higher performance ratio values are obtained at :
  a) Low boiling temperatures
  b) Low compression ratio

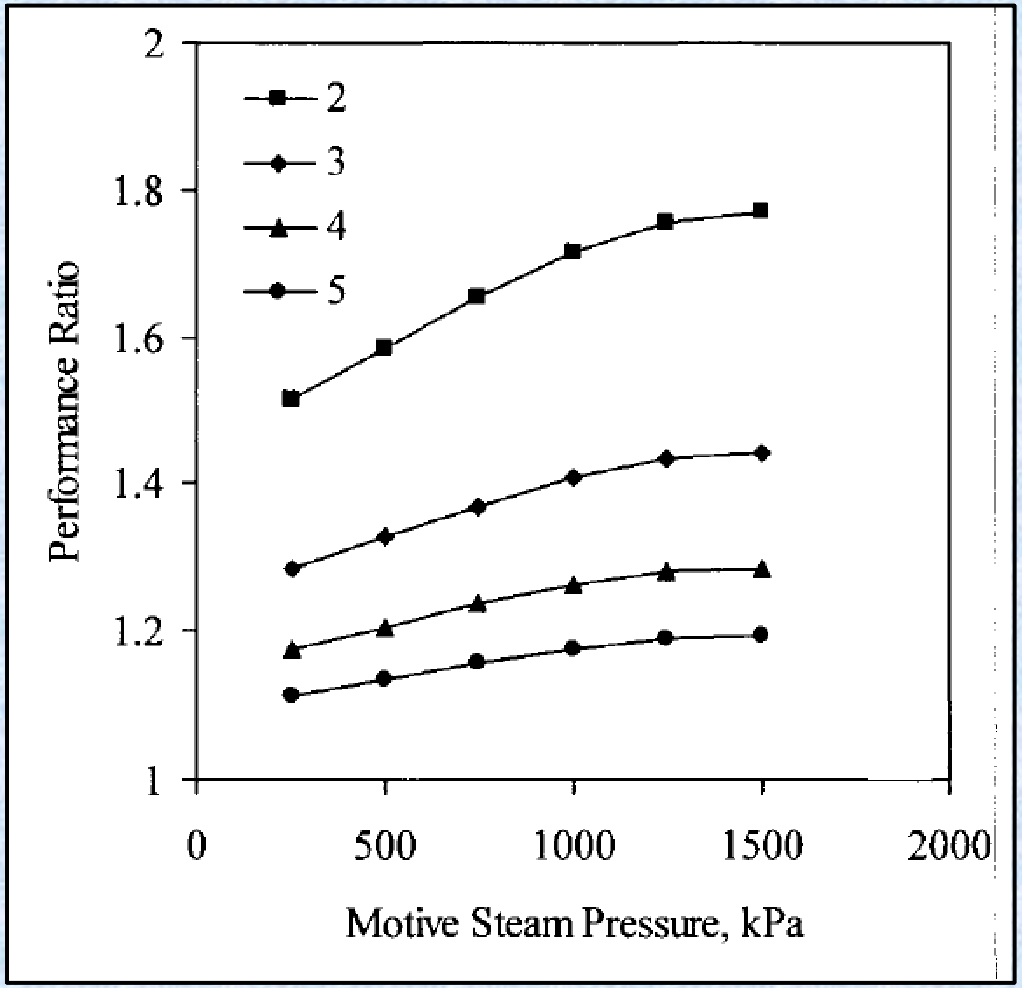
- At low compression ratios, the amount of motive steam consumed to compress the entrained vapor is small. Therefore, the system performance ratio is higher.



3. Variation in the system performance ratio as a function of motive steam pressure P_m and the compression ratio Cr

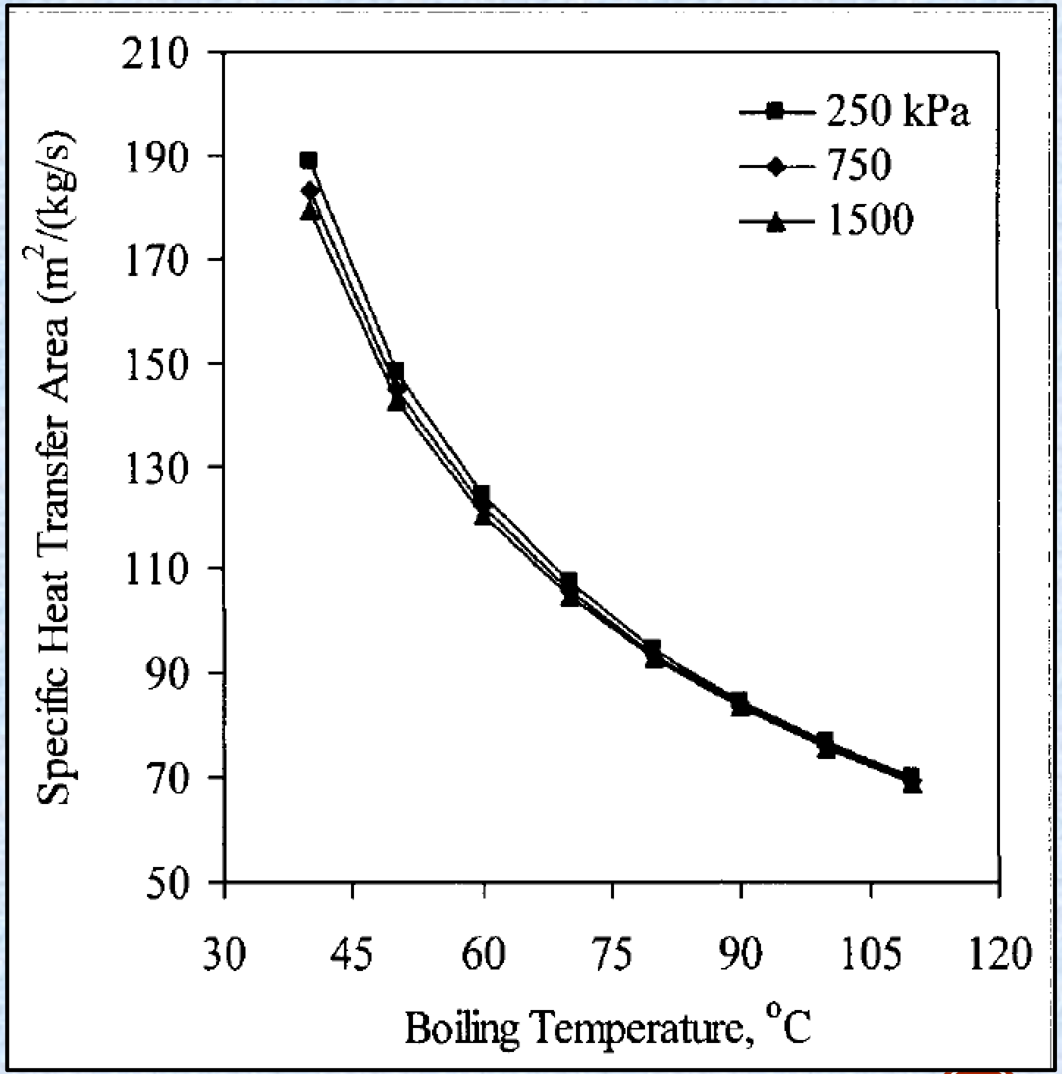
- The sensitivity of the performance ratio to variations in the motive steam pressure is **less pronounced** than those found as a function of the boiling temperature and the compression

- Note: As the pressure of the motive steam is increased its latent is lower. Therefore, to maintain constant latent heat in the compressed vapor, it is necessary to entrain larger amounts of the vapor leaving the evaporator. This reduces the consumed amount of motive steam.



4. Variations in the specific heat transfer area (A) as a function of the boiling temperature T_b and the motive steam pressure P_m

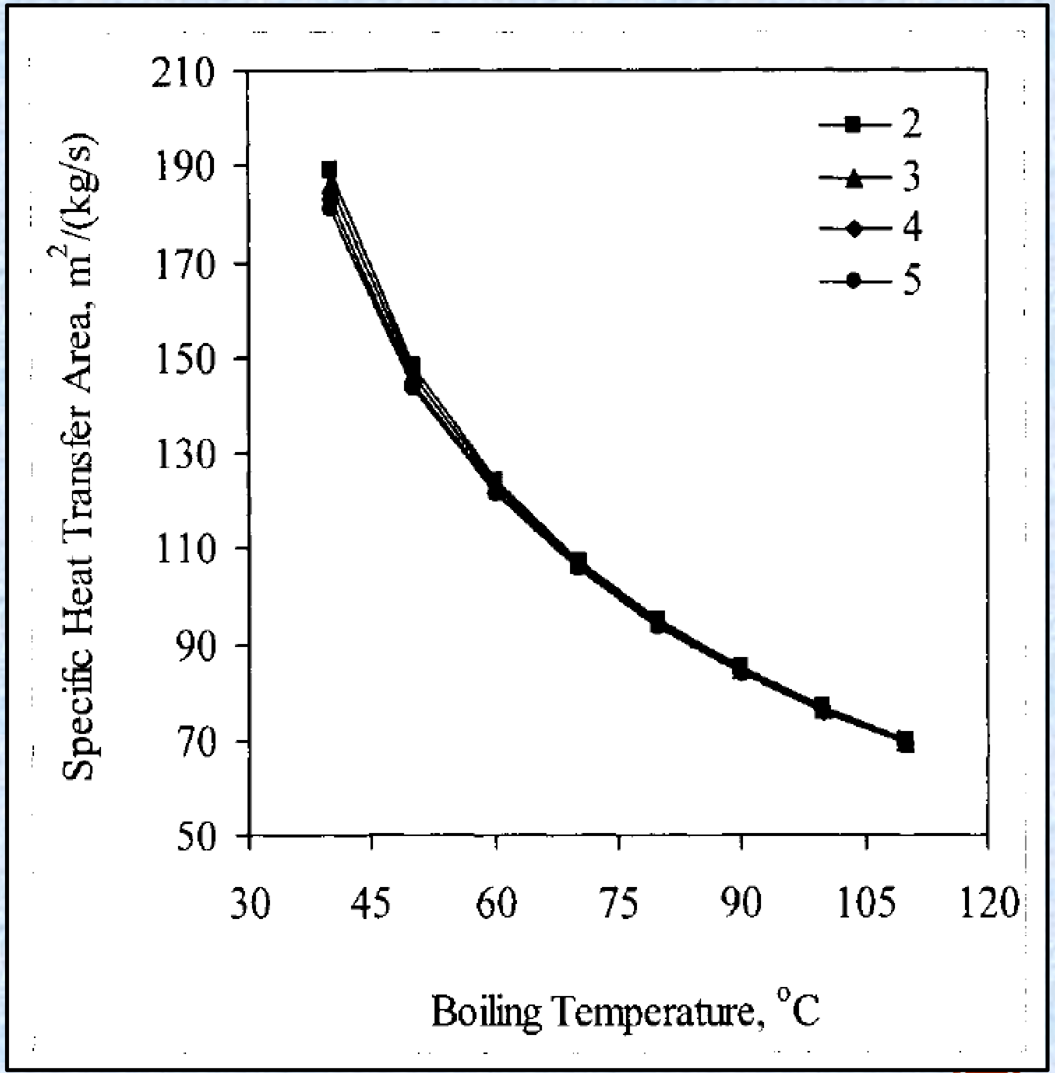
- The specific heat transfer area **decreases drastically** as the boiling temperature is **increased** and this is due to increase in the overall heat transfer coefficient in the evaporator and the condenser at high boiling temperatures.



5. Variations in the specific heat transfer area (A) as a function of the boiling temperature T_b and the compression ratio Cr

- The temperature of the compressed vapor is increased as the compression ratio is elevated. The increase in the temperature of the compressed vapor enhances the rates of heat transfer. This is caused by the increase of the driving force for heat transfer across the evaporator, which is measured by the difference of T_s-T_b.

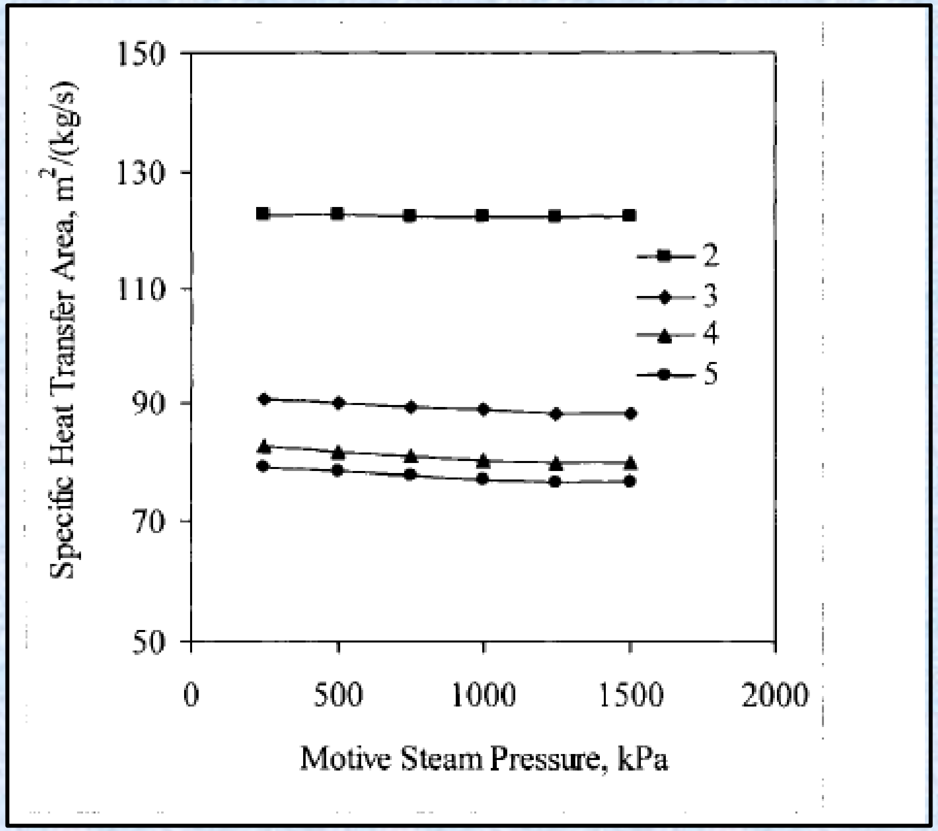
- As a result, the evaporator heat transfer area is reduced at higher compression ratios. Irrespective of this, the heat transfer area increases in the condenser.



6. Variations in the specific heat transfer area (A) as a function of the motive steam pressure P_m and the compression ratio Cr.

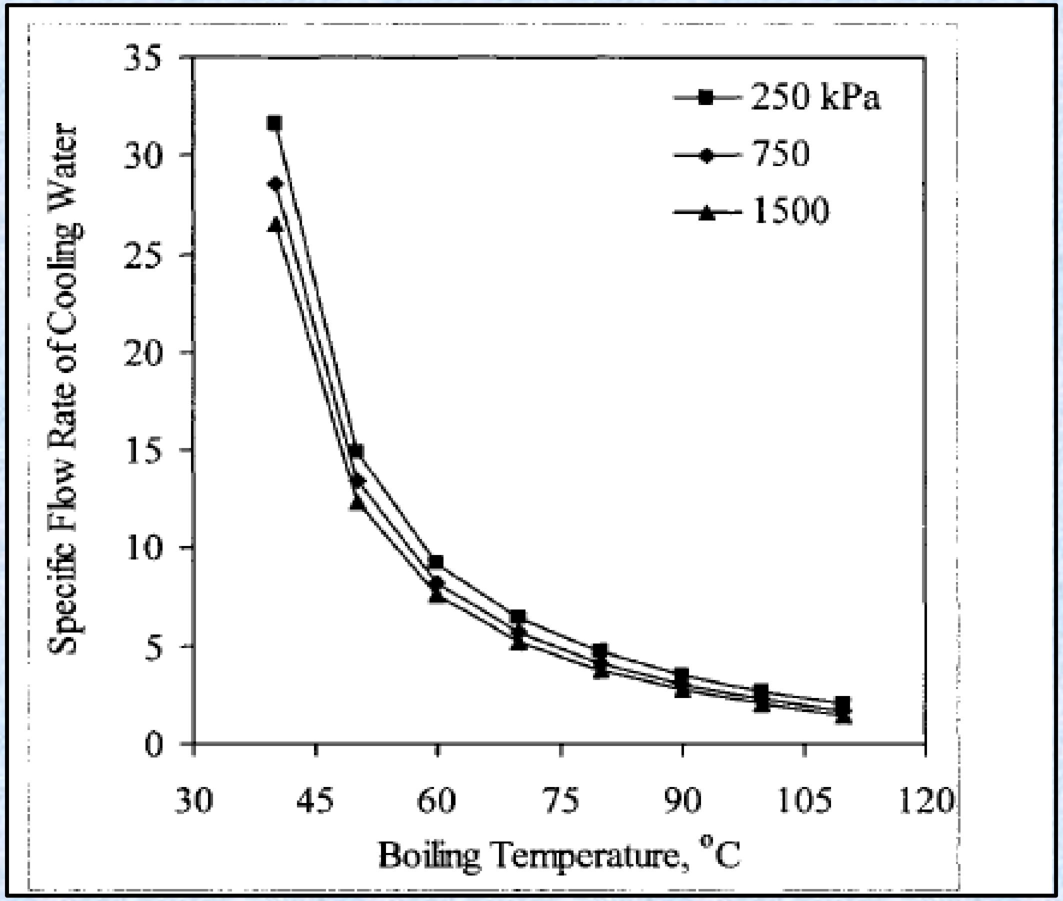
- The specific heat transfer area is insensitive to variations in the motive steam pressure. This is because of limited variations in the overall heat transfer coefficient in the evaporator and condenser as well as the amount of entrained vapor.

- These results are similar to those obtained for the variations in the system performance ratio. (Variations in the performance ratio as a function of the boiling temperature and the motive steam pressure)



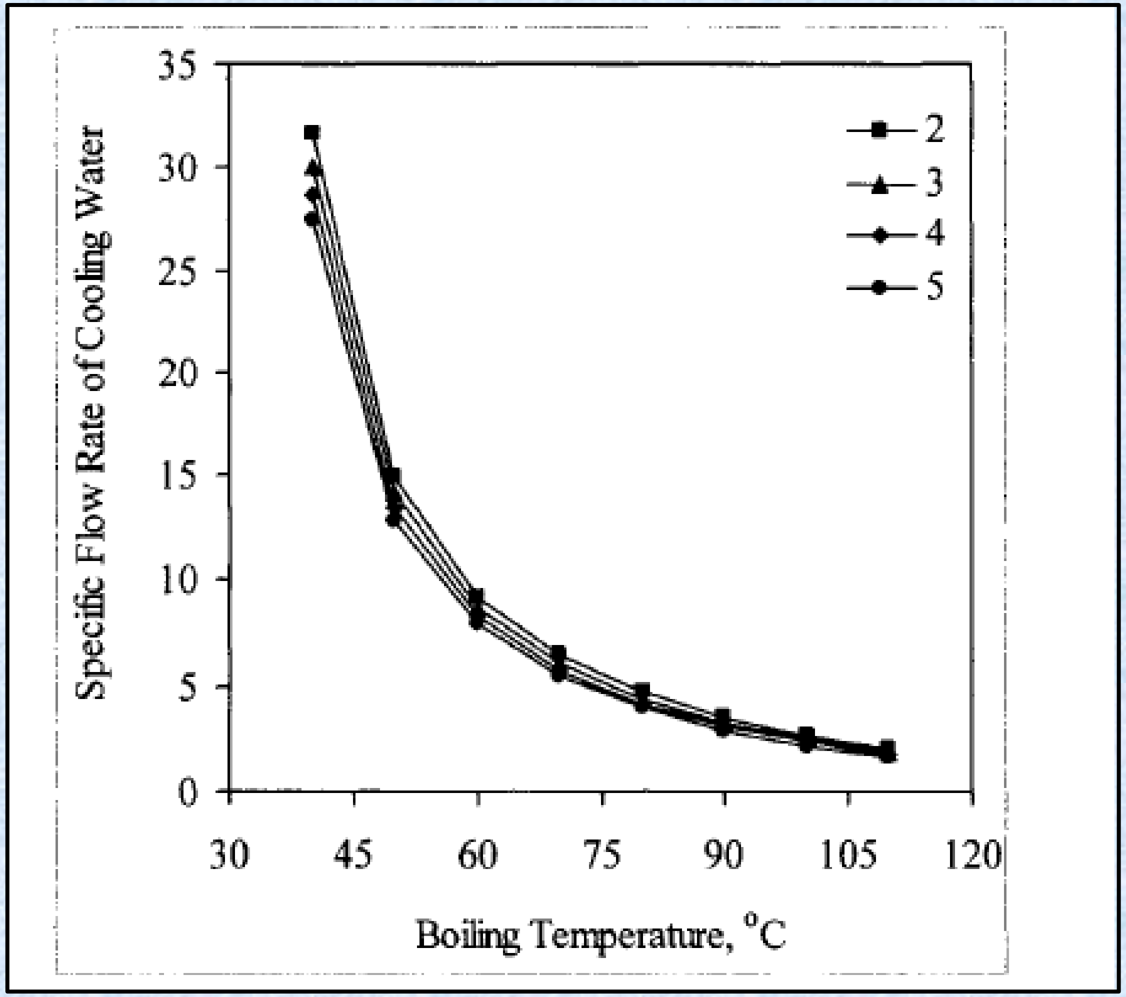
7. Variations in the specific cooling water flow sM_{cw} as a function of boiling temperature T_b and the motive steam pressure P_m

- The specific cooling water flow rate is highly sensitive to variations in the boiling temperature and the motive steam pressure, this is caused by large increase in the overall heat transfer coefficient in the evaporator and condenser.



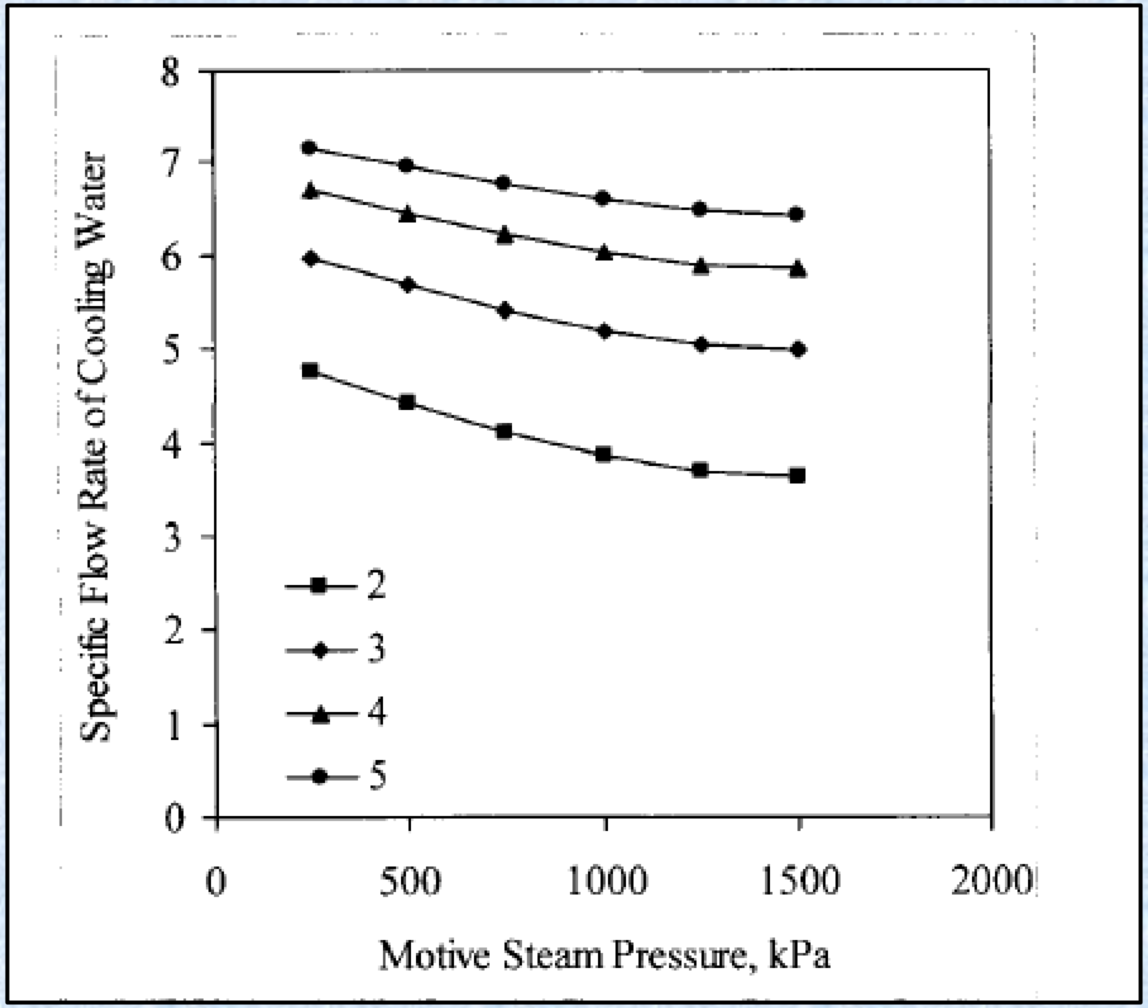
8. Variations in the specific flow rate of cooling water sM_{cw} as a function of boiling temperature T_b and the compression ratio Cr

- The specific cooling water flow rate is highly sensitive to variations in the boiling temperature and the compression ratio , this is caused by large increase in the overall heat transfer coefficient in the evaporator and condenser.



9. Variations in the specific flow rate of cooling water sM_{cw} as a function of the motive steam pressure P_m and the compression ratio Cr.

- Limited sensitivity in the specific cooling water flow rate is found upon the increase in the motive steam pressure.



END OF CHAPTER 6

University of Jordan

Faculty of Engineering and Technology

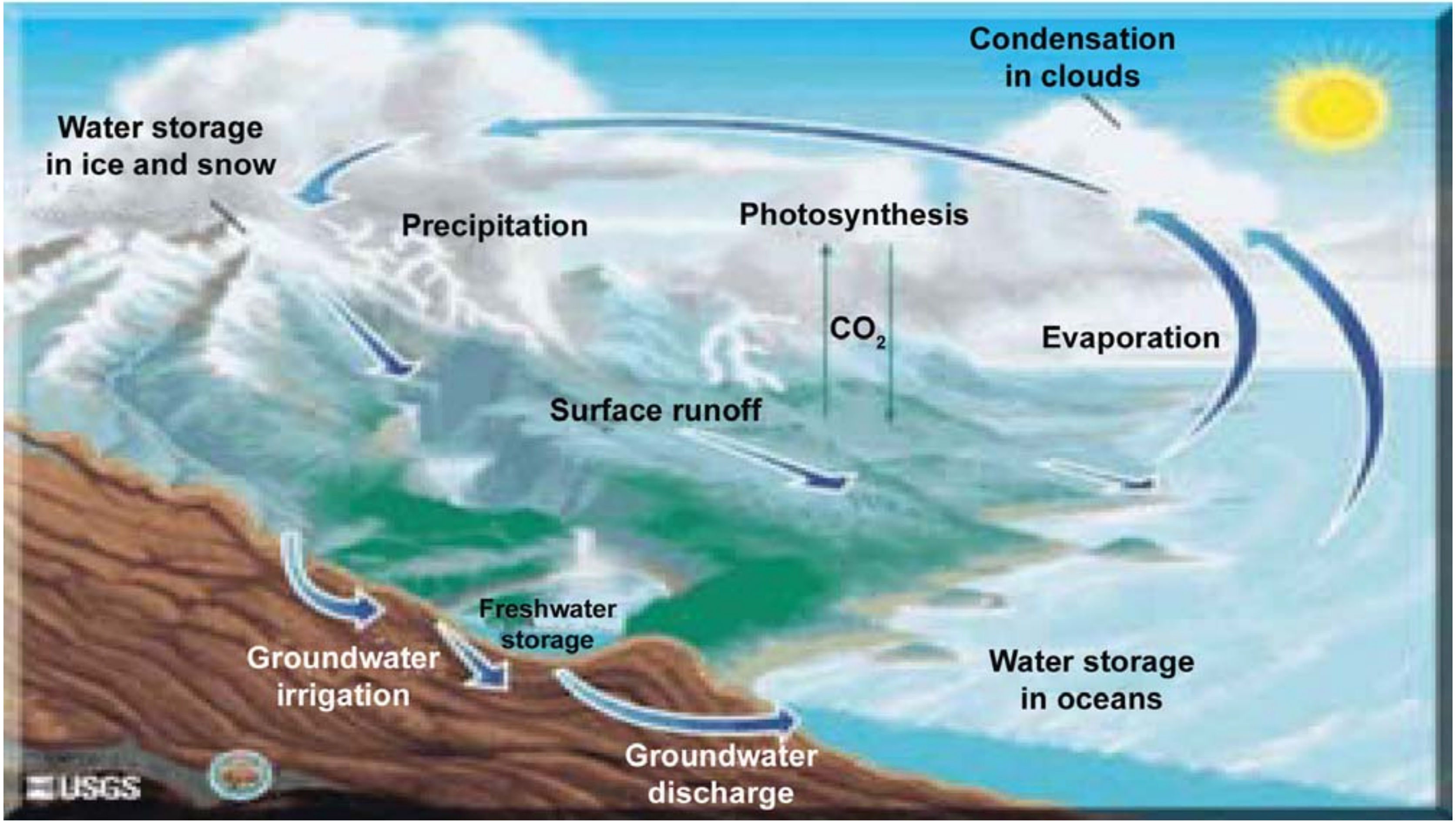
Chemical Engineering Department

Desalination

Dr. Mohammed Rasool Qtaishat

Thermal Desalination Processes

The principle is simple and it is an imitation of what is happening in nature



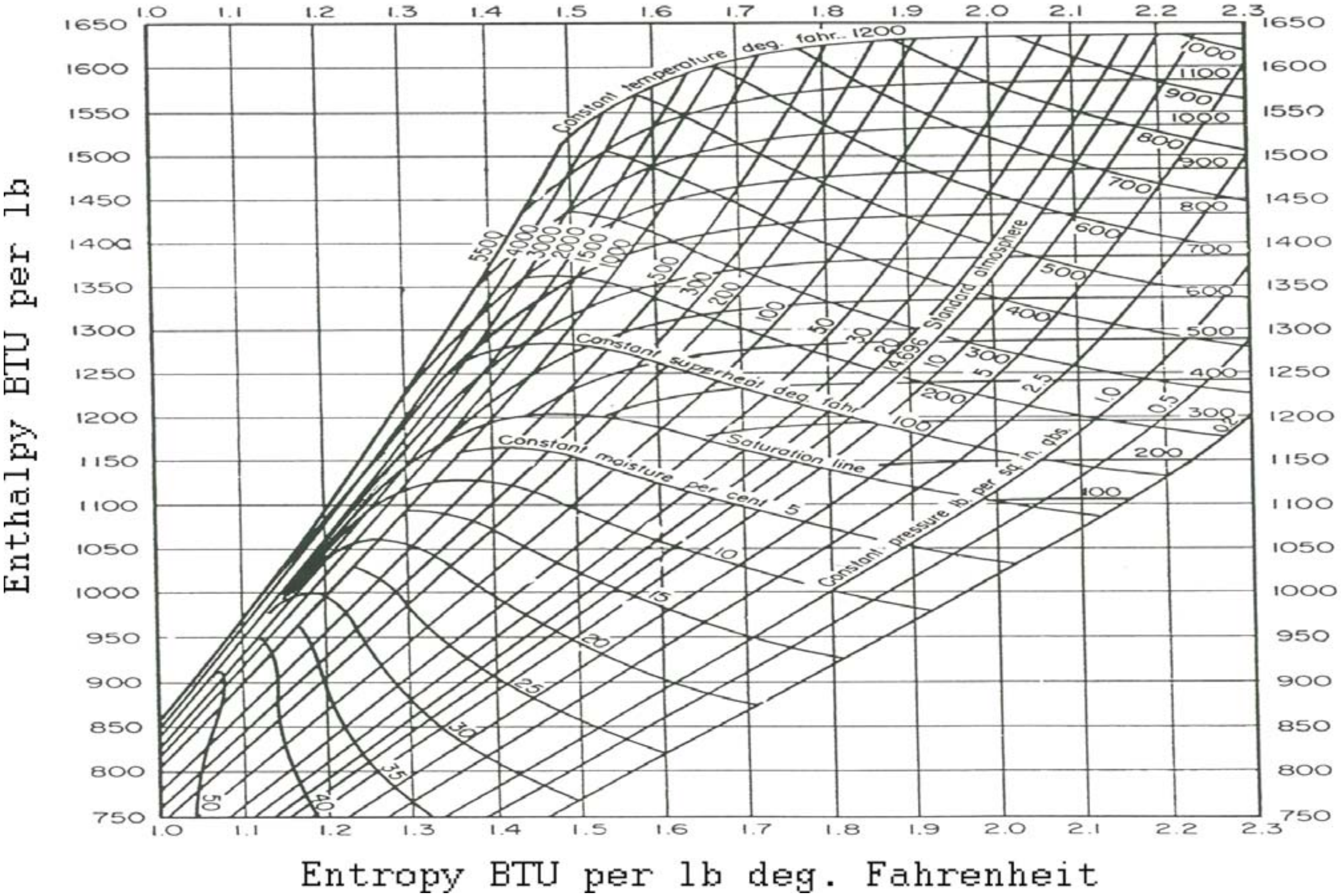
Modern Thermodynamics

- Ancient process were based on a single evaporation and condensation.

- Modern thermal desalination technologies are based on the repetition of evaporation and flashing phenomena at lower temperature and pressure which were understood with the discovery of modern thermodynamics.

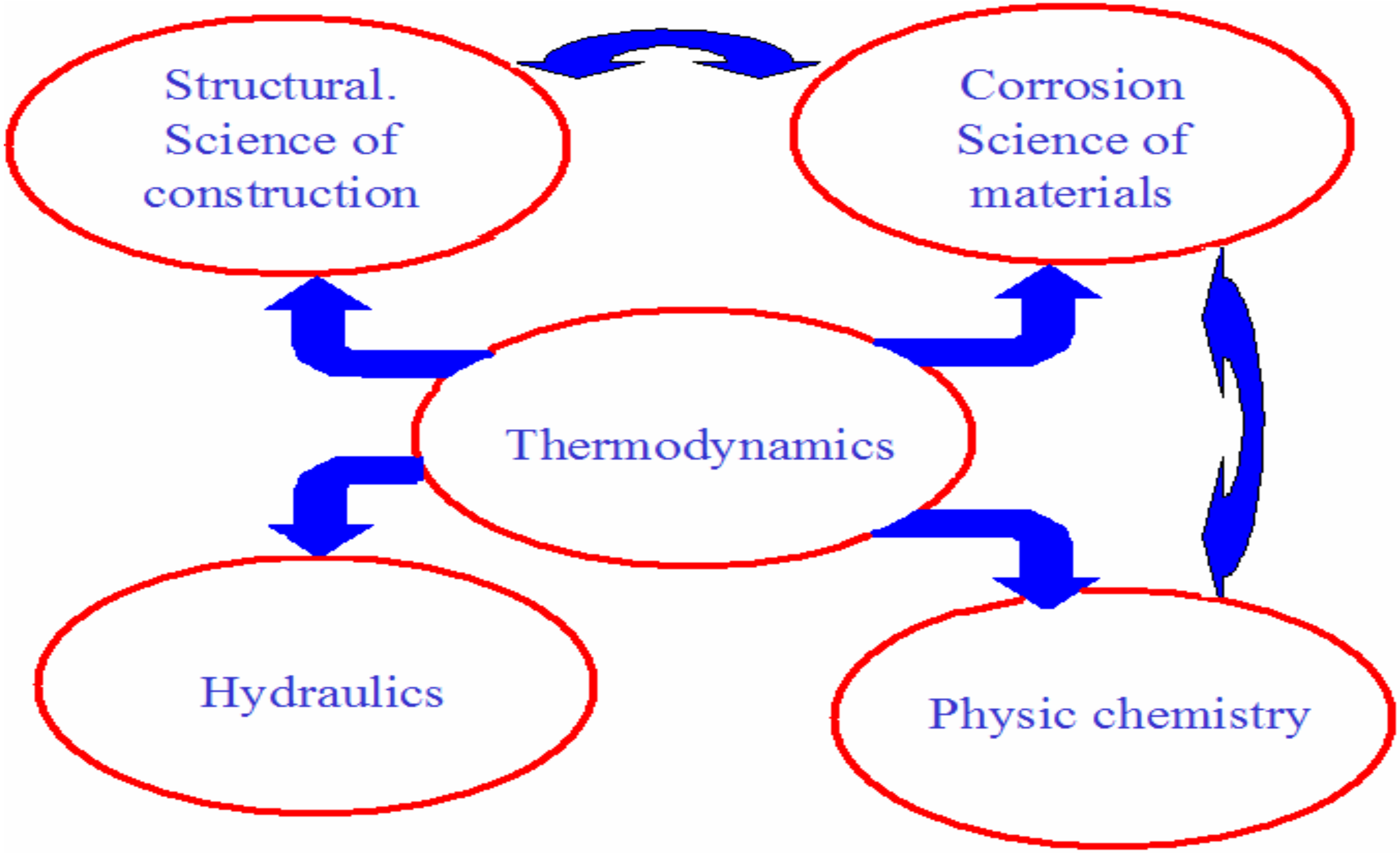
- Mollier diagram: ***An Important Tool***, it gives Relationships between enthalpy entropy temperature and pressure of steam in saturated and overheated conditions.

Modern Thermodynamics



Thermal Desalination: Introduction

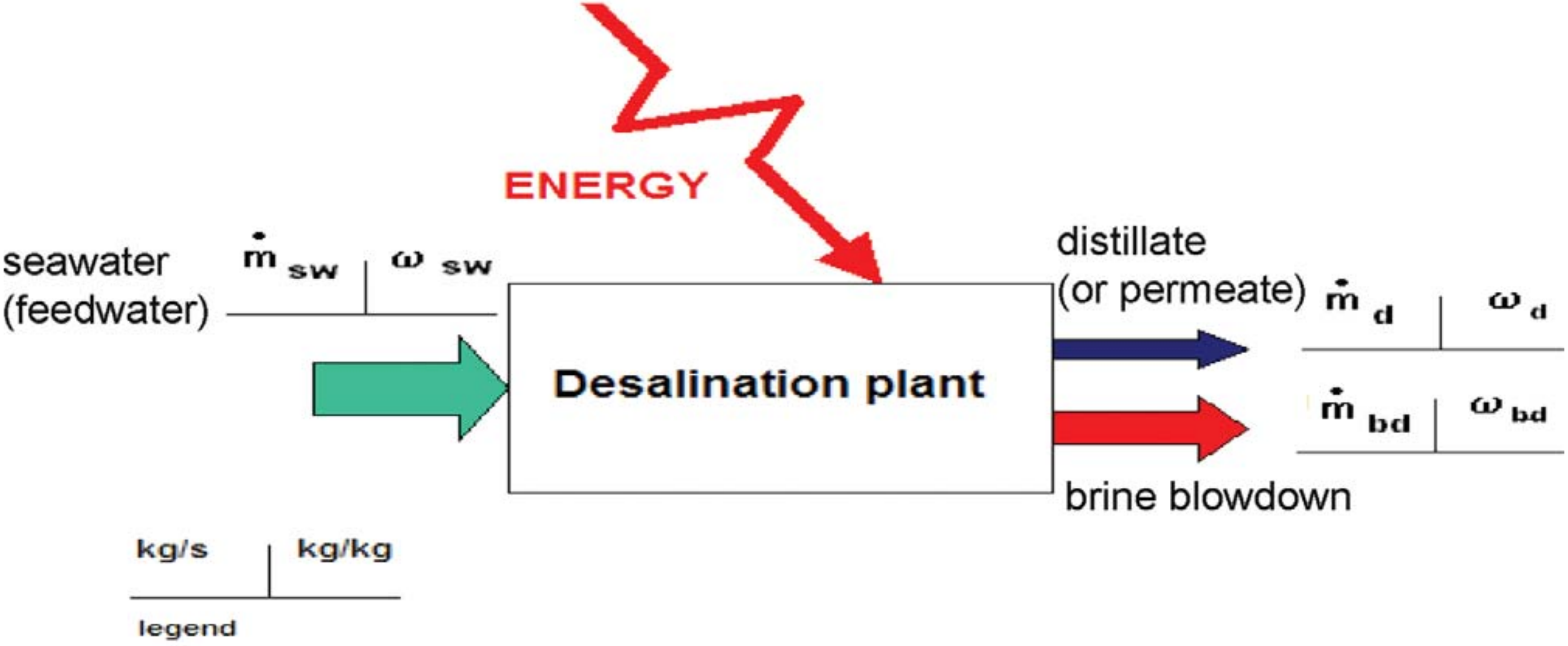
Modern thermal desalination is a technology which involve several disciplines and has reached a high degree of sophistication.



Thermal Desalination: Introduction

- Basic heat and mass balance
- Energy input
- Performance ratio
- Multiple cells concept

Desalination plant basic mass balances



- Regardless of the type of process adopted desalination transforms seawater into concentrated brine and distillate.

Mass Balance Relationships

1) Mass conservation (overall mass balance)

$$m_{sw} = m_{bd} + m_d$$

2) Salt conservation (overall salt balance)

$$m_{sw} \times \omega_{sw} = m_{bd} \times \omega_{bd} + m_d \times \omega_d$$

ω — Salt concentration (kg/kg)

m — Mass flow rate (kg/s)

3) Concentration factor — ratio between blowdown and seawater salt concentration: $Cf_{bd} = \dfrac{\omega_{bd}}{\omega_{sw}}$

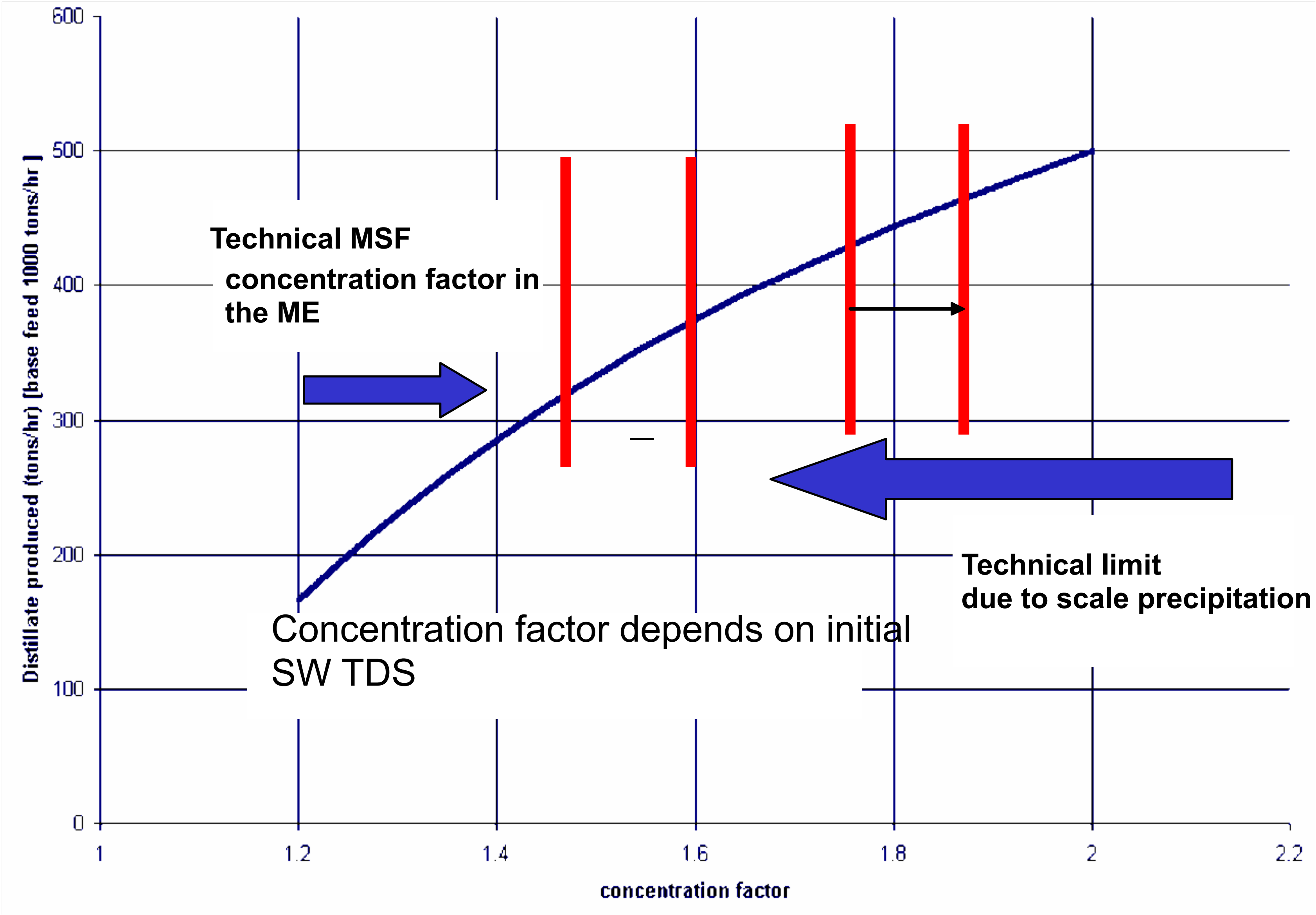
Rearranging, gives $m_d = m_{sw}\left[1 - \dfrac{1}{Cf_{bd}}\right]$,,,, PROVE.

This formula is valid for all Desalination processes including the membrane processes.

Conc. Factor-Production Ratio

- **Theoretically it would be best to concentrate as much as possible. However it is not possible to concentrate seawater–blowdown above a certain limit.**
- **The following constraints occur:**
  - Scale precipitation in tube bundle are more frequent the more salt is concentrated
  - Experience with all systems indicated need for scale control
  - Hot brines easily reached saturation with inorganic species ($Mg(OH)_2$, $CaCO_3$, $CaSO_4$, etc.)
  - Scale restricted flow paths, reduced heat transfer, caused out-ages.

Conc. Factor-Production Ratio



Technical MSF concentration factor in the ME

Technical limit due to scale precipitation

Concentration factor depends on initial SW TDS

A glance at other technologies (RO)

- **Typically the recovery rate for SWRO is 38–45%**

$$PR = \frac{m_p}{m_{sw}} \times 100\% = \frac{m_p}{m_p + m_{conc}} \times 100\%$$

$$PR = \frac{TDS_{con} - TDS_{SW}}{TDS_{con} - T\cancel{D}S_{Perm}}$$

$$Cf = \frac{1}{1 - PR} = \frac{1}{1 - 0.45} = 1.82$$

Example

- **Data Available:**
  - Sea water TDS: 45400 mg/L
  - Desired Distillate flow: 1200 ton/hour
  - Brine blowdown maximum admissible TDS = 58000 mg/L
- **Calculate:**
  - Brine blowdown flow rate
  - Sea water makeup requirement
  - Brine blowdown maximum admissible TDS = 58000 mg/L

- **Answers:** Blowdown flow rate: 4330 ton/h and make up requirement: 5530 ton/h

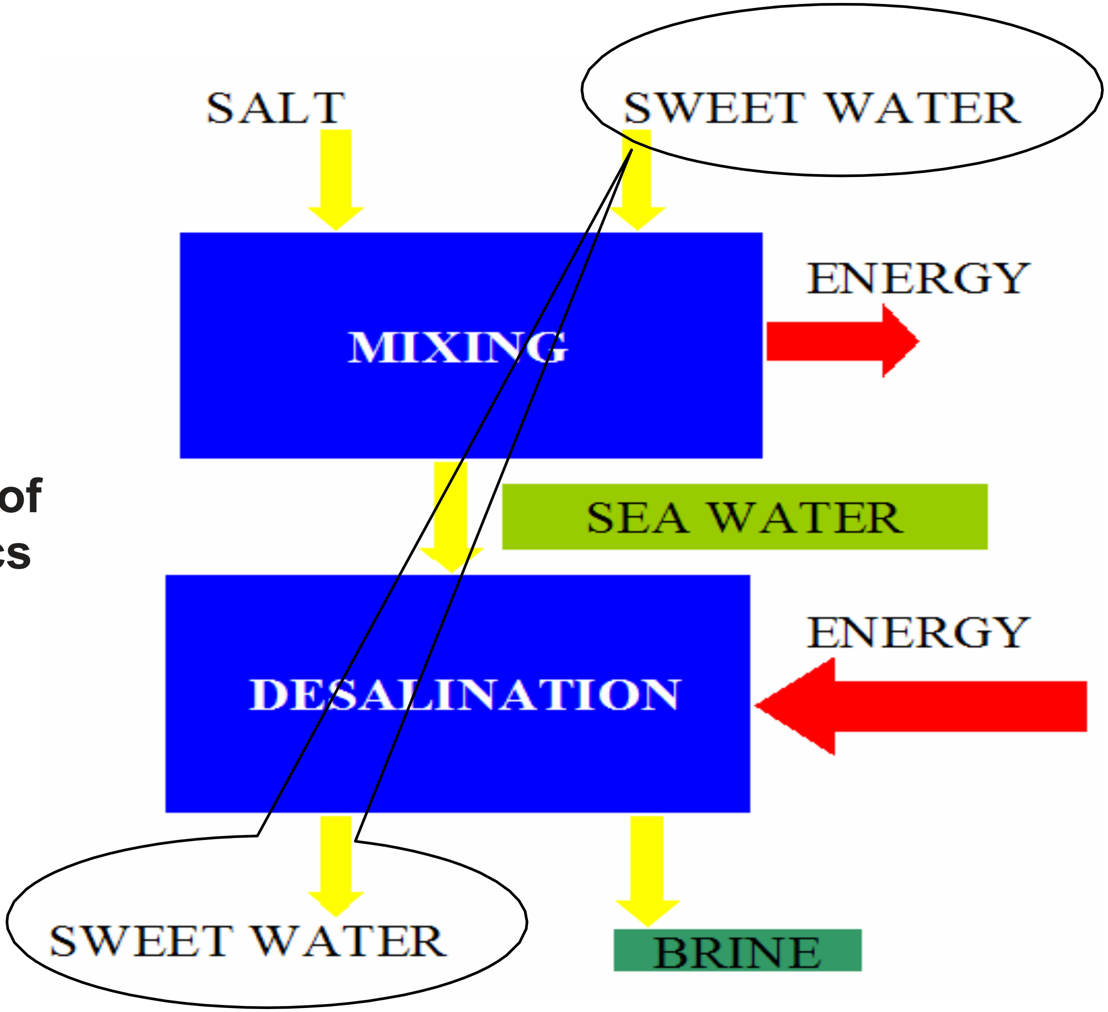
Energy Input Classifications

- **Evaporative processes:**
    - Evaporative processes use thermal energy to produce distilled pure water from sea or brackish water.
    - Evaporative processes rely on a phase change from liquid (in this case brine) to the vapour phase.
    - In this process only the water molecules pass to the vapour phase leaving the other constituents behind in the liquid.
    - The two dominating systems that have evolved are multi stage flash (MSF) and multiple effect distillation (MED).

Energy Input Classifications

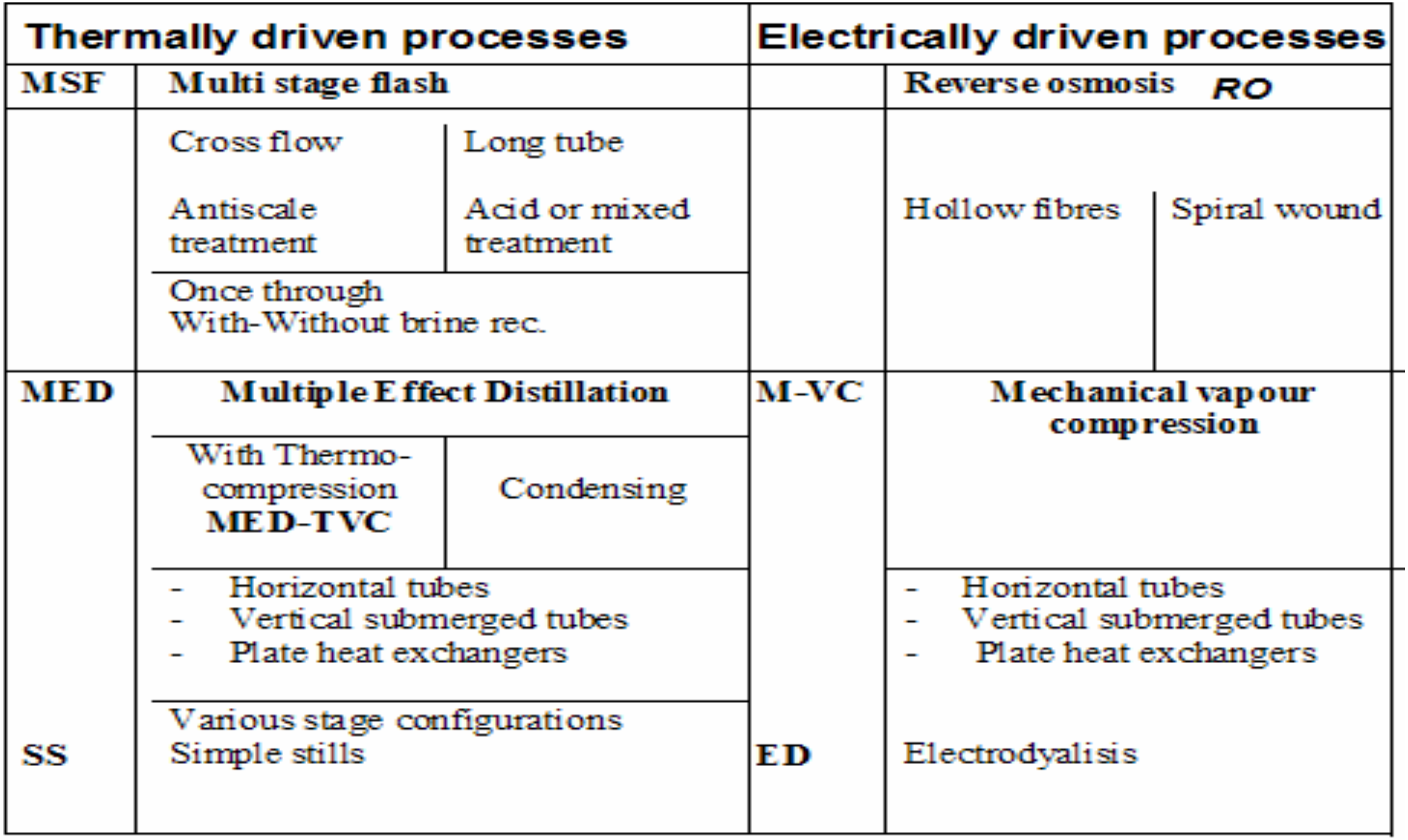
- **Membrane processes:**
  - In membrane processes electric energy is used to pump seawater (or brackish water) through a series of semi permeable membranes to obtain a low salinity permeate as a product.
  - Membrane processes do not rely on a phase change but on the size and transport mobility of water molecules through a permeable membrane.
  - For the separation of fresh water from seawater or brackish water many membrane separation processes are known such as reverse osmosis (RO).

Energy Balance in Desalination

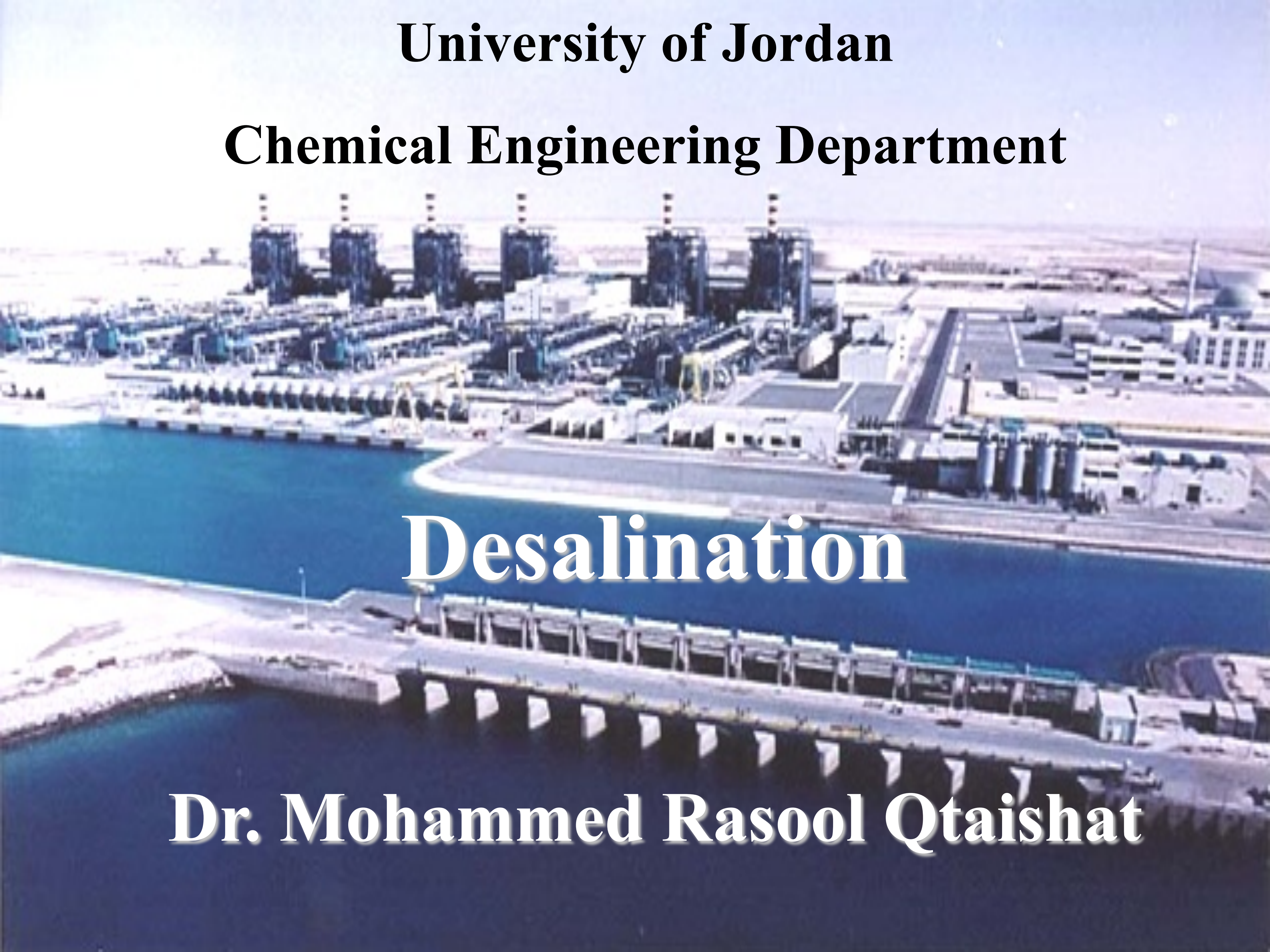


the second law of thermodynamics

OVERVIEW OF THERMALLY-DRIVEN TECHNOLOGIES



Thermally driven processes		Electrically driven processes	
MSF	**Multi stage flash**		**Reverse osmosis** ***RO***
	Cross flow | Long tube Antiscale treatment | Acid or mixed treatment Once through With-Without brine rec.		Hollow fibres | Spiral wound
MED	**Multiple Effect Distillation**	**M-VC**	**Mechanical vapour compression**
	With Thermo-compression **MED-TVC** | Condensing		
	- Horizontal tubes - Vertical submerged tubes - Plate heat exchangers		- Horizontal tubes - Vertical submerged tubes - Plate heat exchangers
SS	Various stage configurations Simple stills	**ED**	Electrodyalisis



University of Jordan

Chemical Engineering Department

Desalination

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Technologies and differences

Technologies and differences: some rule of thumb

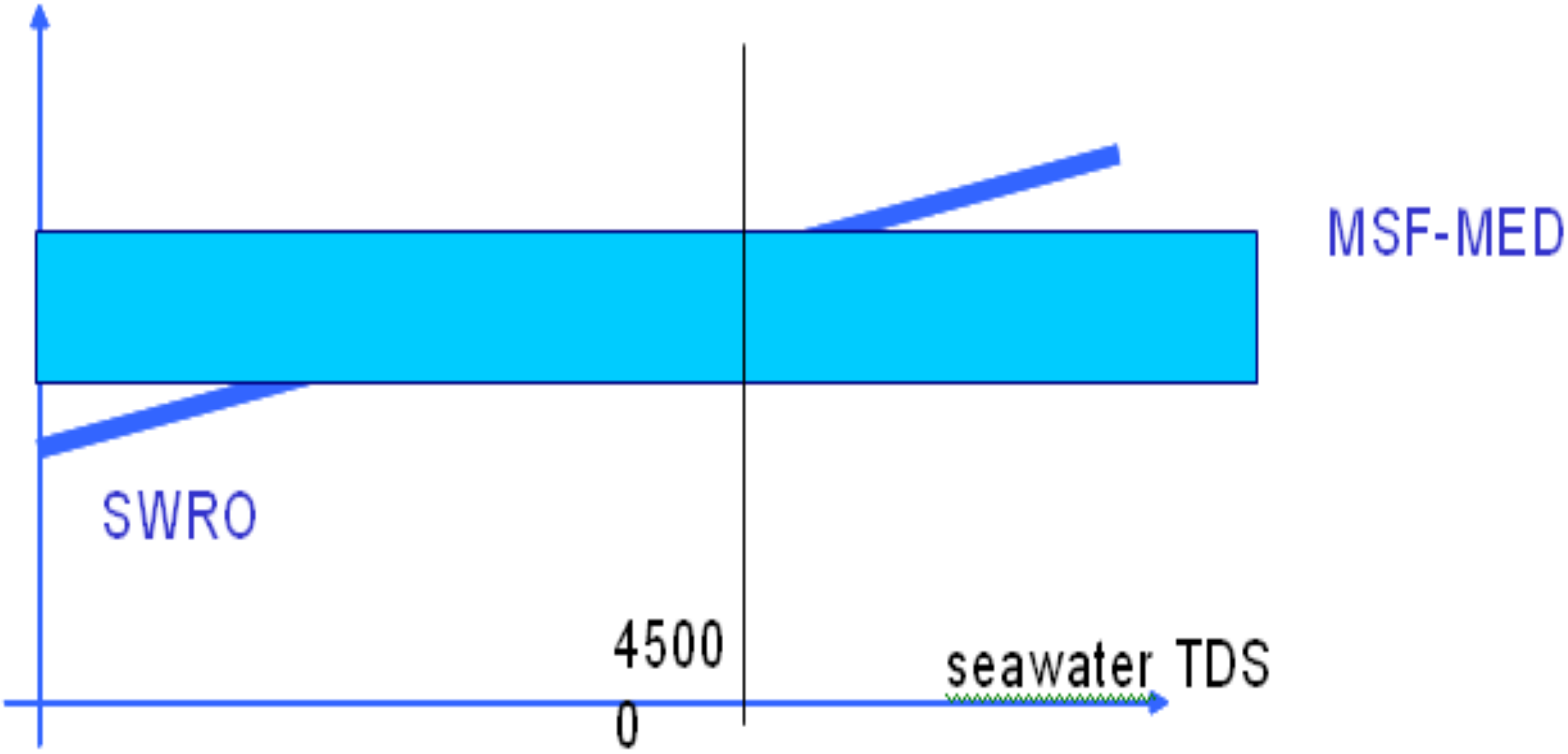
Cost effect: SWRO CAPEX and OPEX are greatly affected by:

- Seawater TDS
- Potable water quality

Cost effect: Thermal CAPEX and OPEX are only partially affected by:

- Seawater TDS
- And practically not affected by potable water quality up to TDS of 25 ppm

Technologies and differences

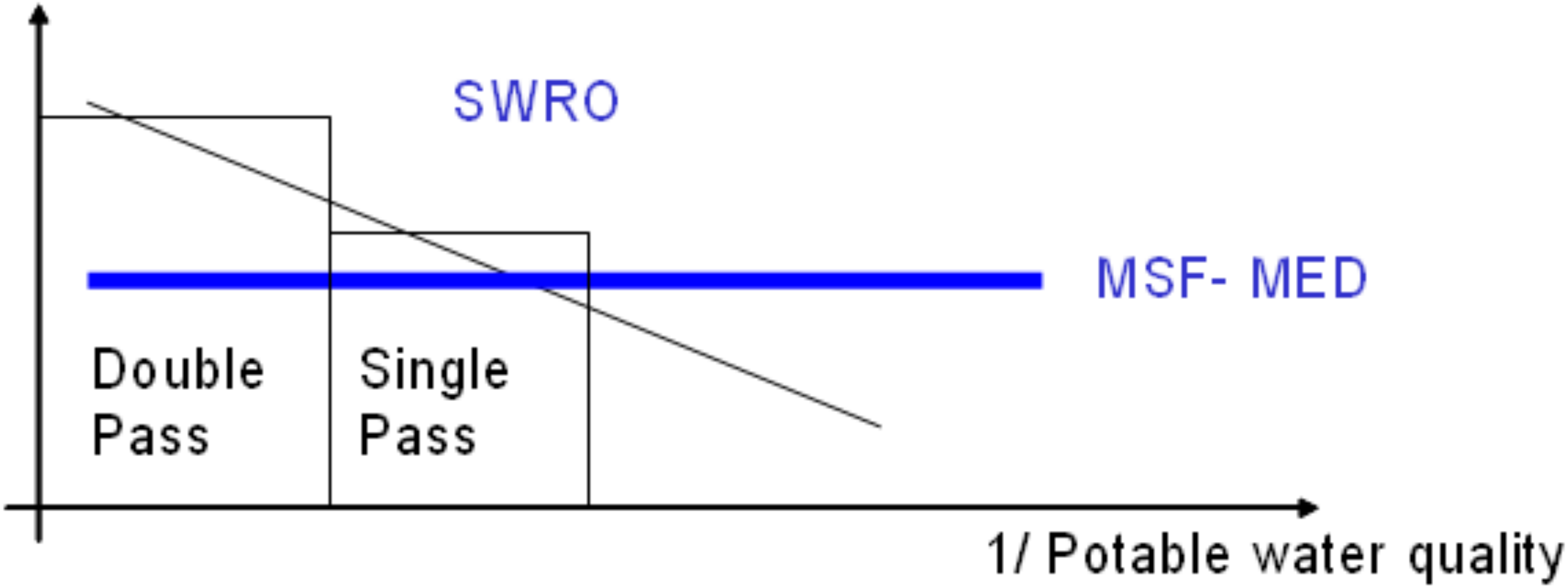


RR α 1/TDS

SW TDS → 5000 RR → 75%-85%

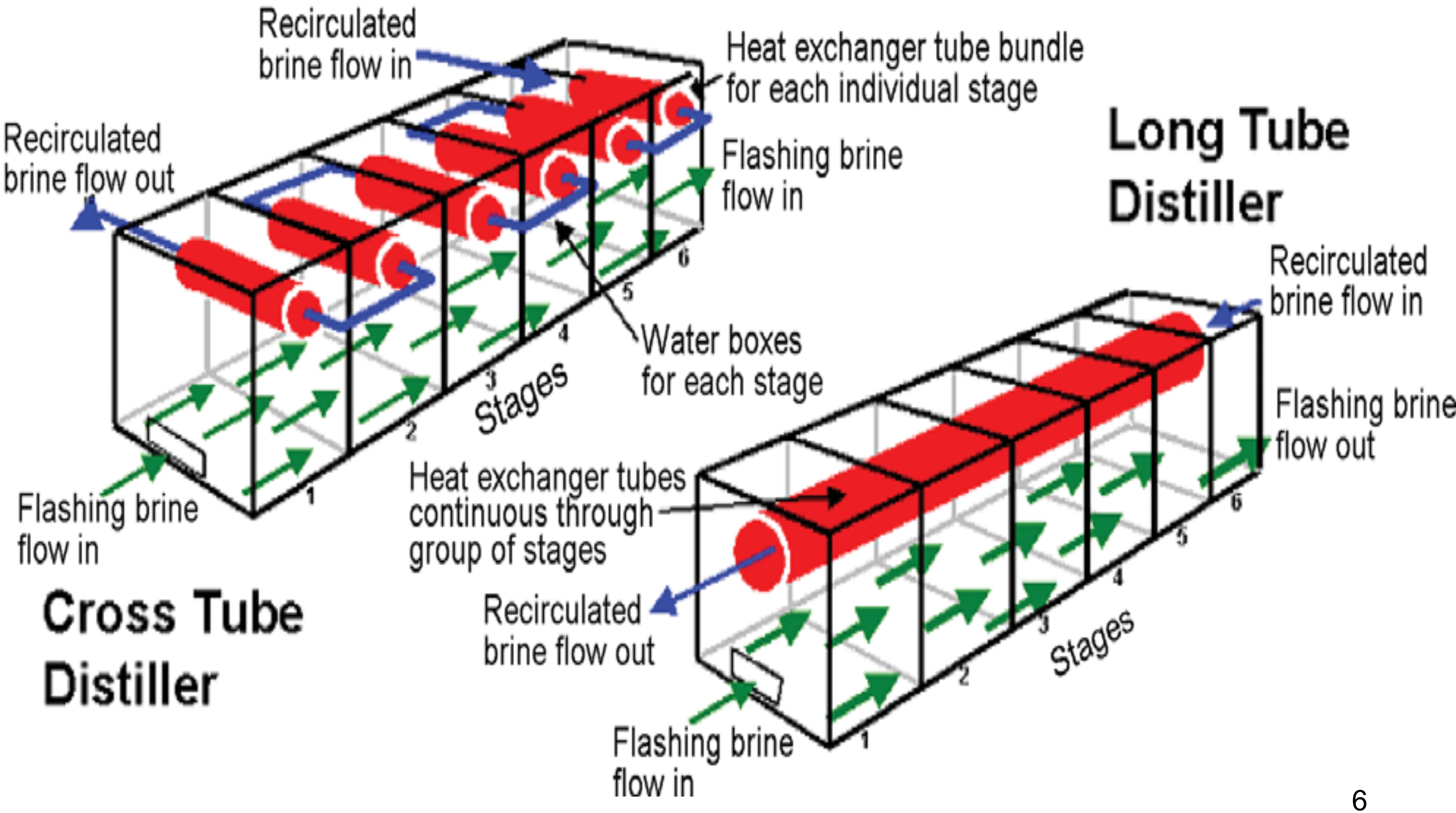
SW TDS → 45000 RR → 40-45%

Technologies and differences

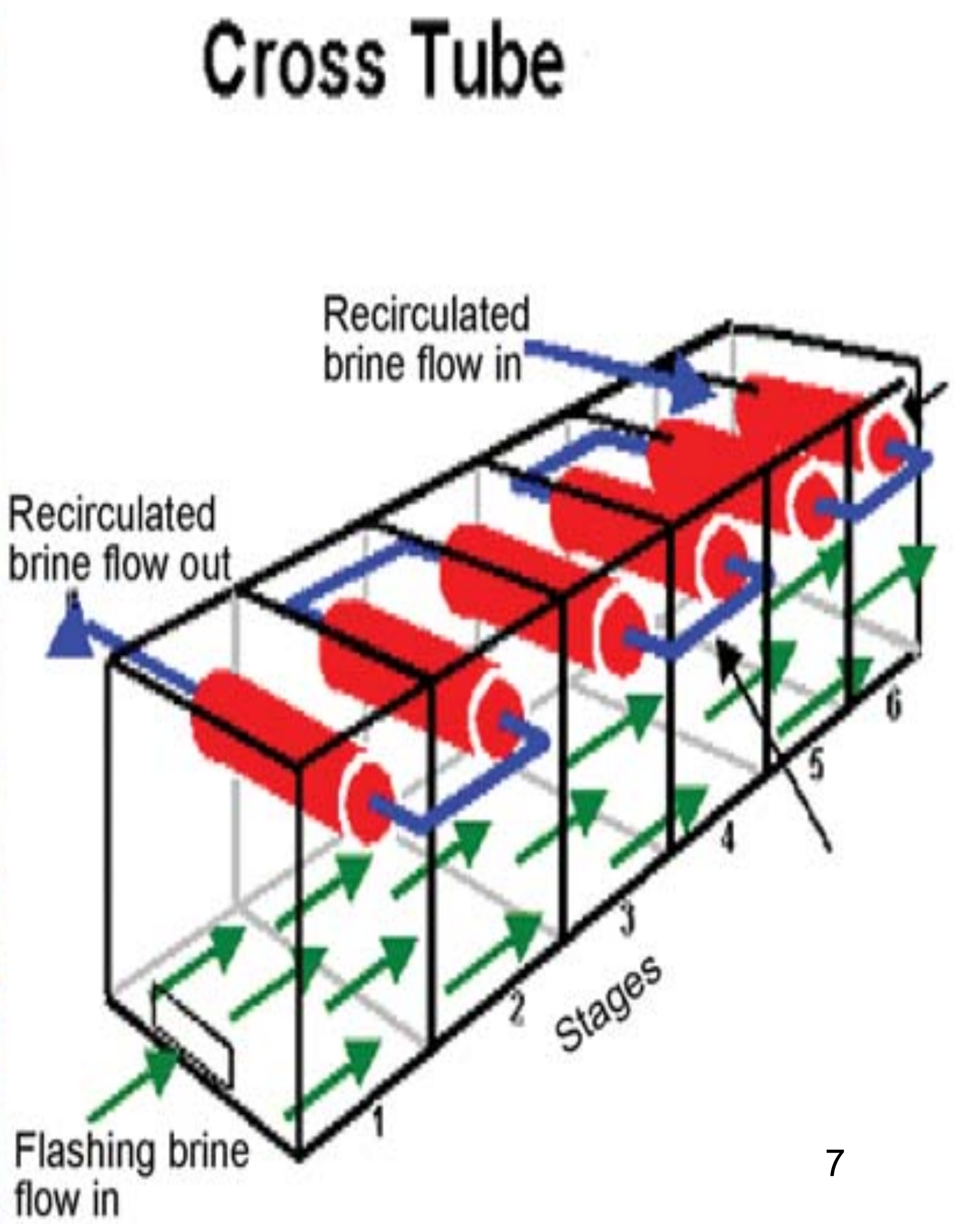


Multi stage flash — dominant technology world-wide

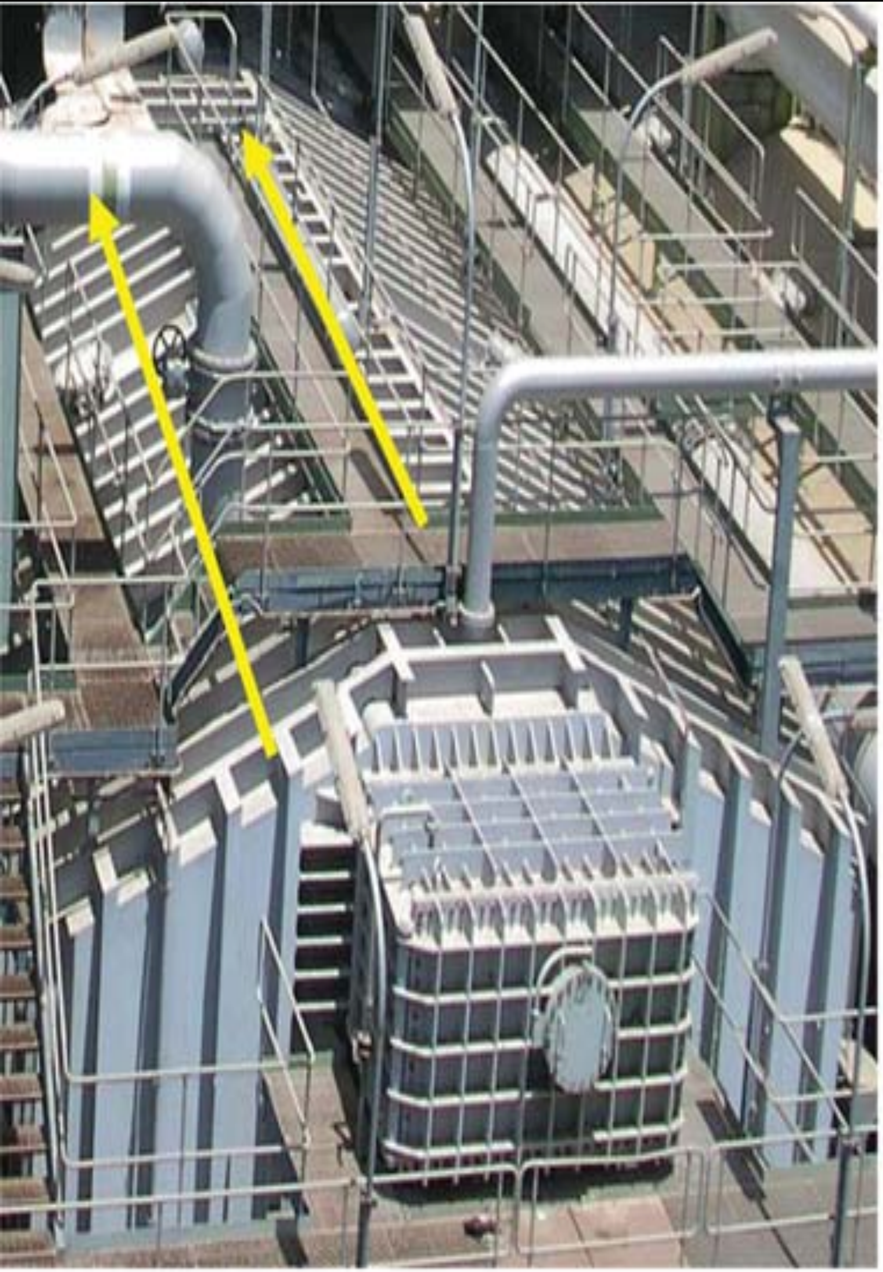
Cross-tube and long-tube MSF distillers

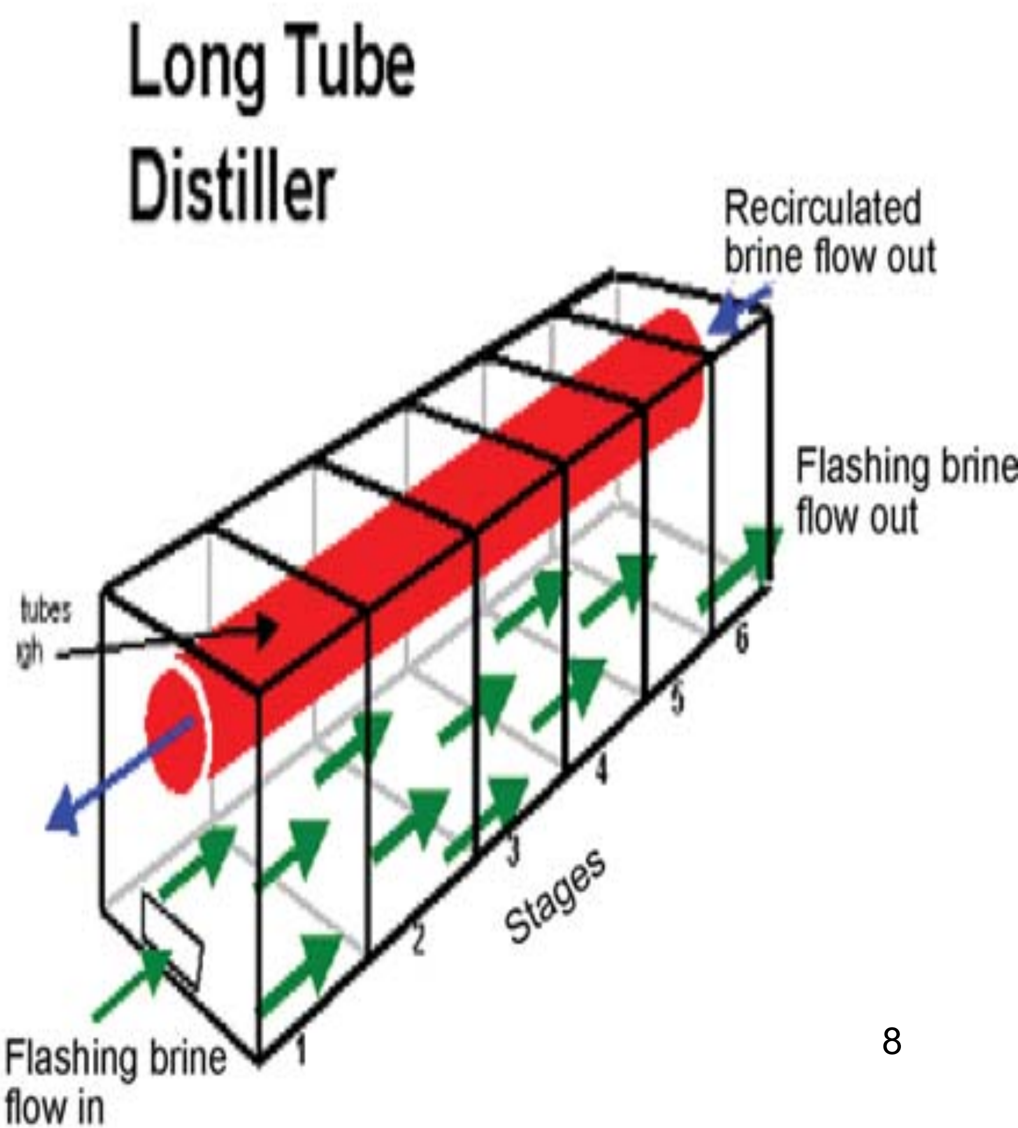


Multi stage flash



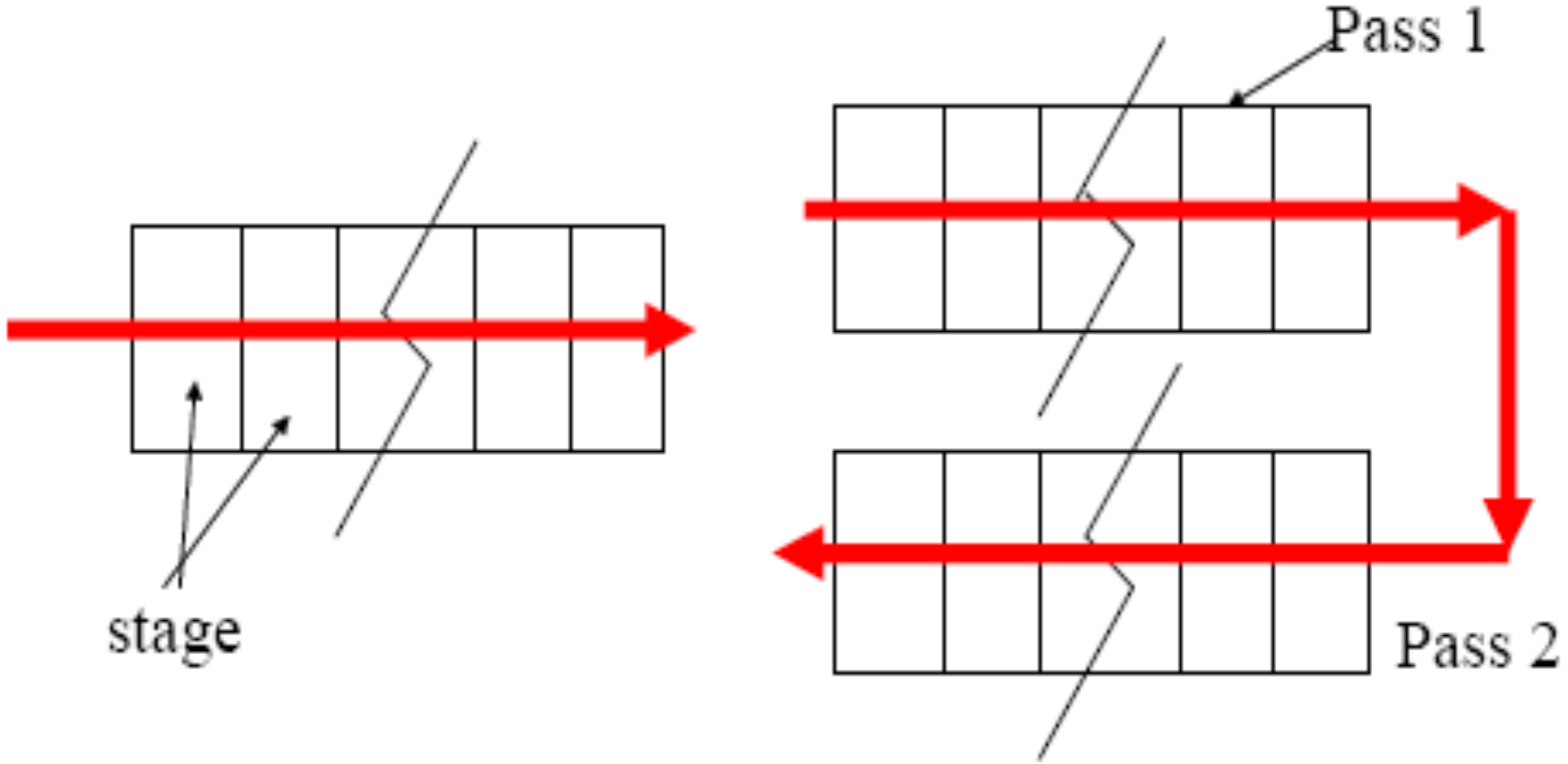
Multi stage flash





Long tube distillers:

- **we need to distinguish between the stages and the passes**



Multiple effect desalination

Evolved from small installation

to relatively large unit size

Multiple effect desalination

With thermo compression

Condesing

Performance Ratio

- **Energy input: performance ratio**

The basic relationship governing the heat transfer &distillate production in the desalination brine heater section is

$$\eta = \frac{\dot{m}_d}{\dot{m}_s} \frac{\Delta H_{ref}}{\Delta H_{h.i.}}$$

ref — reference

h.i. — heat input section

Rearranging

$$\dot{m}_s \updownarrow = \frac{\dot{m}_d \uparrow}{\eta} \frac{\Delta H_{ref}}{\uparrow \Delta H_{h.i.}}$$

Performance Ratio

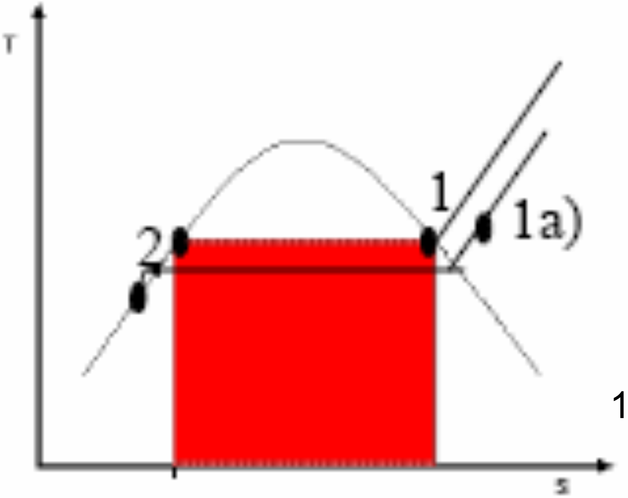
Some rule of thumb:

$$\dot{m}_s \updownarrow = \frac{\dot{m}_d \uparrow}{\eta} \frac{\Delta H_{ref}}{\uparrow \Delta H_{h.i.}}$$

- If the performance ratio increases, steam flow rate decreases.
- Steam flow rate increases if the distillate output increases

The role of the condensation enthalpy

$$\dot{m}_s \downarrow = \frac{\dot{m}_d}{\eta} \frac{\Delta H_{ref}}{\Delta H_{h.i.} \uparrow}$$



Performance Ratio

Some rule of thumb:

The higher the heat exchanged per unit of mass at the heat input section the lower the steam required

Energy input: performance ratio

$$\eta = \frac{\dot{m}_d}{\dot{m}_s}\frac{\Delta H_{ref}}{\Delta H_{h.i.}}$$

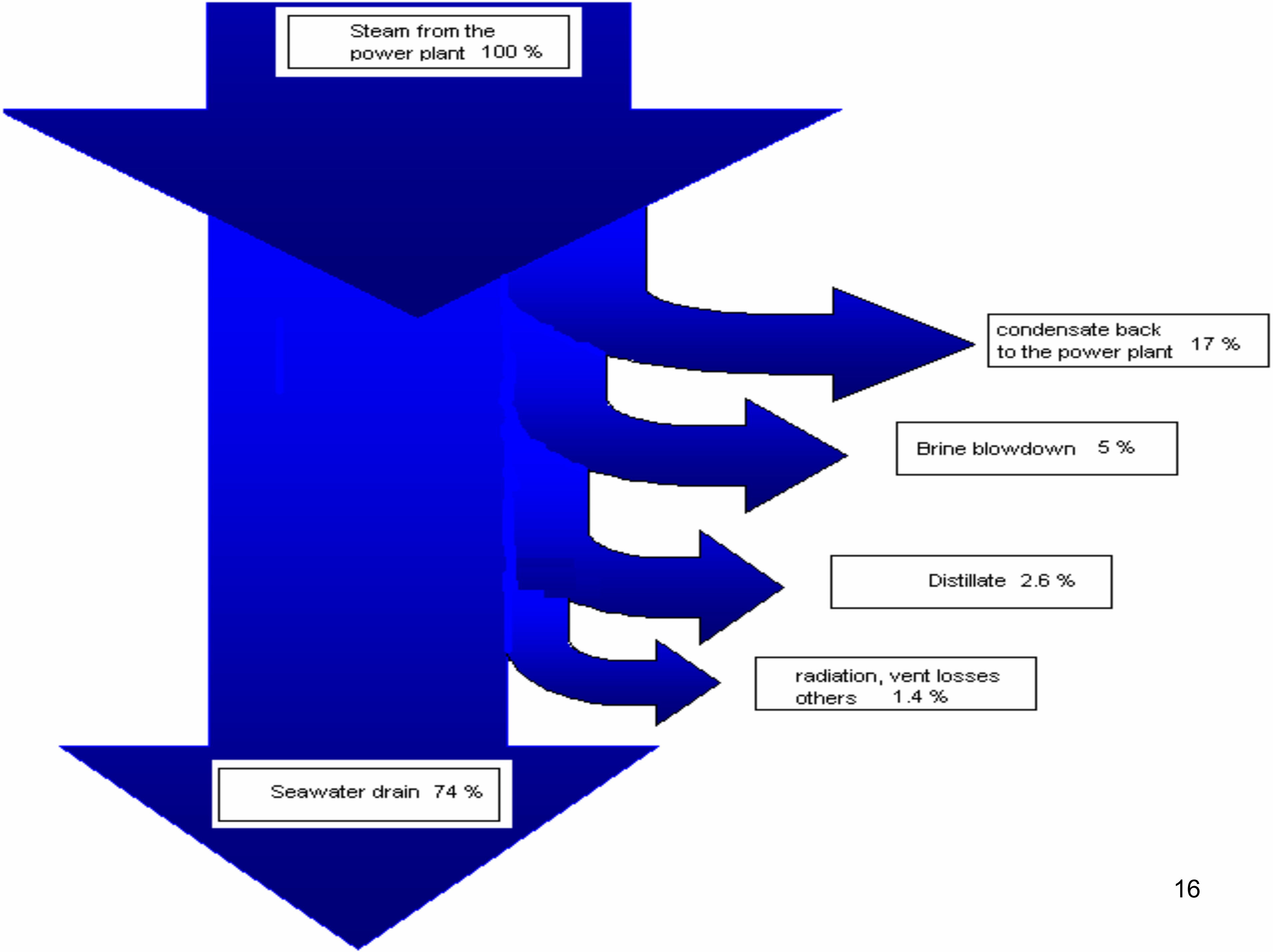
Energy input: GOR

$$GOR = \frac{\dot{m}_d}{\dot{m}_s}$$

$$\eta = GOR \quad \text{if} \quad \frac{\Delta H_{ref}}{\Delta H_{h.i.}} = 1$$

Energy effect

- In fact, as it can be seen from the energy flow diagram below, the great part of the heat input to the MSF system is returned back to the sea with the seawater drain stream.



Steam from the power plant 100 %

condensate back to the power plant 17 %

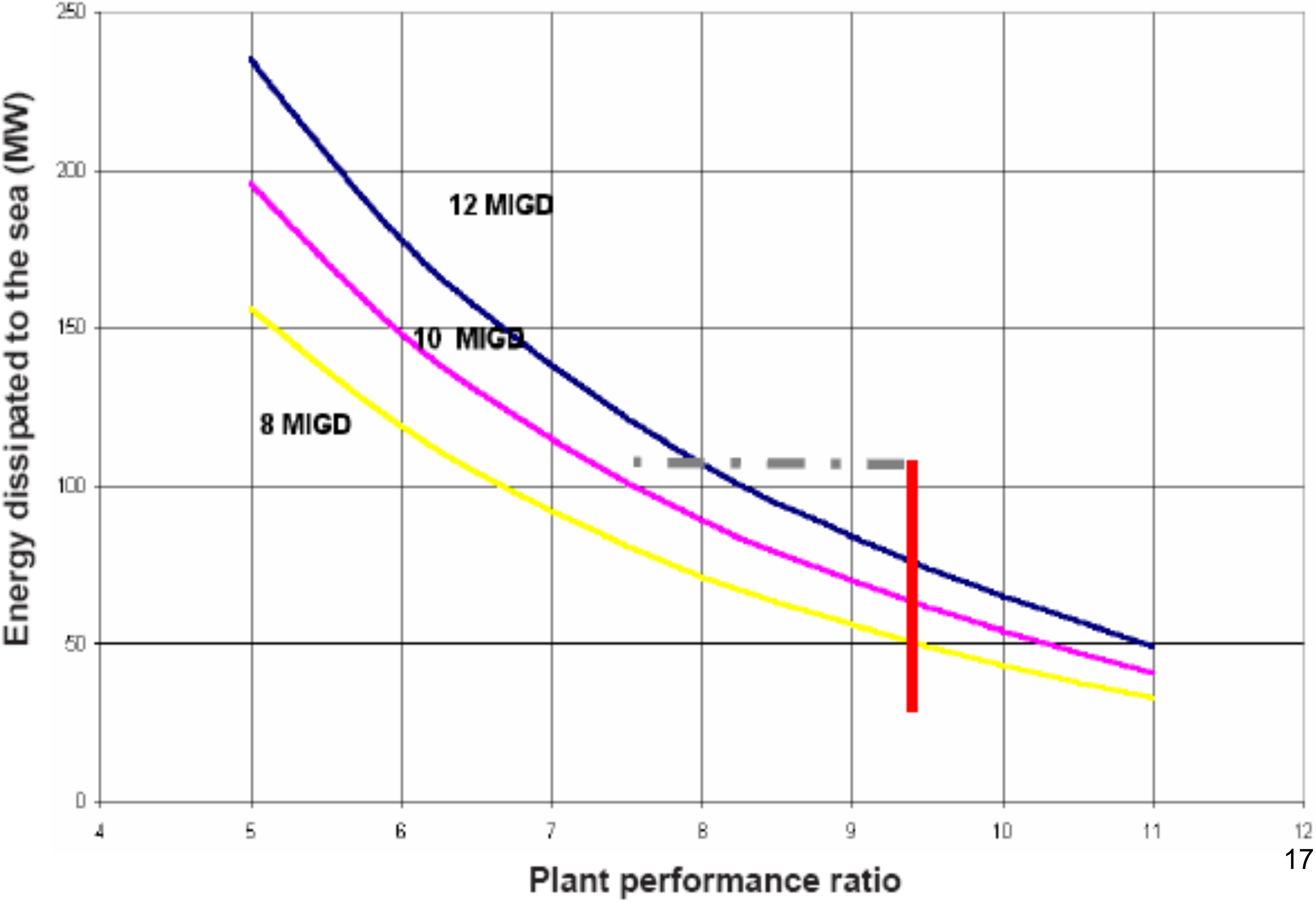
Brine blowdown 5 %

Distillate 2.6 %

radiation, vent losses others 1.4 %

Seawater drain 74 %

Discharge energy and efficiency



Working Example

distillate production $D := 1320 \cdot \frac{tonne}{hr}$

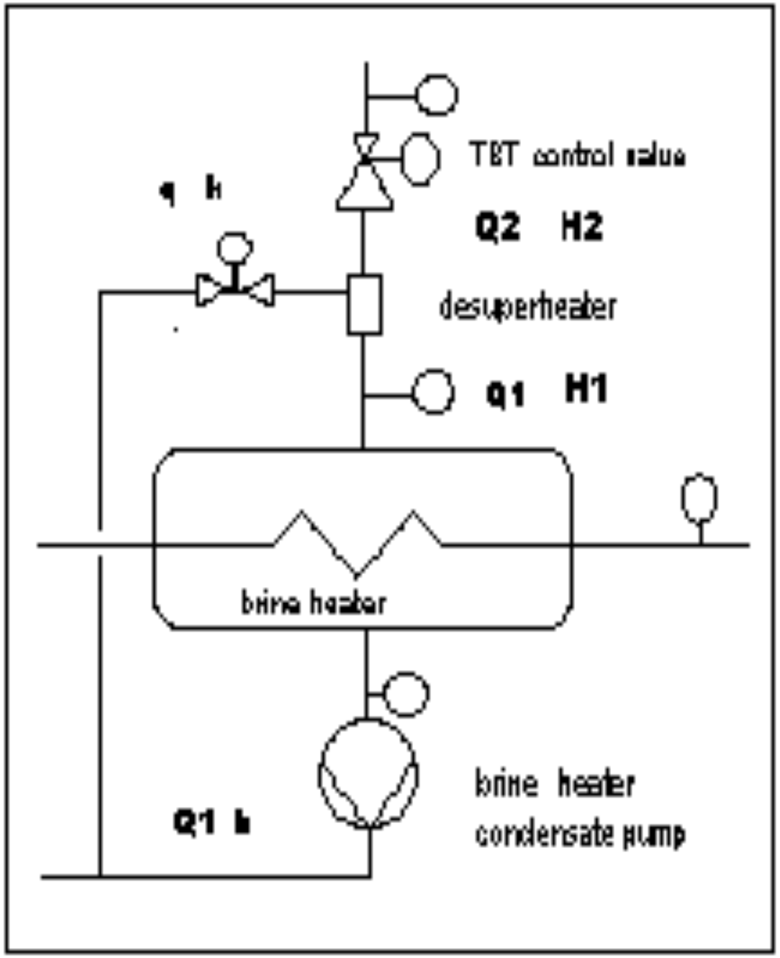
performance ratio $\eta := 7.$

reference enthalpy $Href := 2326 \cdot \frac{kj}{kg}$

Condensate enthalpy
T = 109 °C $h := 457.8 \cdot \frac{kj}{kg}$

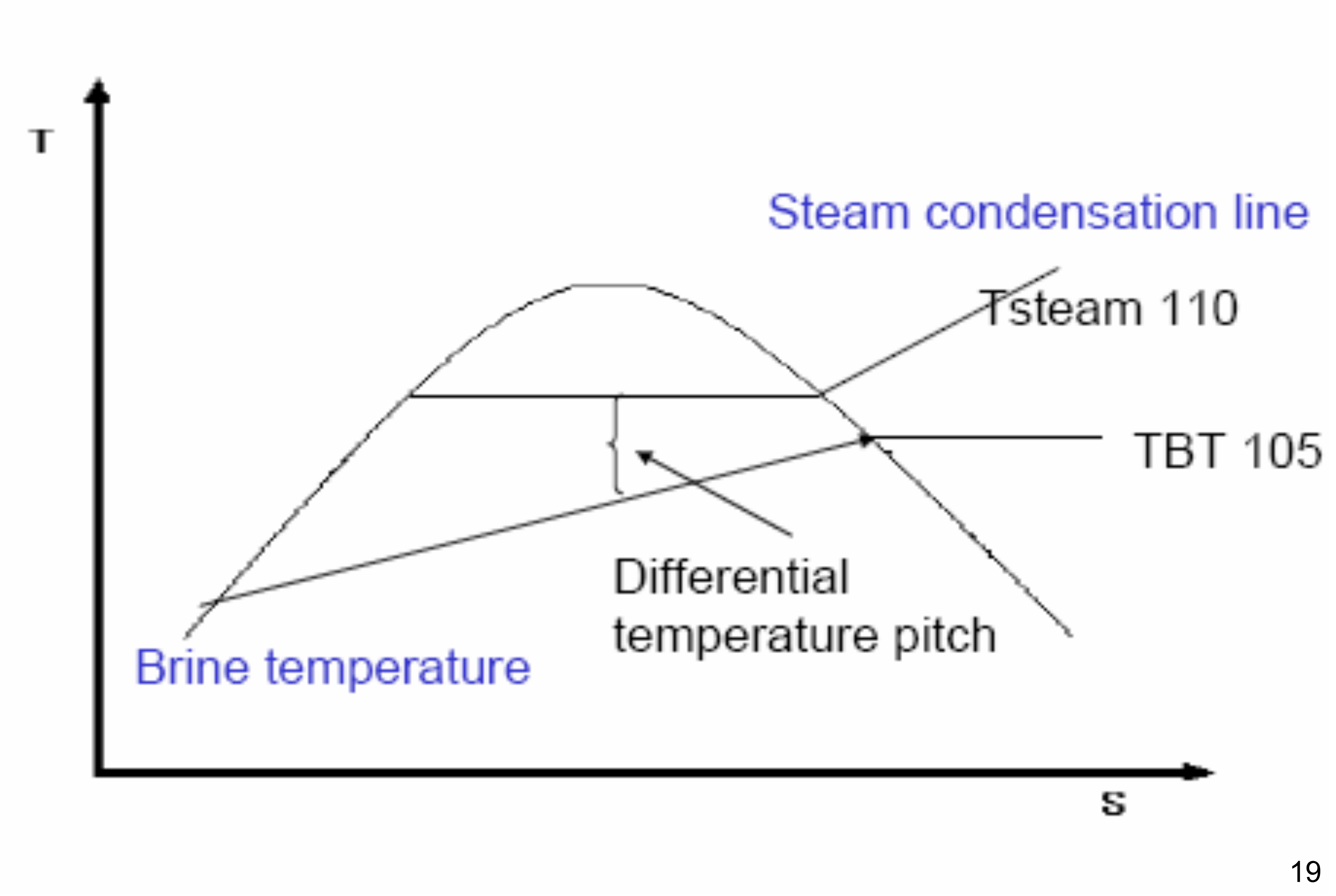
steam to brine heater enthalpy
T 110°C p 1.21 bar $H1 := 2694.0 \cdot \frac{kj}{kg}$

steam flow to be achieved



$$Q1 := \frac{D \cdot (Href)}{\eta \cdot (H1 - h)} \qquad Q1 = 196.144 \cdot \frac{tonne}{hr}$$

Working Example



MSF plant steam to brine heater calculation

Data available:

- Reference enthalpy = 2326 kJ/kg of steam
- Desired distillate flow = 1320 ton/h
- Steam temperature at brine heater = 110ºC
- Condensate return temperature = 109ºC

Calculate:

- Steam–condensate flow rate to–from the plant

From steam tables

Steam enthalpy H = 2694 kJ/kg

Condensate enthalpy h = 457.8 kJ/kg

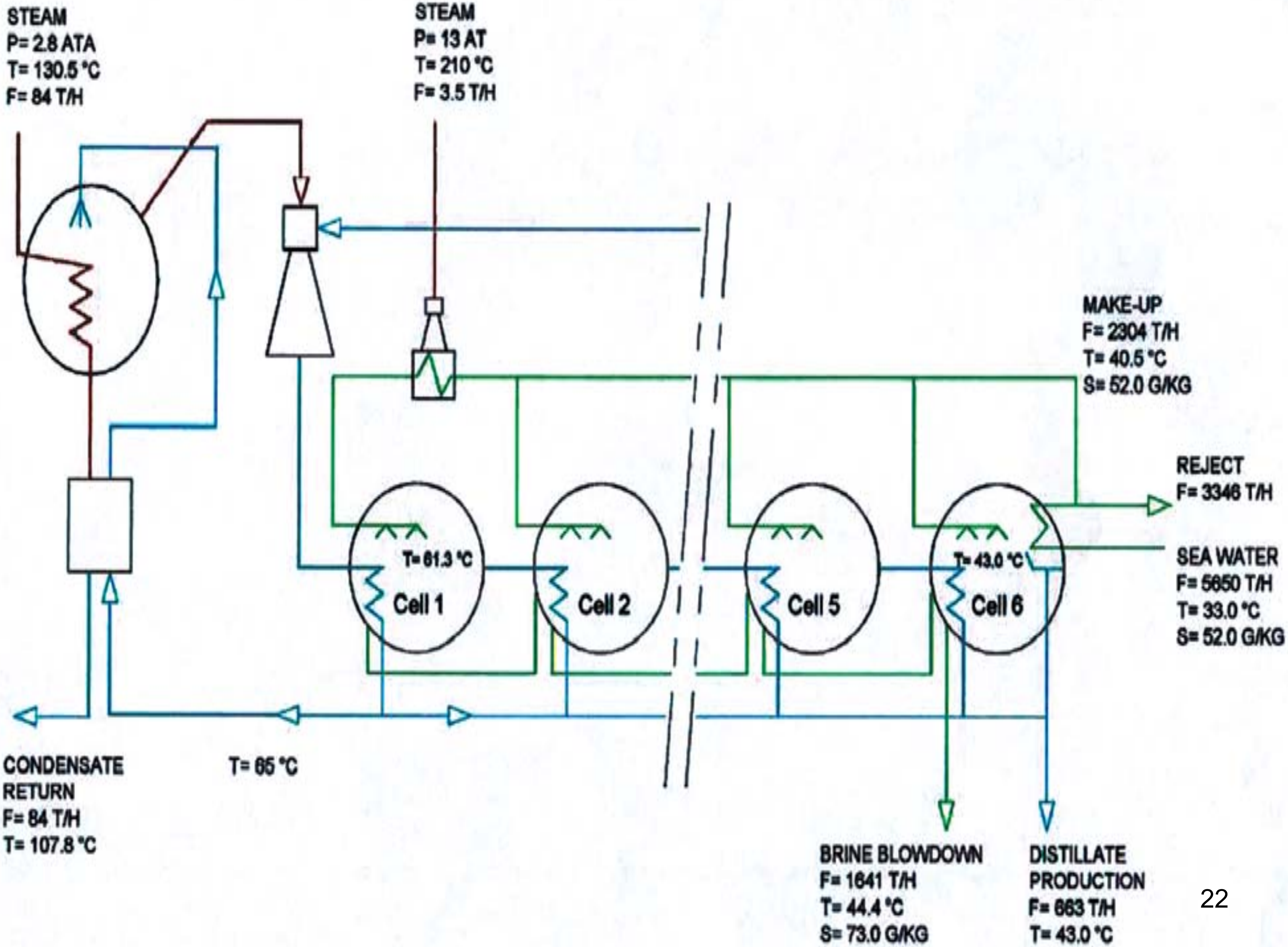
MSF plant steam to brine heater calculation

- Enthalpy exchanged at desalination heat input stage

$$\Delta H = (2694 - 457.8)\ \mathrm{kJ/kg} = 2236\ \mathrm{kJ/kg}$$

$$\eta = \frac{\dot{m}_d}{\dot{m}_s}\frac{\Delta H_{ref}}{\Delta H_{bh}} \Rightarrow 7 = \frac{1320}{\dot{m}_s}\cdot\frac{2326}{2236}\cdot\frac{\mathrm{ton}}{\mathrm{h}}\cdot\frac{\mathrm{kJ}}{\mathrm{kg}}\cdot\frac{\mathrm{kg}}{\mathrm{kJ}}\cdot\frac{\mathrm{h}}{\mathrm{ton}}$$

$$\dot{m}_s = \frac{1320}{7}\cdot\frac{2326}{2236}\cdot\frac{\mathrm{ton}}{\mathrm{h}} = 196.16\cdot\frac{\mathrm{ton}}{\mathrm{h}}$$


STEAM
P= 2.8 ATA
T= 130.5 °C
F= 84 T/H
STEAM
P= 13 AT
T= 210 °C
F= 3.5 T/H
MAKE-UP
F= 2304 T/H
T= 40.5 °C
S= 52.0 G/KG
REJECT
F= 3346 T/H
SEA WATER
F= 5650 T/H
T= 33.0 °C
S= 52.0 G/KG
T= 61.3 °C
T= 43.0 °C
Cell 1
Cell 2
Cell 5
Cell 6
CONDENSATE
RETURN
F= 84 T/H
T= 107.8 °C
T= 65 °C
BRINE BLOWDOWN
F= 1641 T/H
T= 44.4 °C
S= 73.0 G/KG
DISTILLATE
PRODUCTION
F= 663 T/H
T= 43.0 °C

MED operating data

Calculate:

- Performance ratio based on $H_{ref} = 2326$ kJ/kg
- Specific energy consumption

MED operating data

- **Performance ratio**

Steam enthalpy *H* = 2721 kJ/kg

Condensate enthalpy *h* = 452.2 kJ/kg

$$\eta = \frac{\dot{m}_d}{\dot{m}_s}\frac{\Delta H_{ref}}{\Delta H_{S.T.}} \Rightarrow \frac{663}{84}\cdot\frac{2326}{(2721-452.2)}\cdot\frac{\text{ton}}{\text{h}}\cdot\frac{\text{h}}{\text{ton}}\cdot\frac{\text{kJ}}{\text{kg}}\cdot\frac{\text{kg}}{\text{kJ}}$$

$$\eta = \frac{663}{84}\cdot\frac{2326}{2268.8} = 8.09\cdot(-)$$

MED operating data

- **Specific heat consumption**

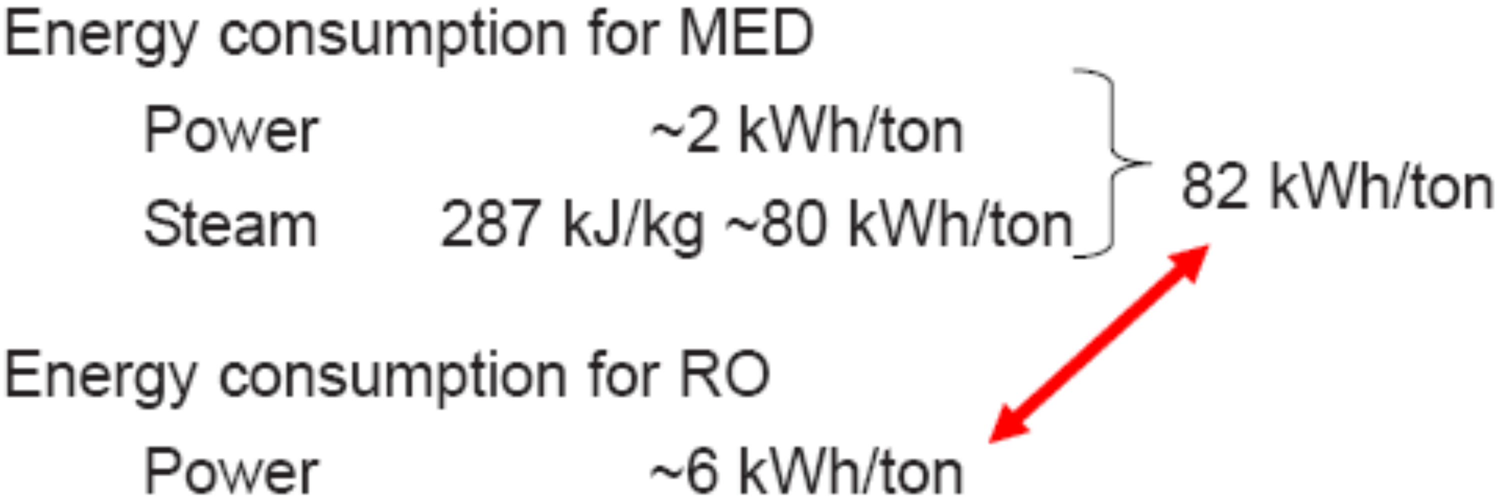
Steam enthalpy $H = 2721$ kJ/kg

Condensate enthalpy $h = 452.2$ kJ/kg

$$\Delta H_{S.T.} = (2721 - 452.2) \cdot \frac{\text{kJ}}{\text{kg}} = 2268.8 \cdot \frac{\text{kJ}}{\text{kg}}$$

$$E = \Delta H_{S.T.} \cdot \frac{\dot{m}_c}{\dot{m}_d} = 2268.8 \cdot \frac{84}{663} \frac{\text{kJ}}{\text{kg}} \cdot \frac{\text{ton}}{\text{h}} \cdot \frac{\text{h}}{\text{ton}} = \boxed{287 \cdot \frac{\text{kJ}}{\text{kg}}}$$

Understanding the real value: Exergy



Energy consumption for MED

Power ~2 kWh/ton

Steam 287 kJ/kg ~80 kWh/ton

82 kWh/ton

Energy consumption for RO

Power ~6 kWh/ton

Second law of thermodynamics:

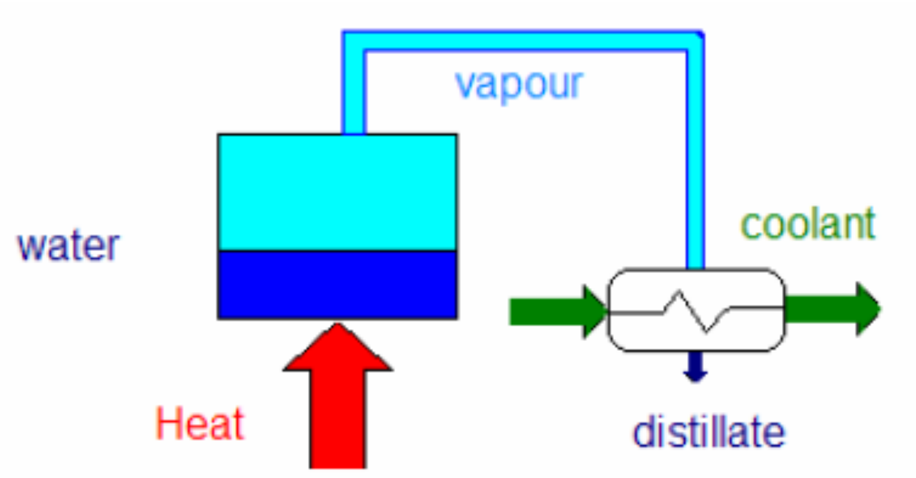
how much do we pay for the 80 kWh/h, how much electric power could be worth?

The Multiple Cells Concept

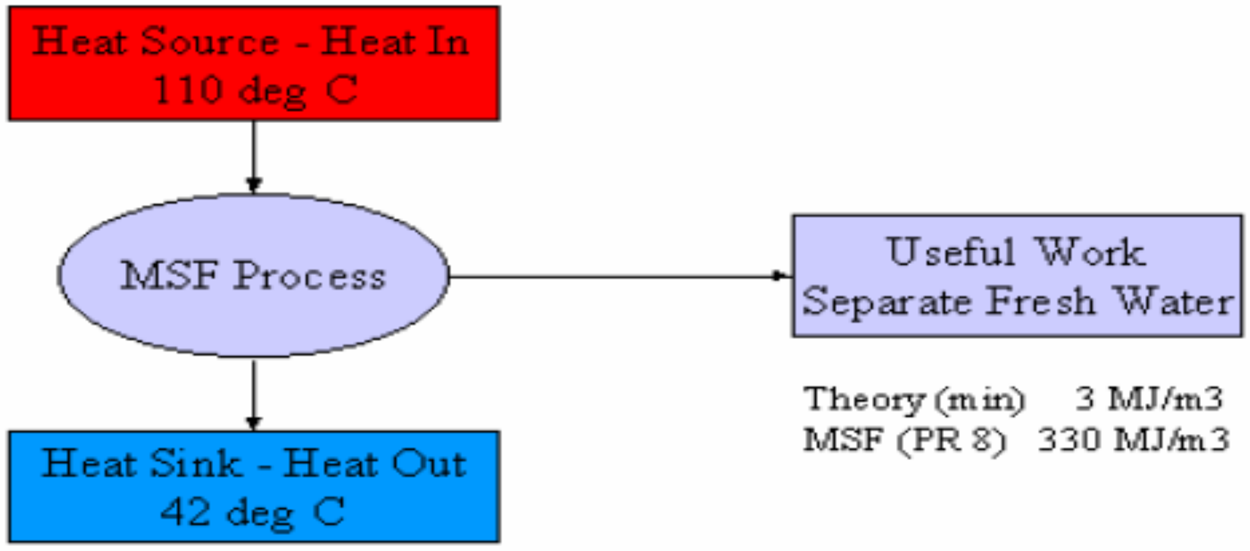
- **The problem with simple “single effect” stills**

- Heat is added to raise the feed water to its boiling point
- Additional heat (the heat of vaporization) is added to convert the hot water to steam at the same temperature
- (The heat of vaporization may be 6–7 times the heat needed just to heat the water)
- All of this heat is then lost to the condensing water (or air)

The Multiple Cells Concept



Thermal Process



Theory (min) 3 MJ/m3
MSF (PR 8) 330 MJ/m3

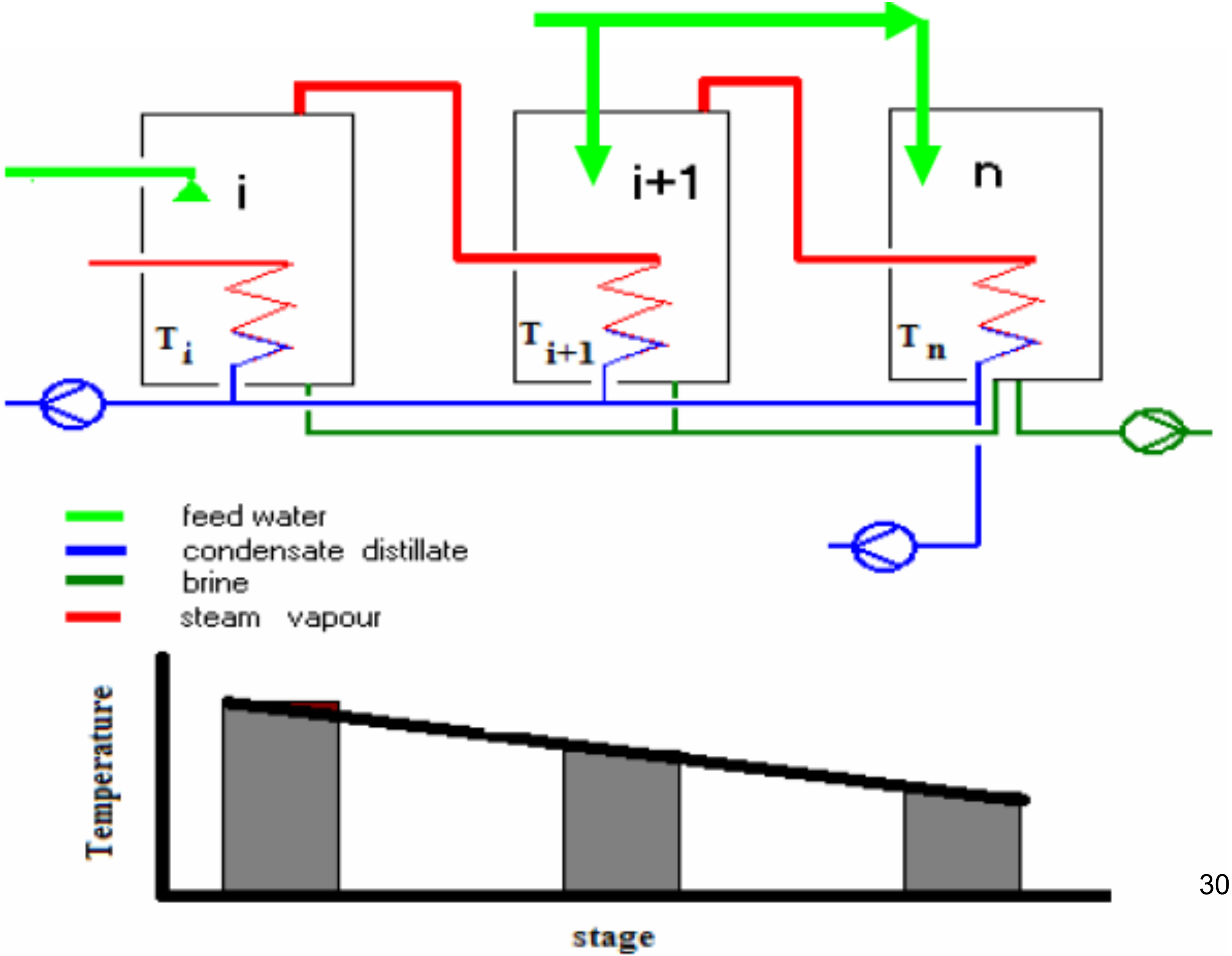
The concept of multiple stage: (early 1990)

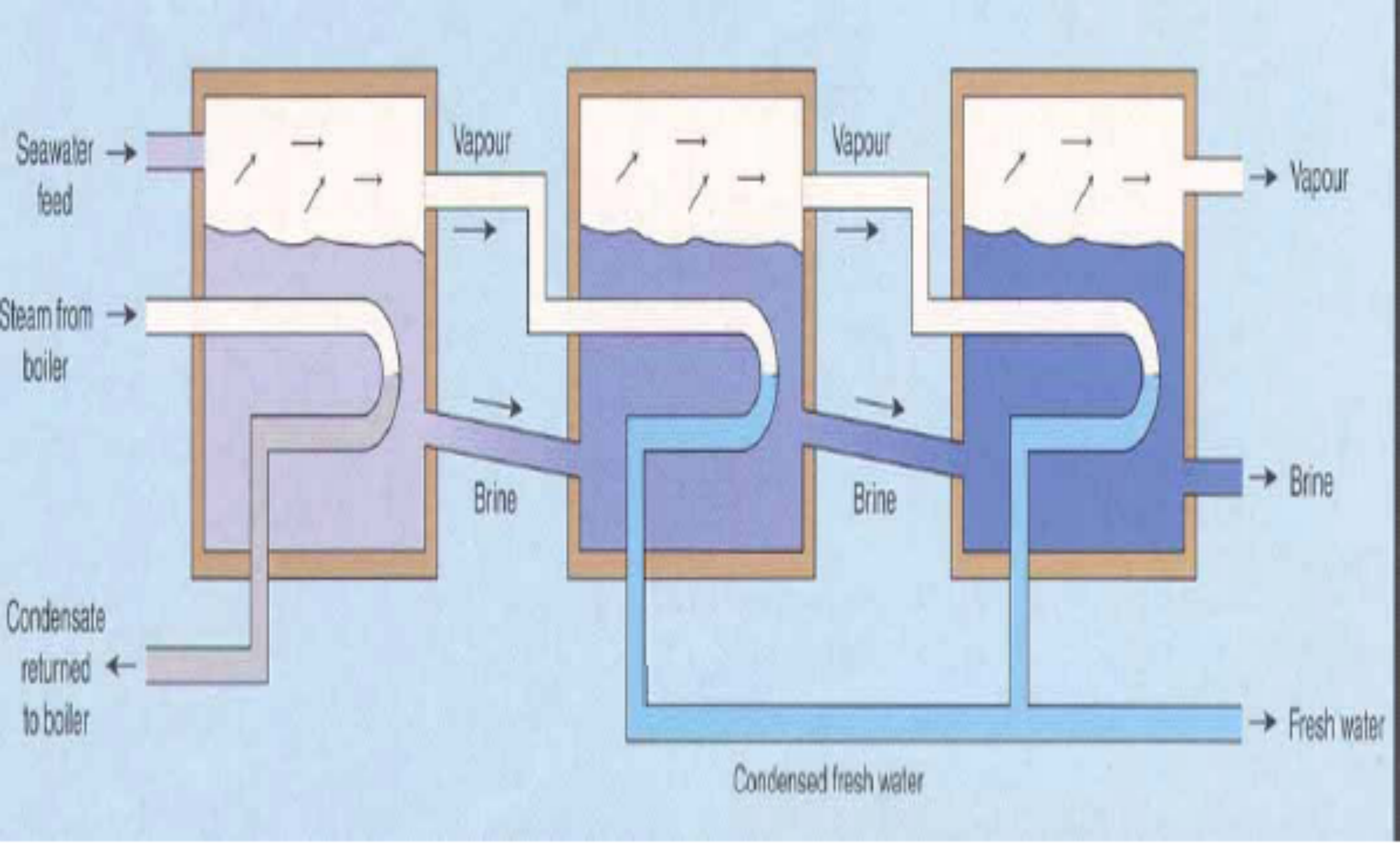
The boiling point of water is lower at reduced pressure.

- The heat released during condensation can be utilized to evaporate additional vapor from the condensing water if the condensing water is held at a reduced pressure.

- This new vapor may now in turn be condensed by water at a yet lower pressure, again generating new vapor.

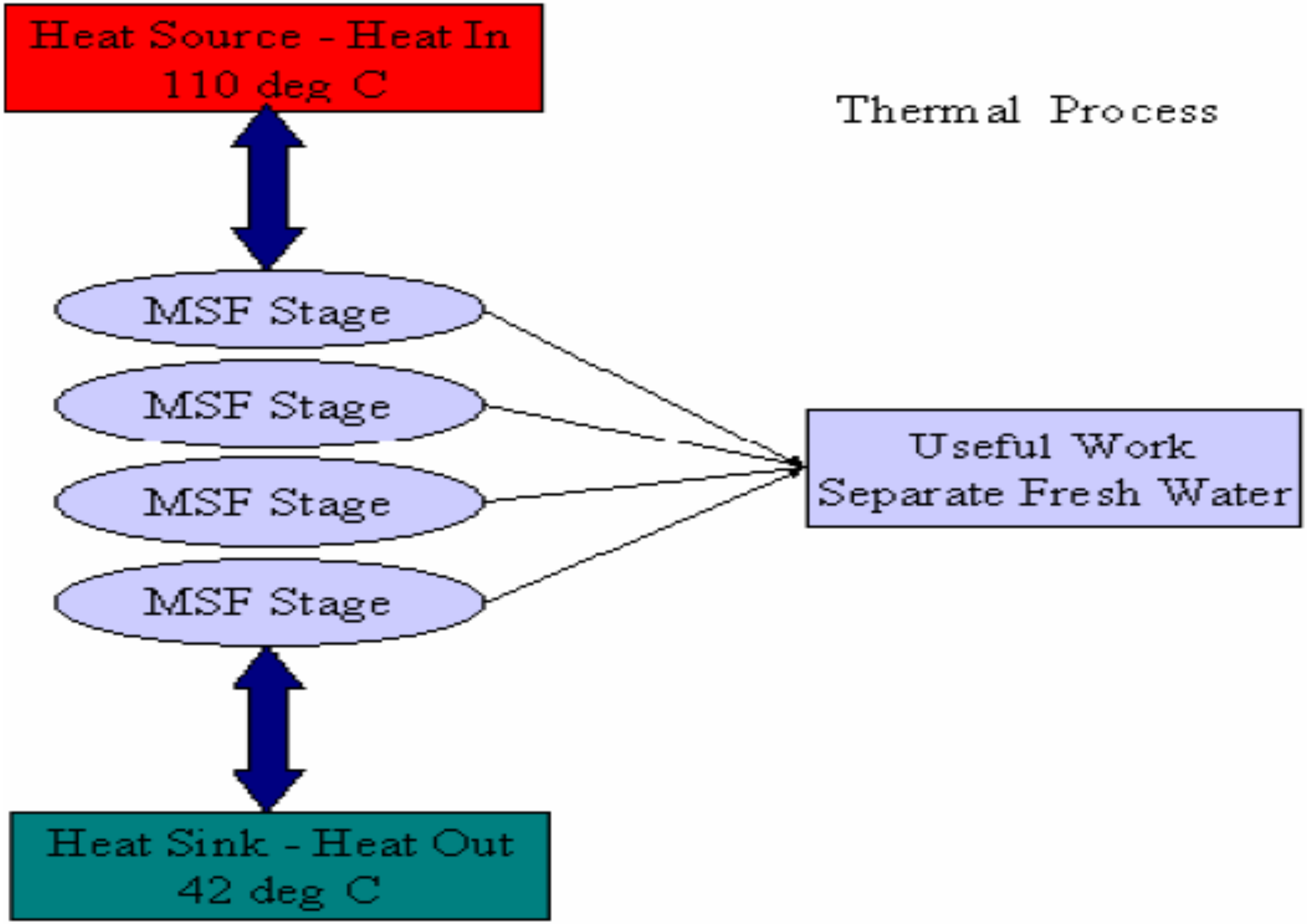
Schematic diagram





- Thus the heat is "recycled" within the system. Energy efficiency is a function of the number of effects.

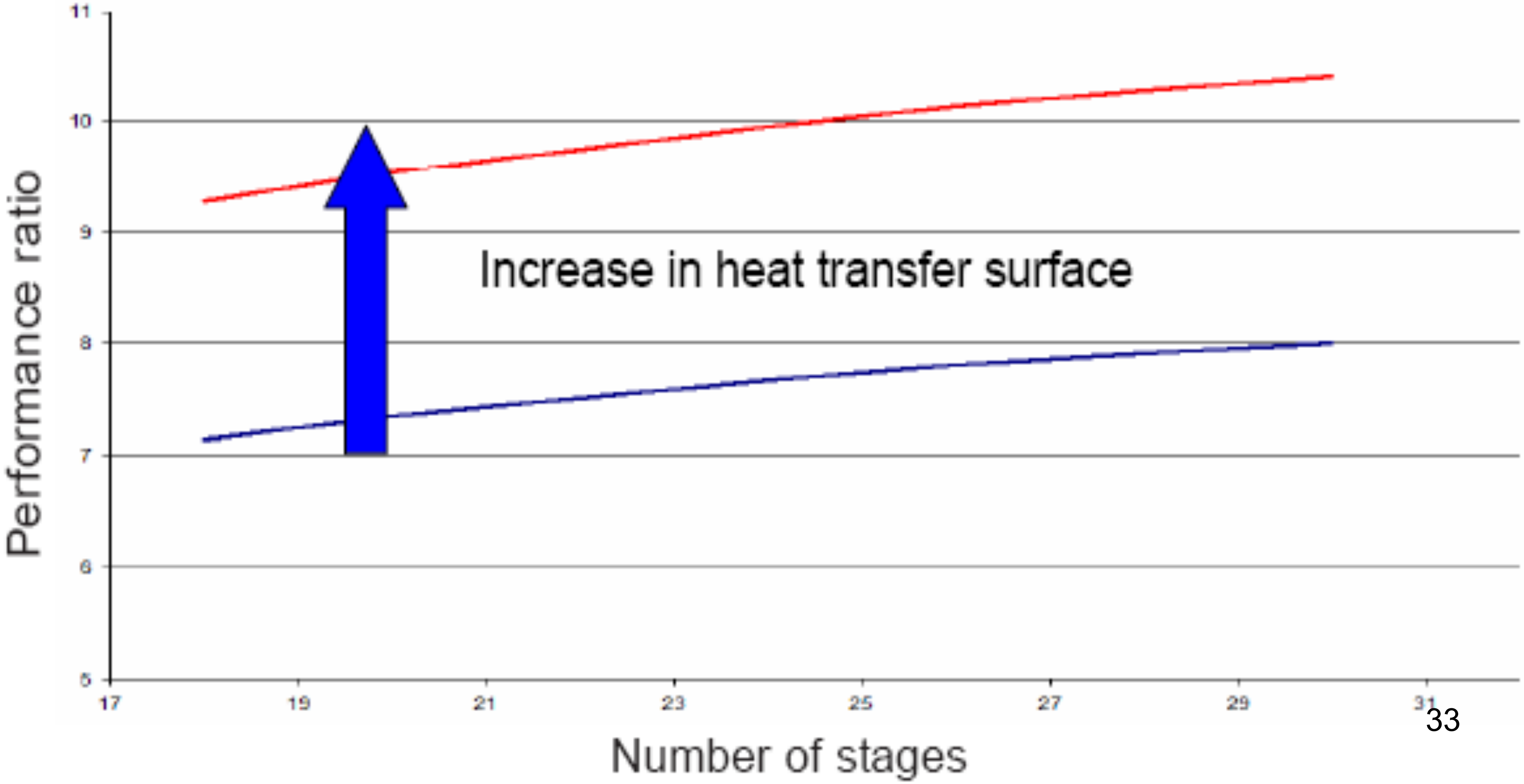
MSF example


Heat Source - Heat In
110 deg C
Thermal Process
MSF Stage
MSF Stage
MSF Stage
MSF Stage
Useful Work
Separate Fresh Water
Heat Sink - Heat Out
42 deg C

MSF example

- **Some rule of thumb**

At constant heat transfer surface increasing the number of stages brings about an increase in plant performance ratio



COMBINED POWER AND THERMAL DESALINATION PLANTS

- **Interfaces with the rest of the yard**
- **Typical layout**
- **Power desalination plant combination**

INTERFACES WITH THE REST OF THE YARD

Interfaces with the rest of the plant

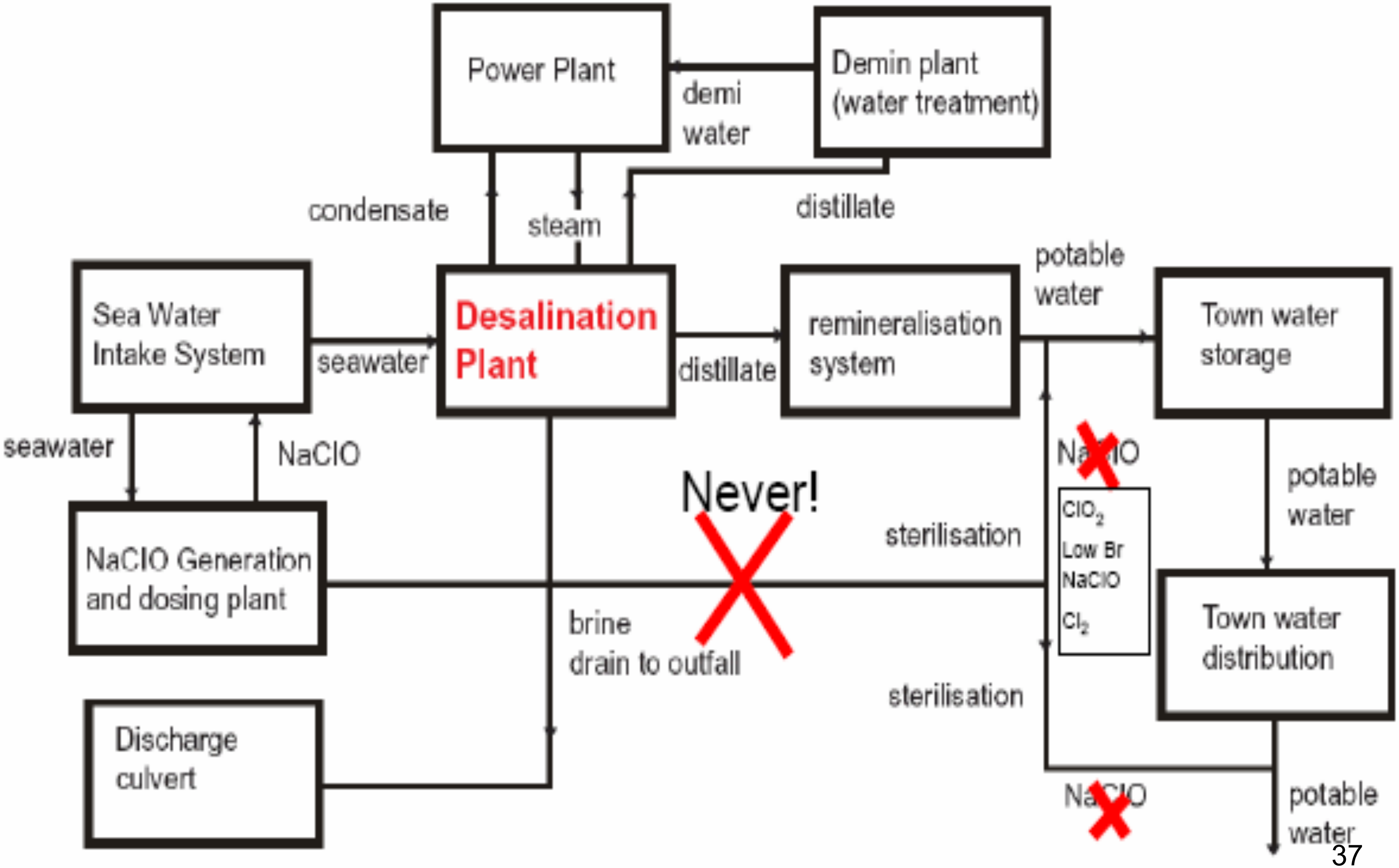
These include several engineering work packages part of the desali- nation BOP.

- Seawater intake system
- Discharge outfall
- Chlorine generation and dosing
- Remineralisation (potabilisation)
- Town water and storage
- Potable water pumping
- Etc.
- Demineralisation plant

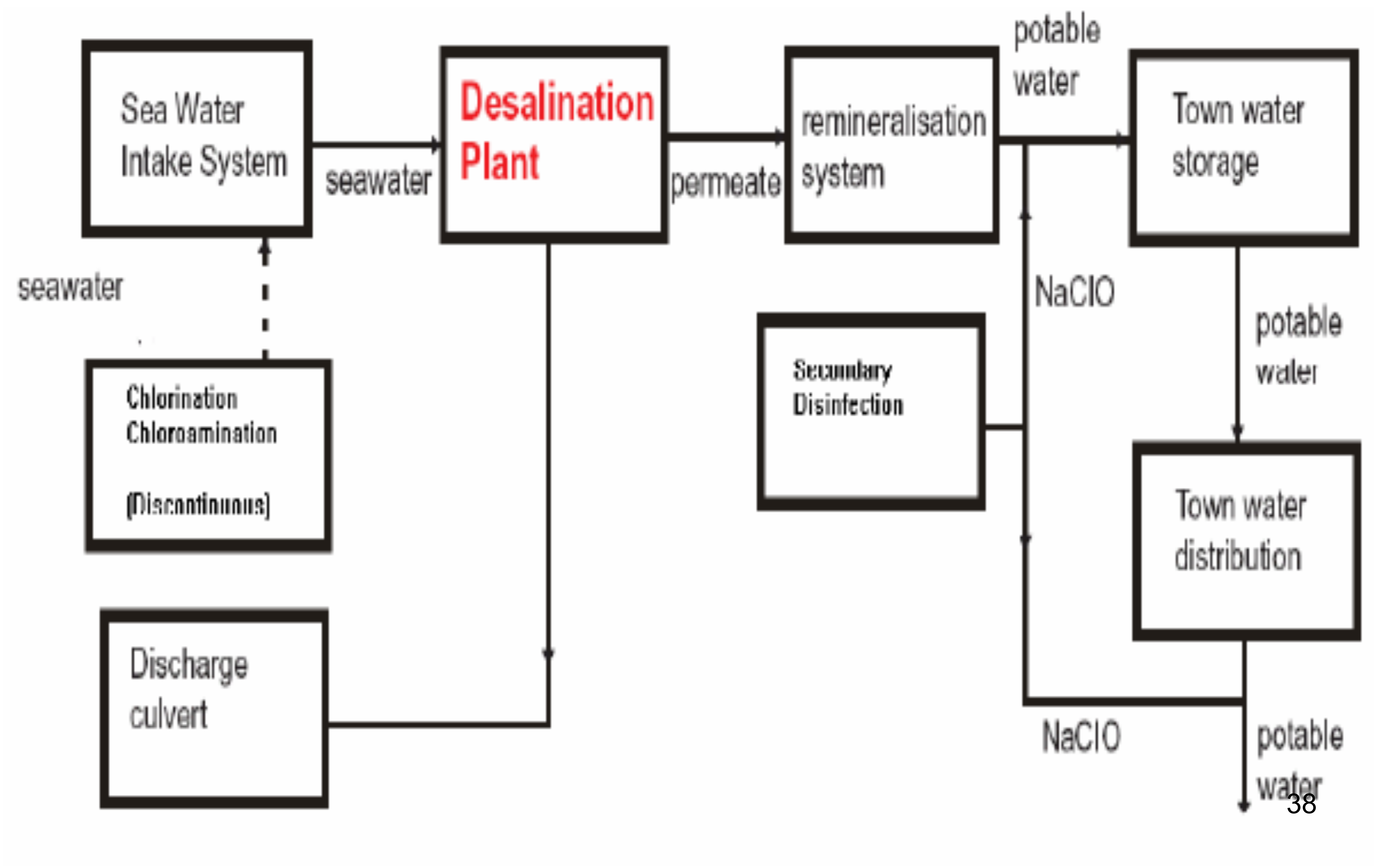
INTERFACES WITH THE REST OF THE YARD

- Steam condensate system various options:
  - Power plant
  - Heat recovery steam generator
  - Auxiliary steam generator
  - Others

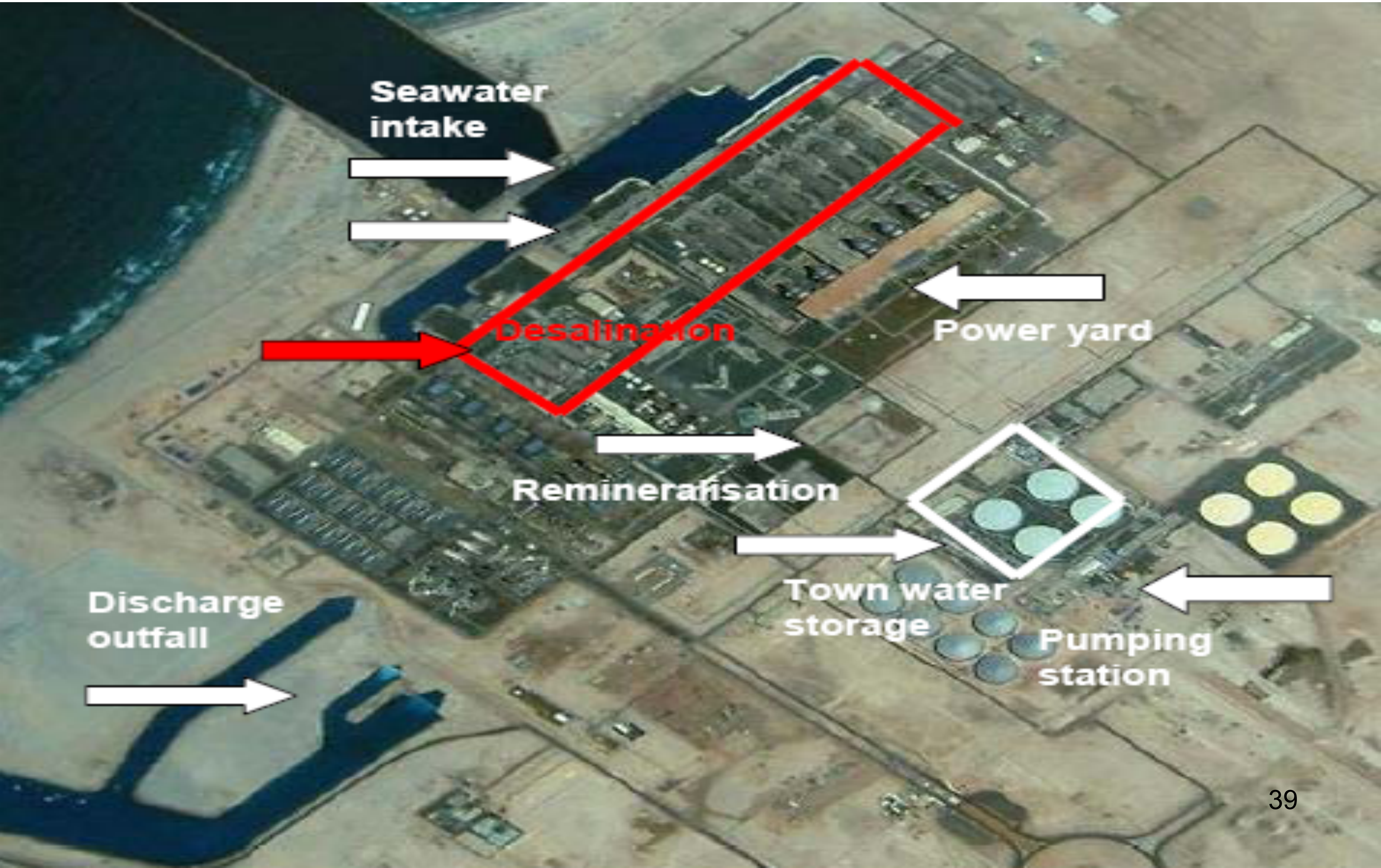
Interfaces with the rest of the plant: thermal



Interfaces with the rest of the plant: membrane



Interfaces with the rest of the plant
Typical layout



Typical layout

Typical layout

Classification of process interfaces for thermally-driven desalination plants

1. With associated power generator

  1.1. Steam from heat recovery steam generation

  1.2. Steam from back pressure steam turbine (combined cycle)

  1.3. Steam from condensing steam turbine

2. Without associated power generator

  2.1. Steam from auxiliary boiler

  2.2. Heat from other sources: i.e. solar, etc.

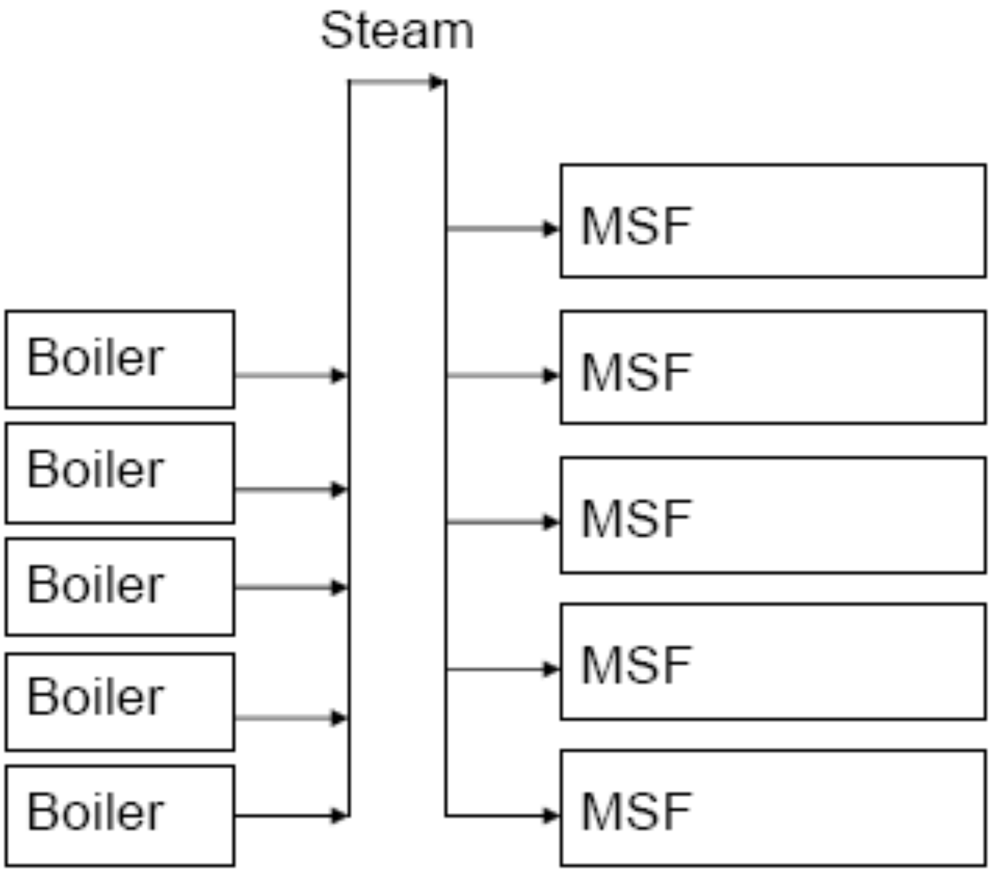
Matching power and water generation

- Matching power and water generation is important in order to reduce power and water cost by effective co-generation.

- **What are the targets?**

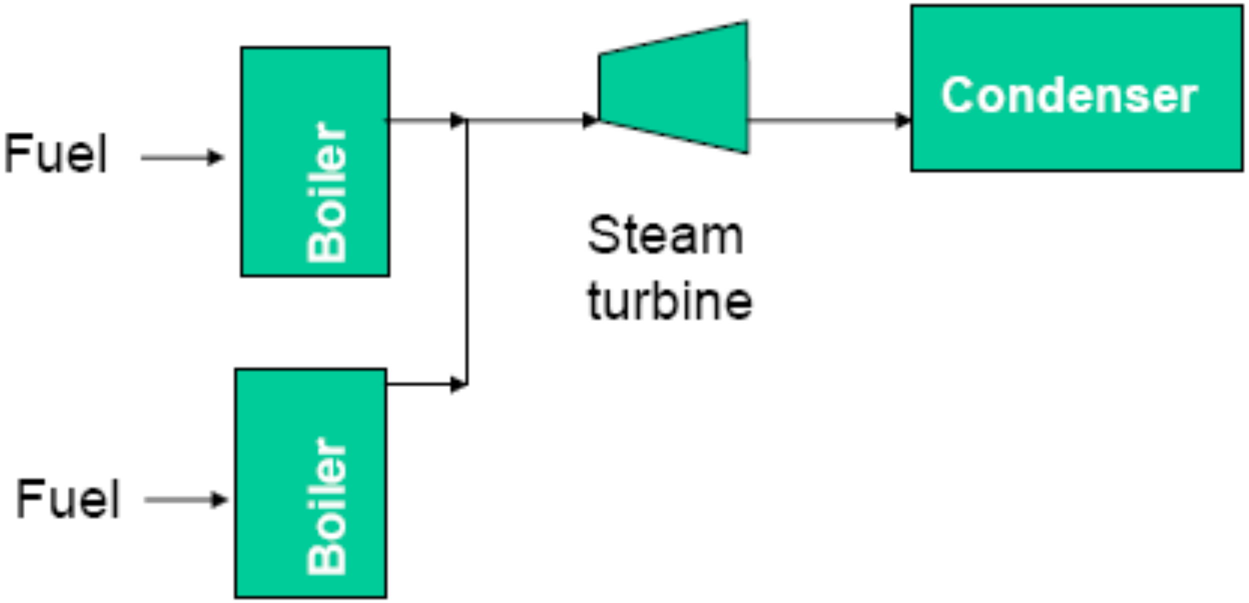
- Identify a heat load that allows cost effective power generation.

Matching power and water generation: option 1



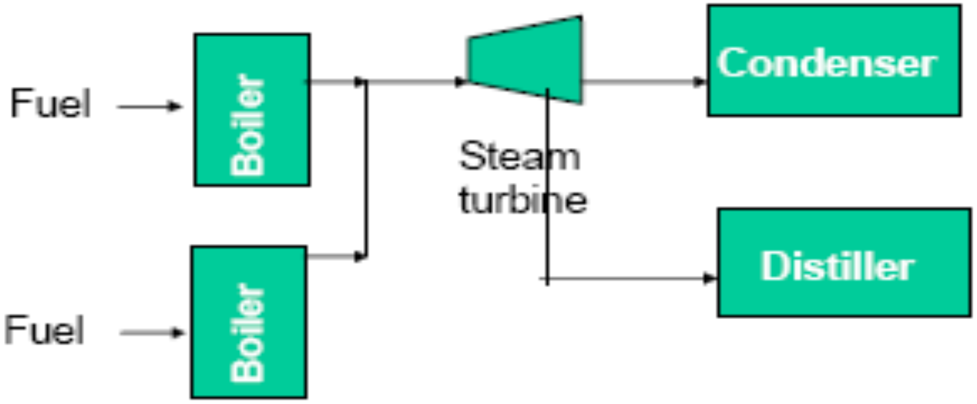
Big kettle

Matching power and water generation: option 2

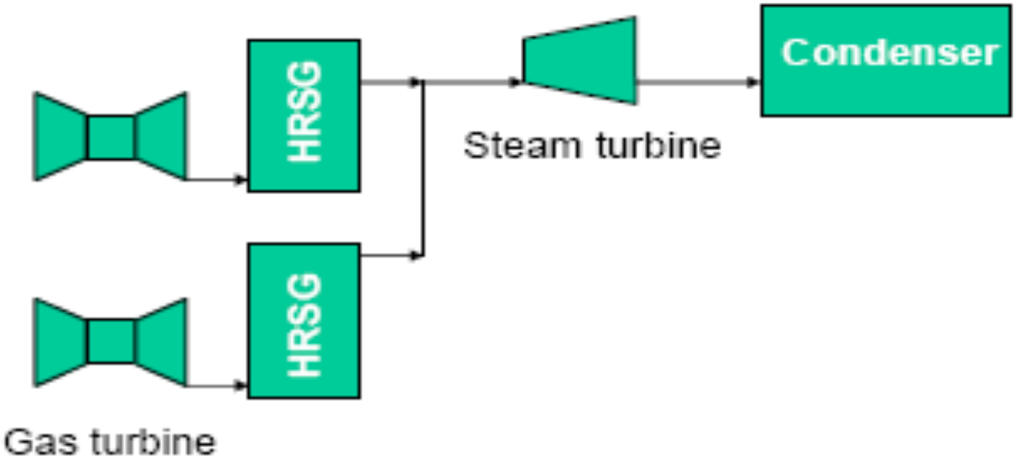


Conventional power plant

Matching power and water generation

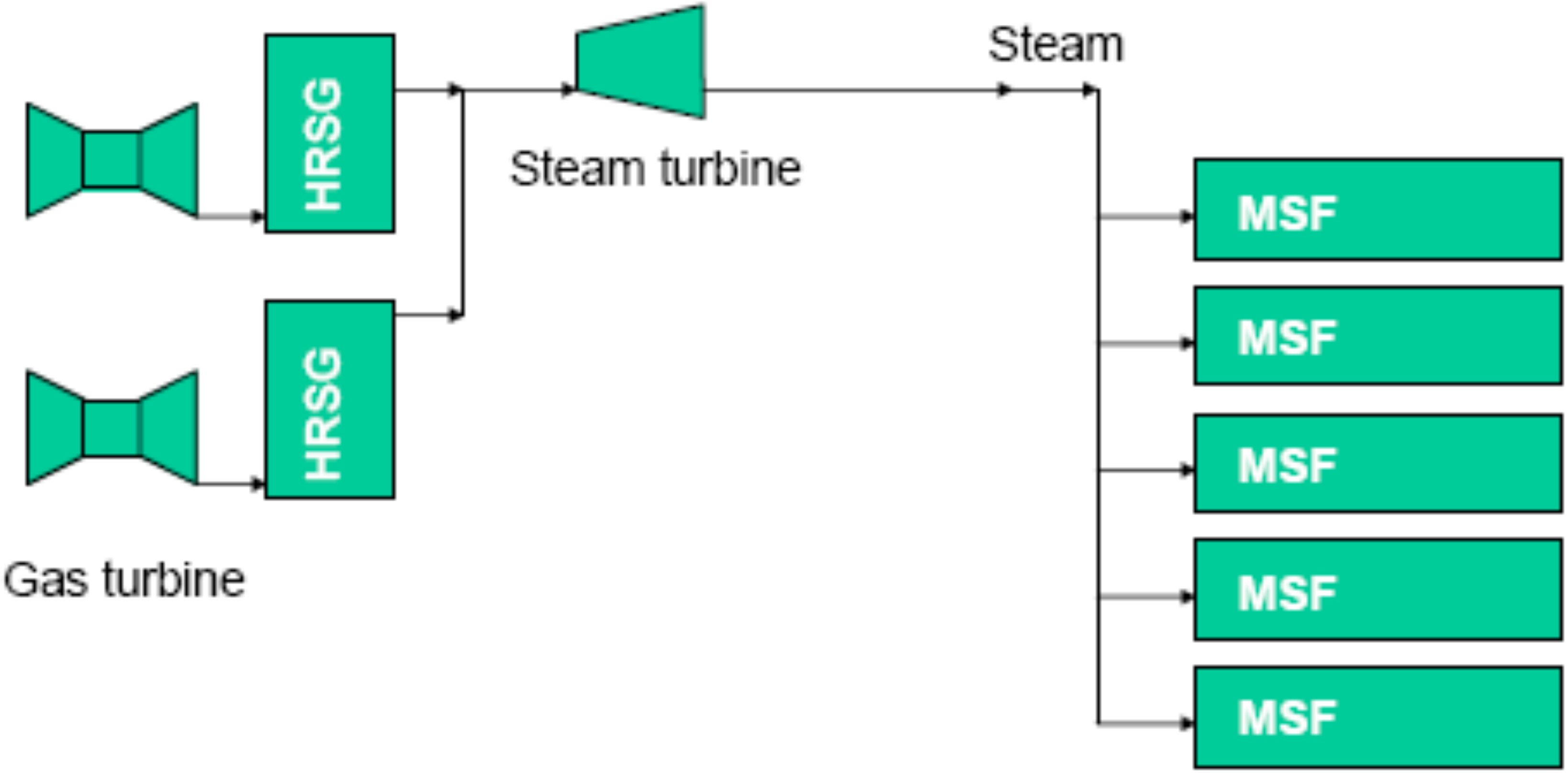


Conventional power plant - **Cogeneration**



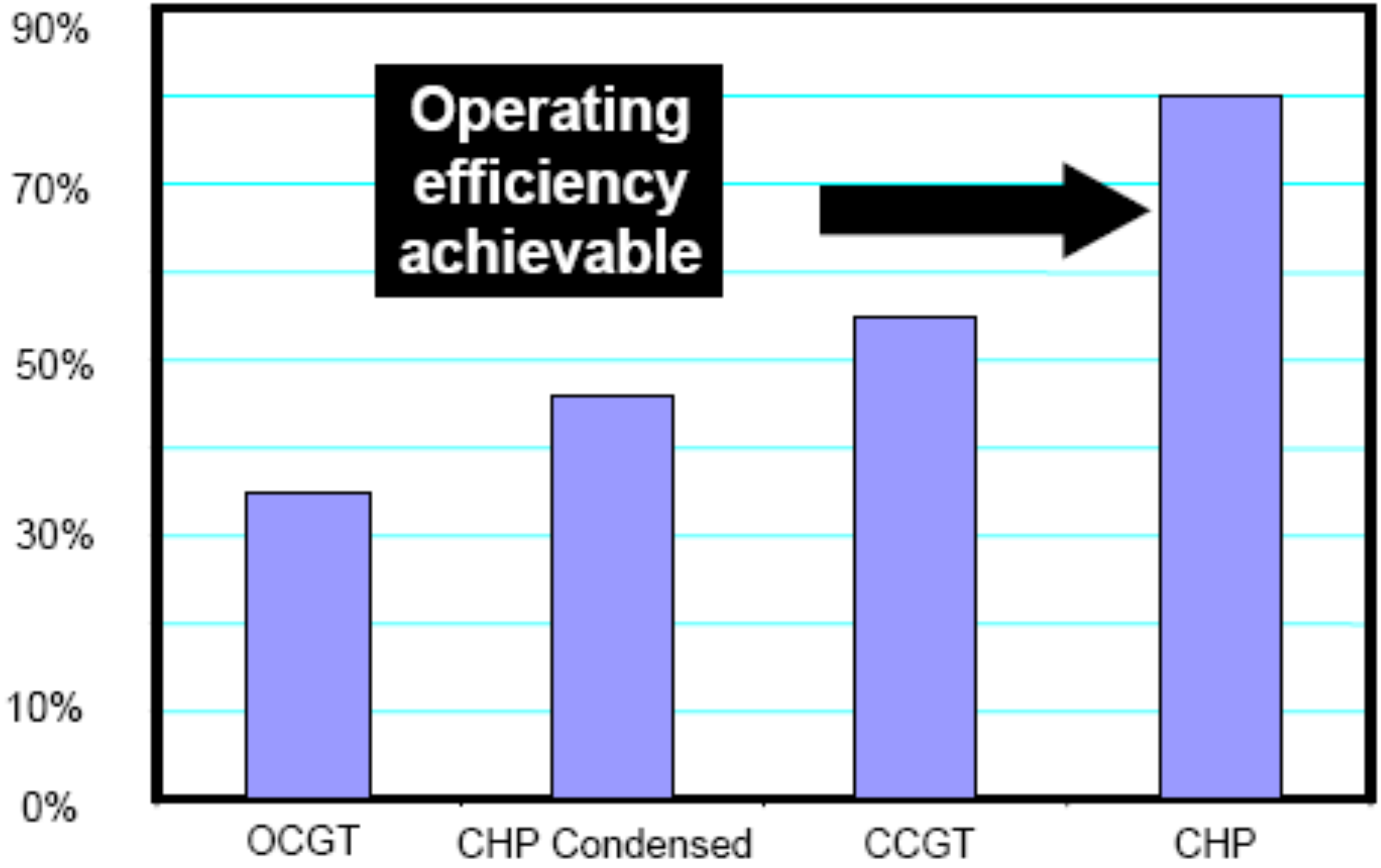
CCGT power plant

Matching power and water generation

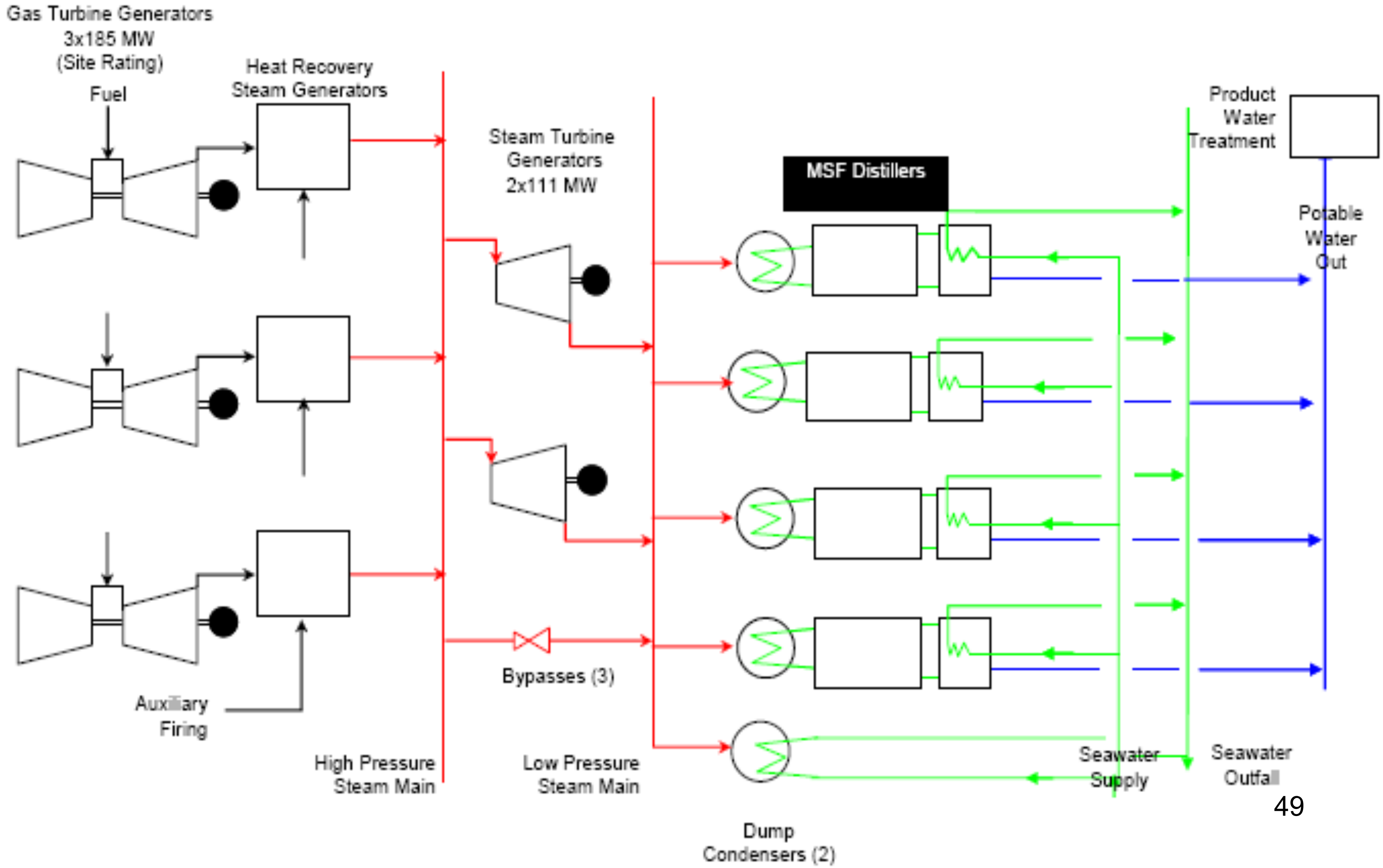


Combined heat and power (CHP) - **Cogeneration**

Operating efficiencies



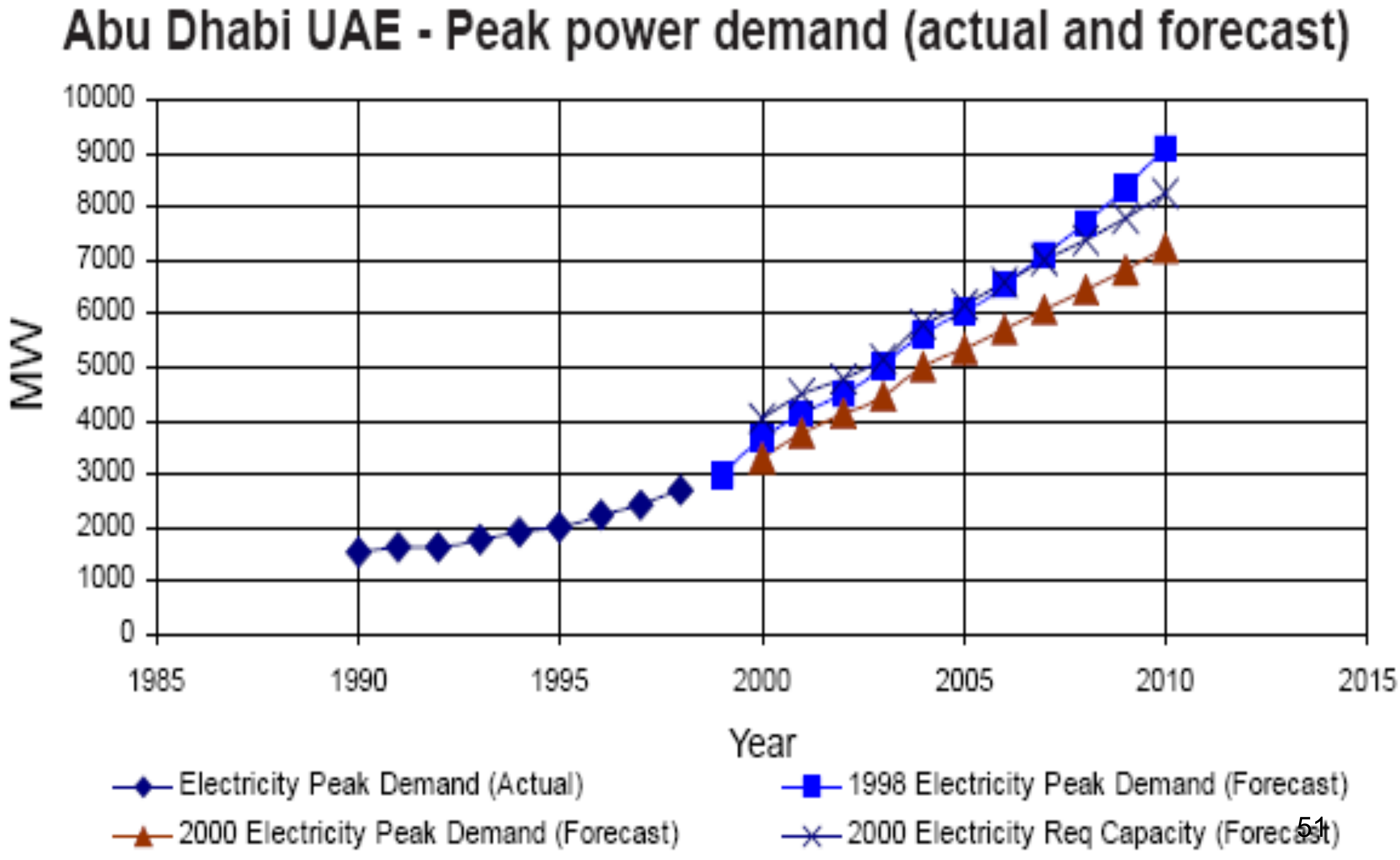
Combined cycle and MSF distillers block flow diagram



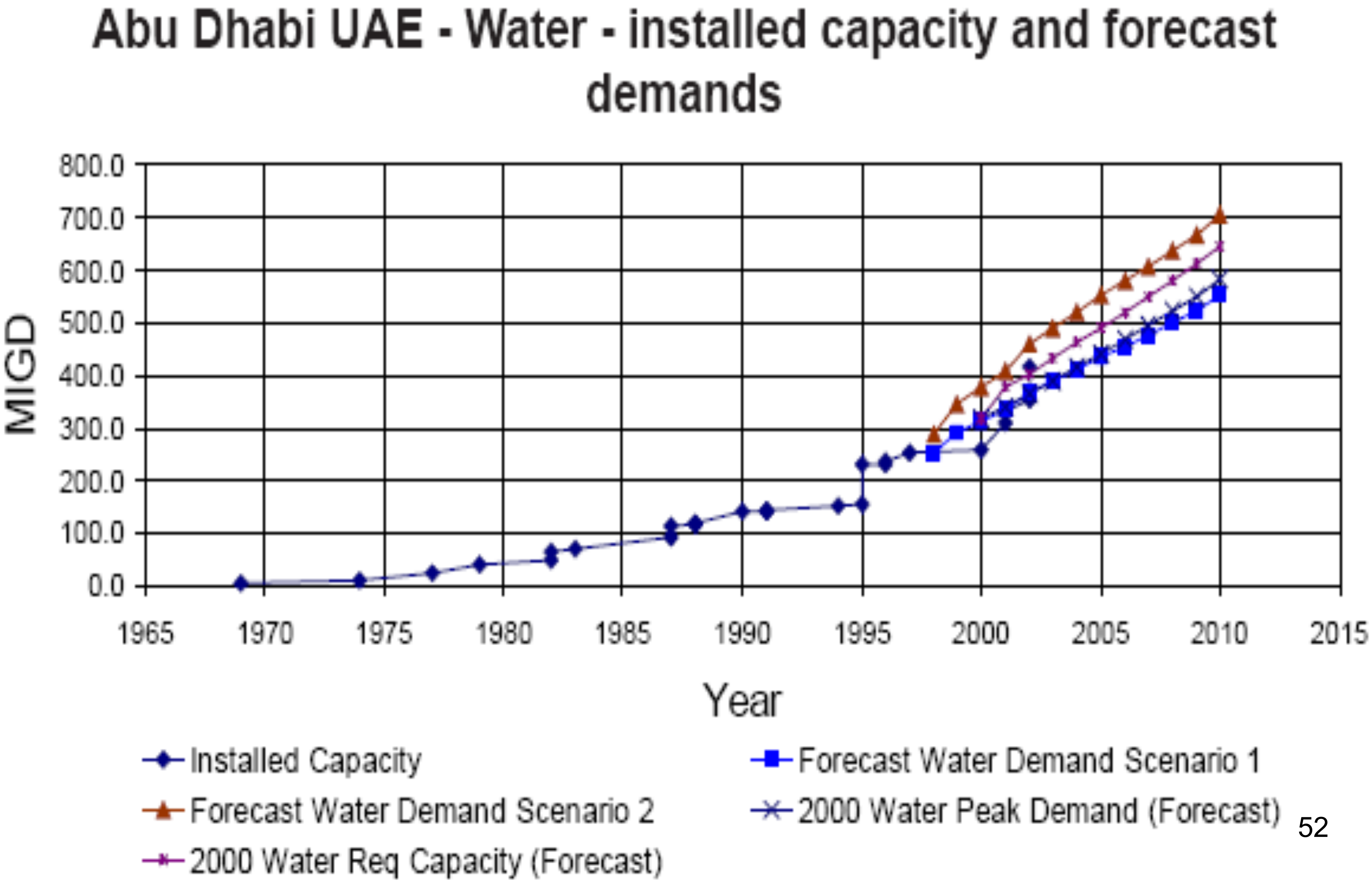
HYBRID SYSTEMS

- Optimising a combined cycle what is the maximum steam we can make available to desalination?
- From 1.2 to 1.5 ton/h of steam per each MW installed.
- How much water can be produced? Approximately 1 MIGD every 18 MW installed.

HYBRID SYSTEMS



HYBRID SYSTEMS



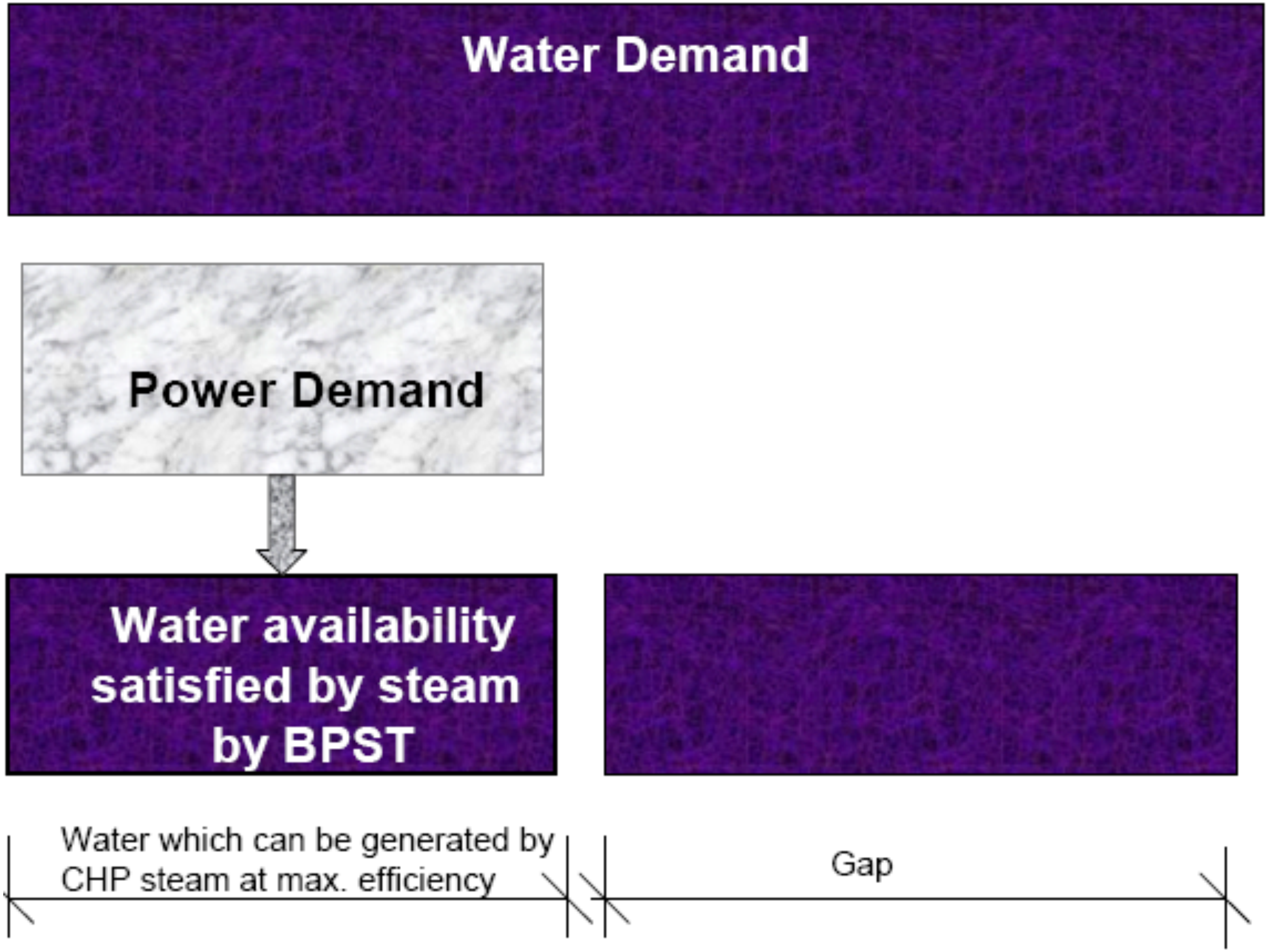
Situation analysis

Additional: 4000 MW, 350 MIGD Ratio MW/MIGD = 11–12.5 < 18

4000 MW can generate 250 MIGD of thermal desalination with an optimised CCGT.

There is no sufficient power demand to generate enough steam for the thermal desalination through a cogeneration process.

Situation analysis



Water Demand

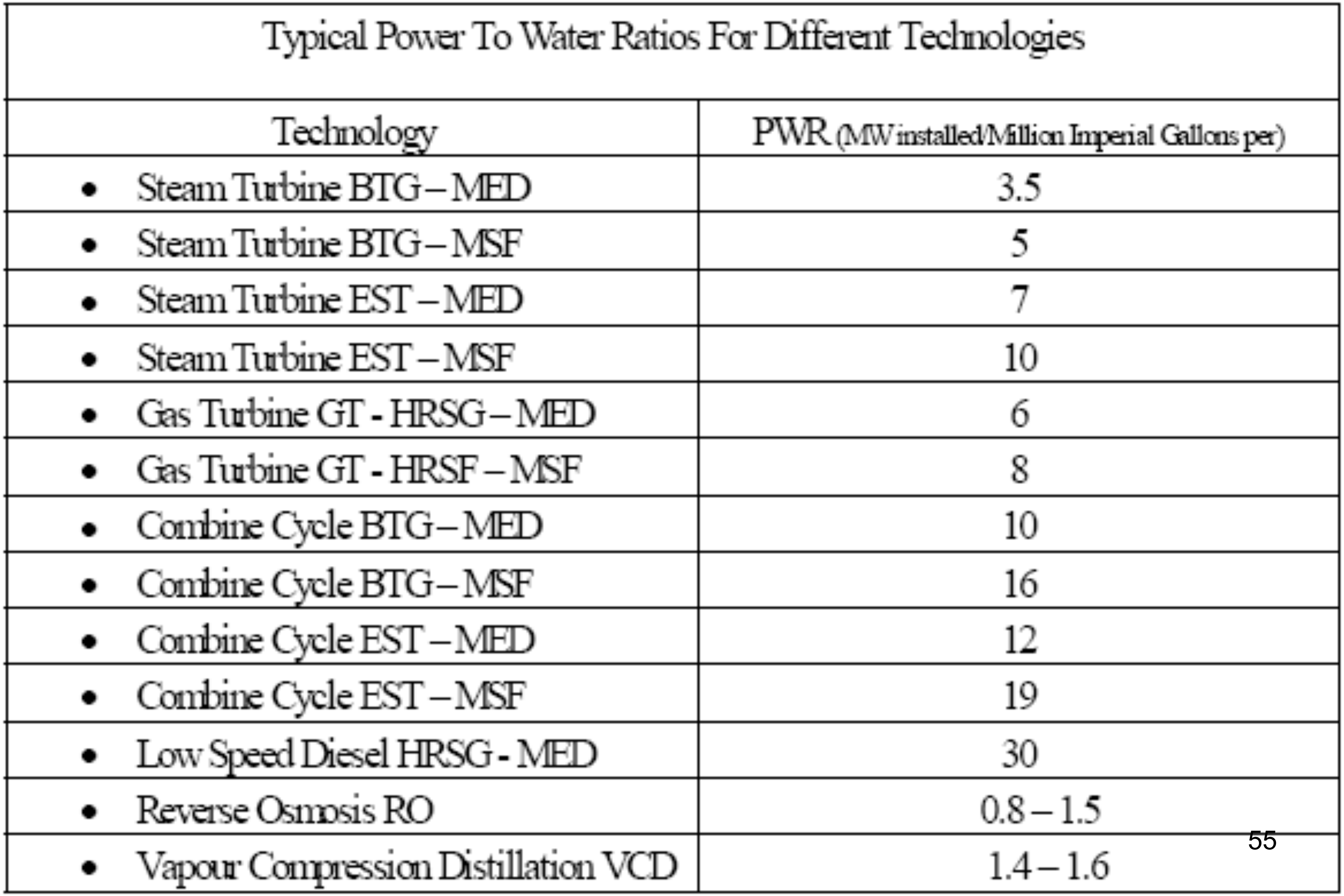
Power Demand

Water availability satisfied by steam by BPST

Water which can be generated by CHP steam at max. efficiency

Gap

Situation analysis

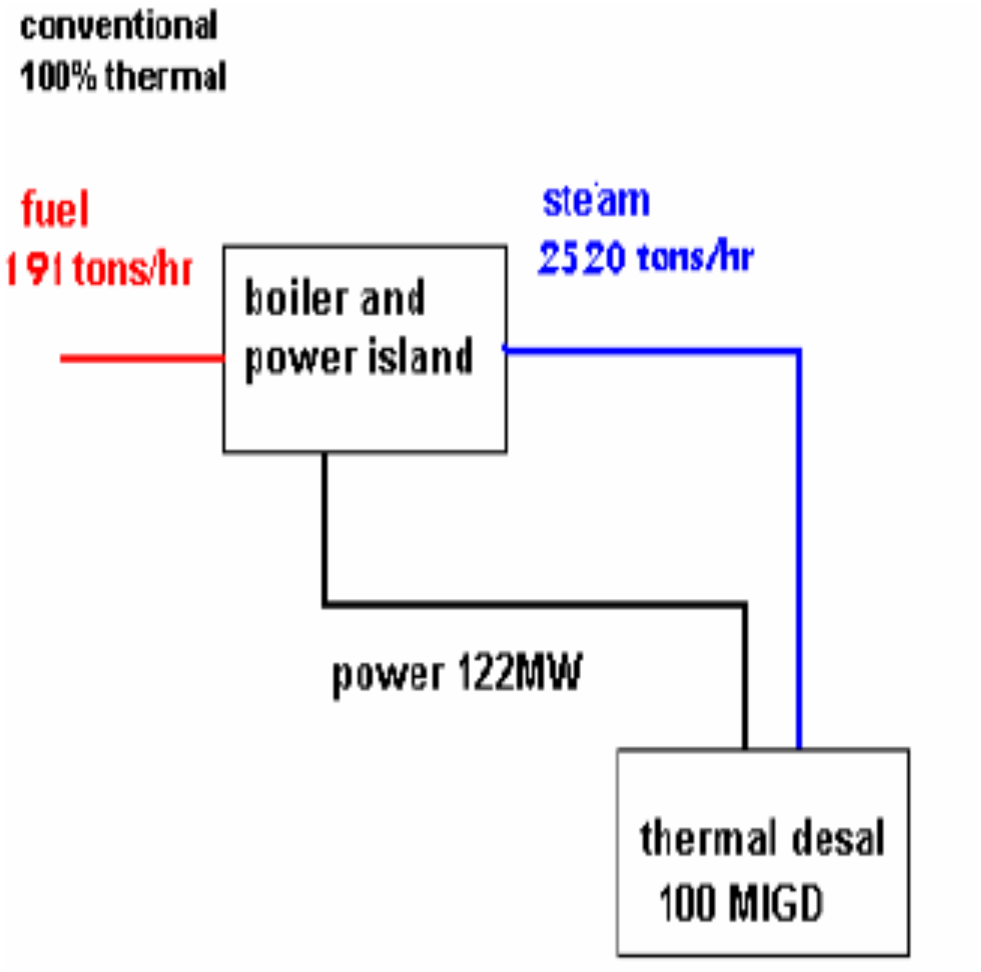


Typical Power To Water Ratios For Different Technologies

Technology	PWR (MW installed/Million Imperial Gallons per)
• Steam Turbine BTG – MED	3.5
• Steam Turbine BTG – MSF	5
• Steam Turbine EST – MED	7
• Steam Turbine EST – MSF	10
• Gas Turbine GT - HRSG – MED	6
• Gas Turbine GT - HRSF – MSF	8
• Combine Cycle BTG – MED	10
• Combine Cycle BTG – MSF	16
• Combine Cycle EST – MED	12
• Combine Cycle EST – MSF	19
• Low Speed Diesel HRSG - MED	30
• Reverse Osmosis RO	0.8 – 1.5
• Vapour Compression Distillation VCD	1.4 – 1.6

Case study

400 MW power + 100 MIGD : PWR = 4



200 ton/h of fuel
to produce

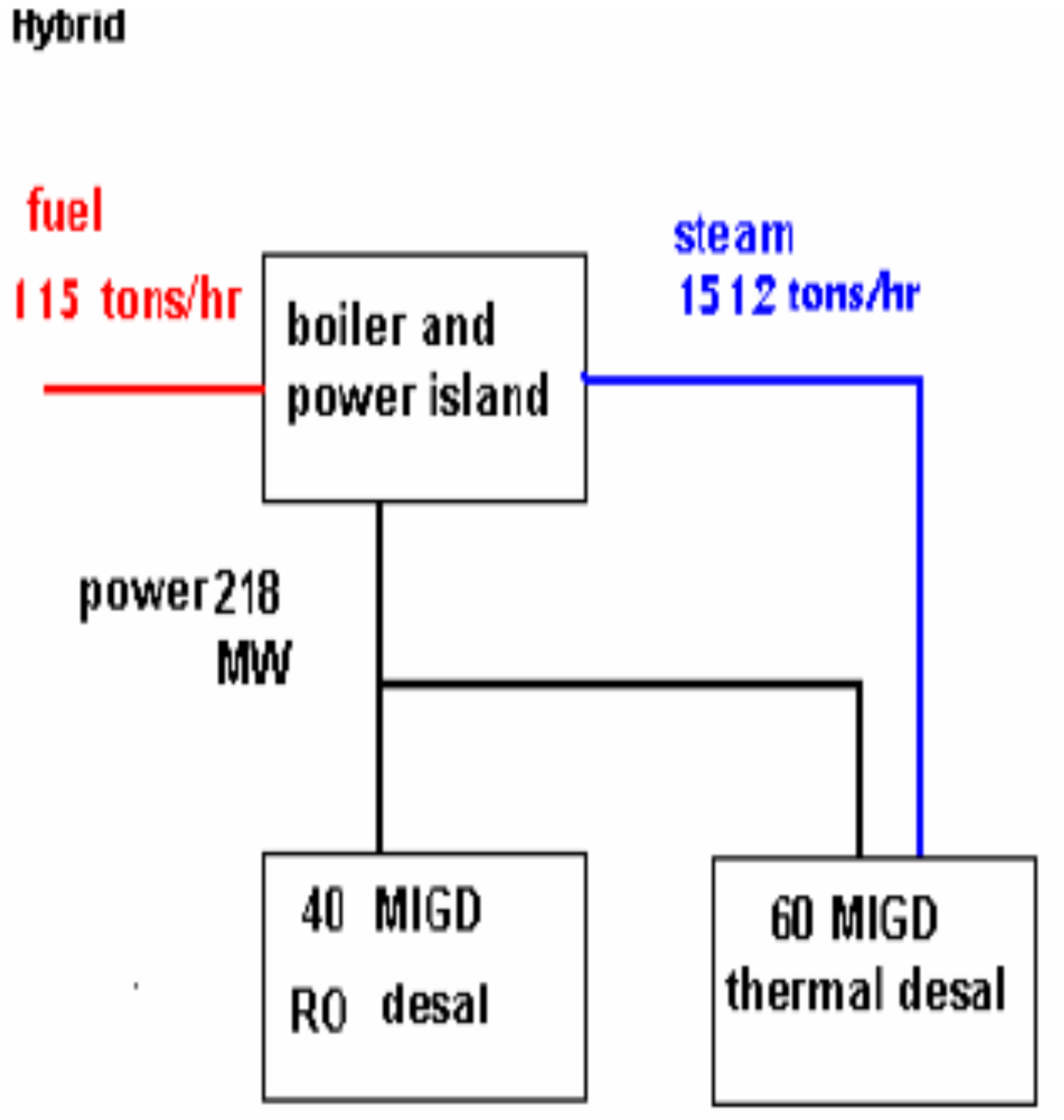
100 MIGD

80 MUS$/y

Option 1

100%thermal

Case study

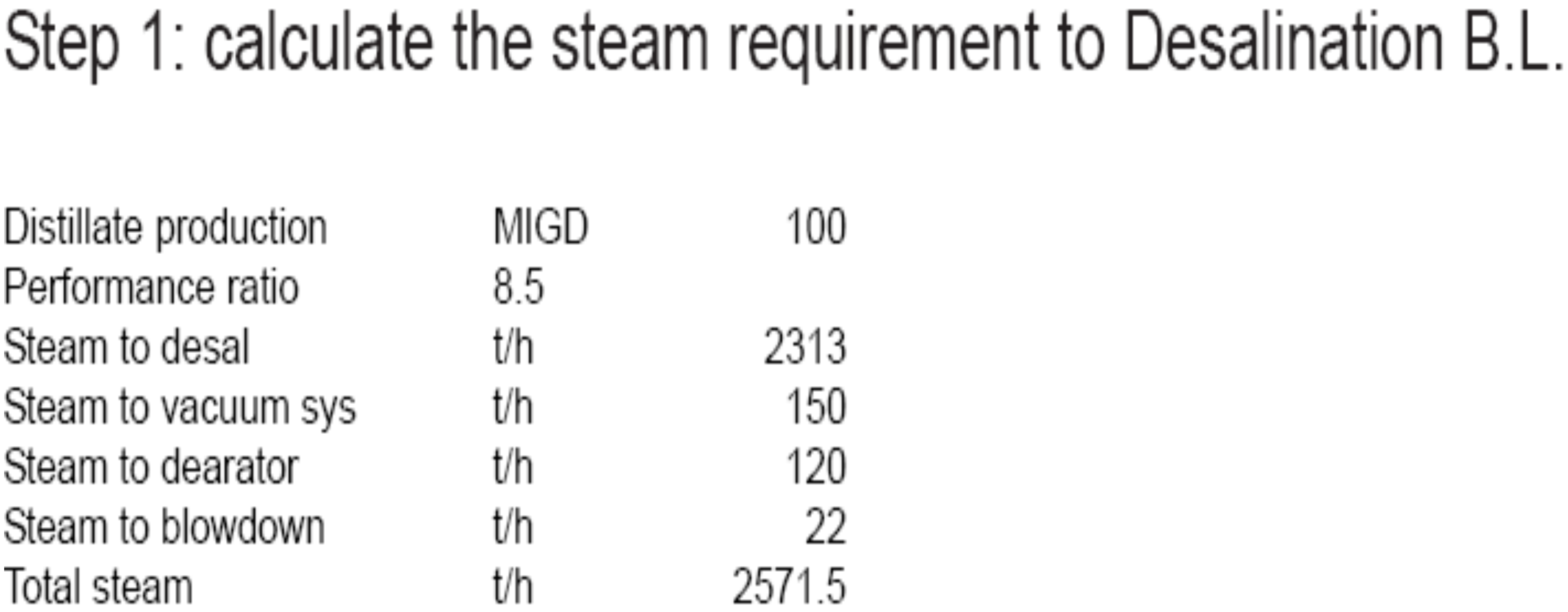


Fuel costs halved

Option 2

60%thermal

Working example



Step 1: calculate the steam requirement to Desalination B.L.

Distillate production	MIGD	100
Performance ratio	8.5	
Steam to desal	t/h	2313
Steam to vacuum sys	t/h	150
Steam to dearator	t/h	120
Steam to blowdown	t/h	22
Total steam	t/h	2571.5

Working example

distillate production $D := 100 \cdot \mathrm{MIGD}$

$$D = 18900 \cdot \frac{\mathrm{tonne}}{\mathrm{hr}}$$

performance ratio $\eta := 8.5$

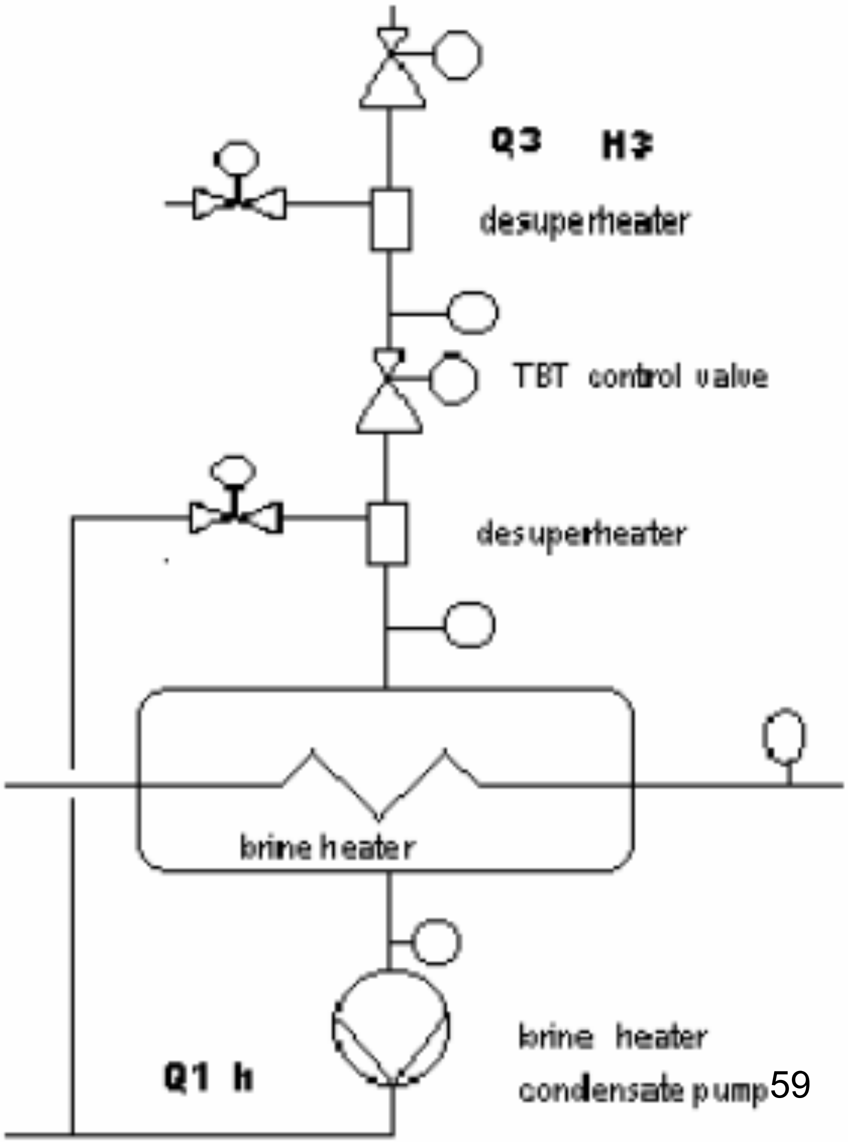
reference enthalpy $\mathrm{Href} := 2326 \cdot \frac{\mathrm{kj}}{\mathrm{kg}}$

Condensate enthalpy $h := 514 \cdot \frac{\mathrm{kj}}{\mathrm{kg}}$

steam to brine heater enthalpy $\mathrm{H1} := 2750 \cdot \frac{\mathrm{kj}}{\mathrm{kg}}$

steam flow to be achieved

$$\mathrm{Q1} := \frac{D \cdot (\mathrm{Href})}{\eta \cdot (\mathrm{H1} - h)} \qquad \mathrm{Q1} = 2.313 \cdot 10^{3} \cdot \frac{\mathrm{tonne}}{\mathrm{hr}}$$



Working example

Step 2: calculate the fuel requirement

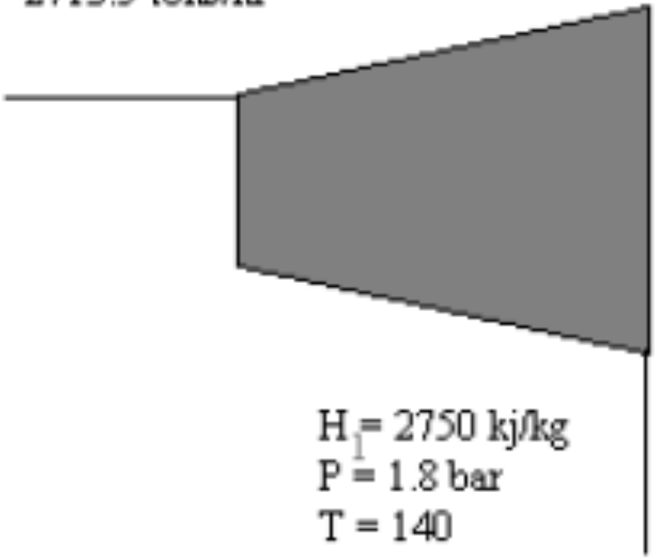
ST outlet enthalpy	kJ/kg	2750
ST inlet enthalpy	kJ/kg	3563
Heat required	GJ/h	9668.201
Fuel LCV	Btu/ncft	947
Conversion	kJ/m^3	37.259
Fuel LCV	kJ/m^3	35284.27
Fuel required	nm^3/h	274008.8

H_0 =3563 kj/kg

P = 60 bar

T = 560

steam flow = 2713.5 tons/hr



H_1 = 2750 kj/kg

P = 1.8 bar

T = 140

steam flow = 2713.5 tons/hr

$$H0 := 3563 \cdot \frac{kj}{kg}$$

$$H := Qt \cdot (H0)$$

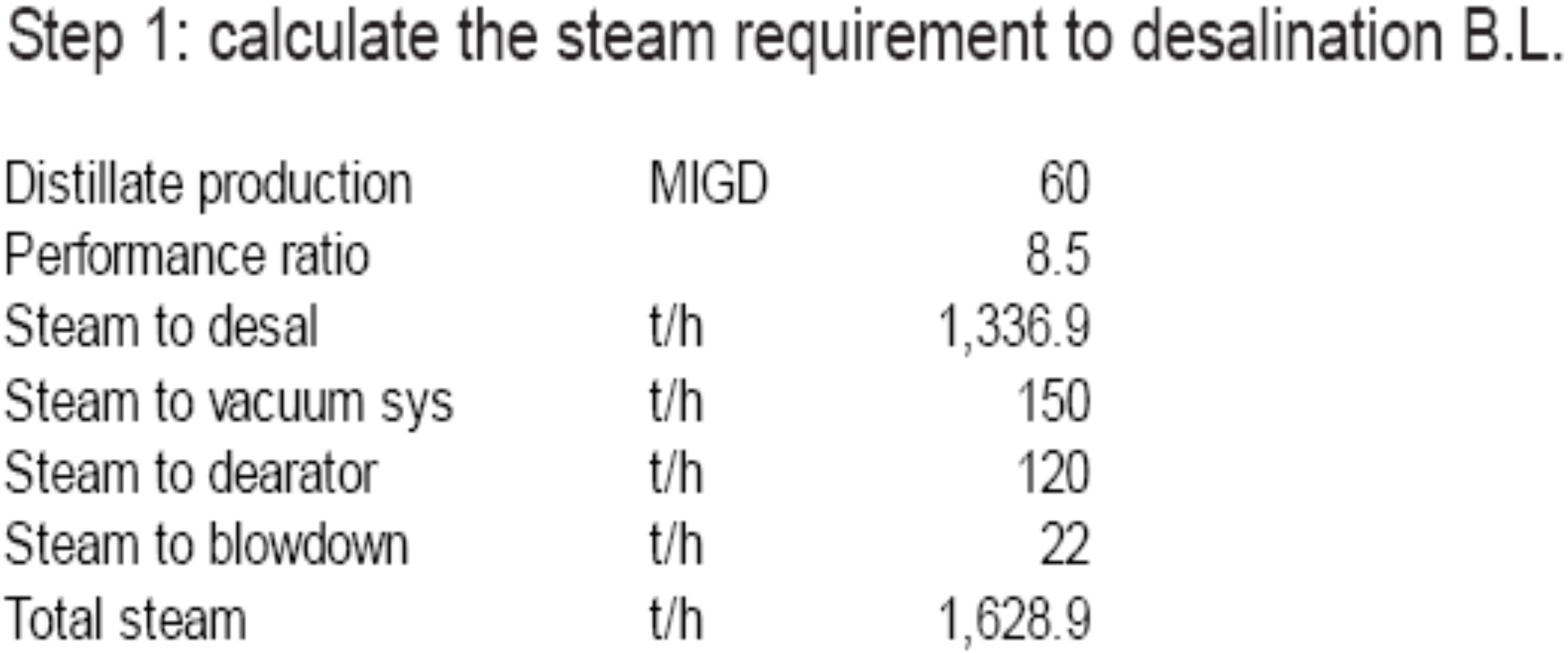
$$H = 9.668 \cdot 10^3 \cdot \frac{GJ}{hr}$$

Working example

Step 3: calculate the fuel costs

Fuel required	nm^3/h	274009
Fuel cost	US$/GJ	1.1
Annual fuel costs	US$/y	93160848

Case 2: 100 MIGD 60 % thermal, 40% RO



Step 1: calculate the steam requirement to desalination B.L.

Distillate production	MIGD	60
Performance ratio		8.5
Steam to desal	t/h	1,336.9
Steam to vacuum sys	t/h	150
Steam to dearator	t/h	120
Steam to blowdown	t/h	22
Total steam	t/h	1,628.9

Case 2: 100 MIGD 60 % thermal, 40% RO

Step 2: calculate the fuel requirement

Total steam	t/h	1,628.9
ST outlet enthalpy	kJ/kg	2840
ST inlet enthalpy	kJ/kg	3557.83
Heat required	GJ/h	5795.344
Fuel LCV	Btu/ncft	947
Conversion	Btu/ncft	37.259
Fuel LCV	kJ/m^3	35284.27
Fuel required	nm^3/h	164247.2

Step 3: calculate the fuel costs

Heat required	GJ/h	5795.344
Fuel required	nm^3/h	164247.2
Fuel cost	US$/GJ	1.1
Annual fuel costs	US$/y	55843931

Commentary

- Case 1 would still be able to produce more than 300 MW (i.e.1100 MR 800 MW) against more fuel
- The power however would be idle

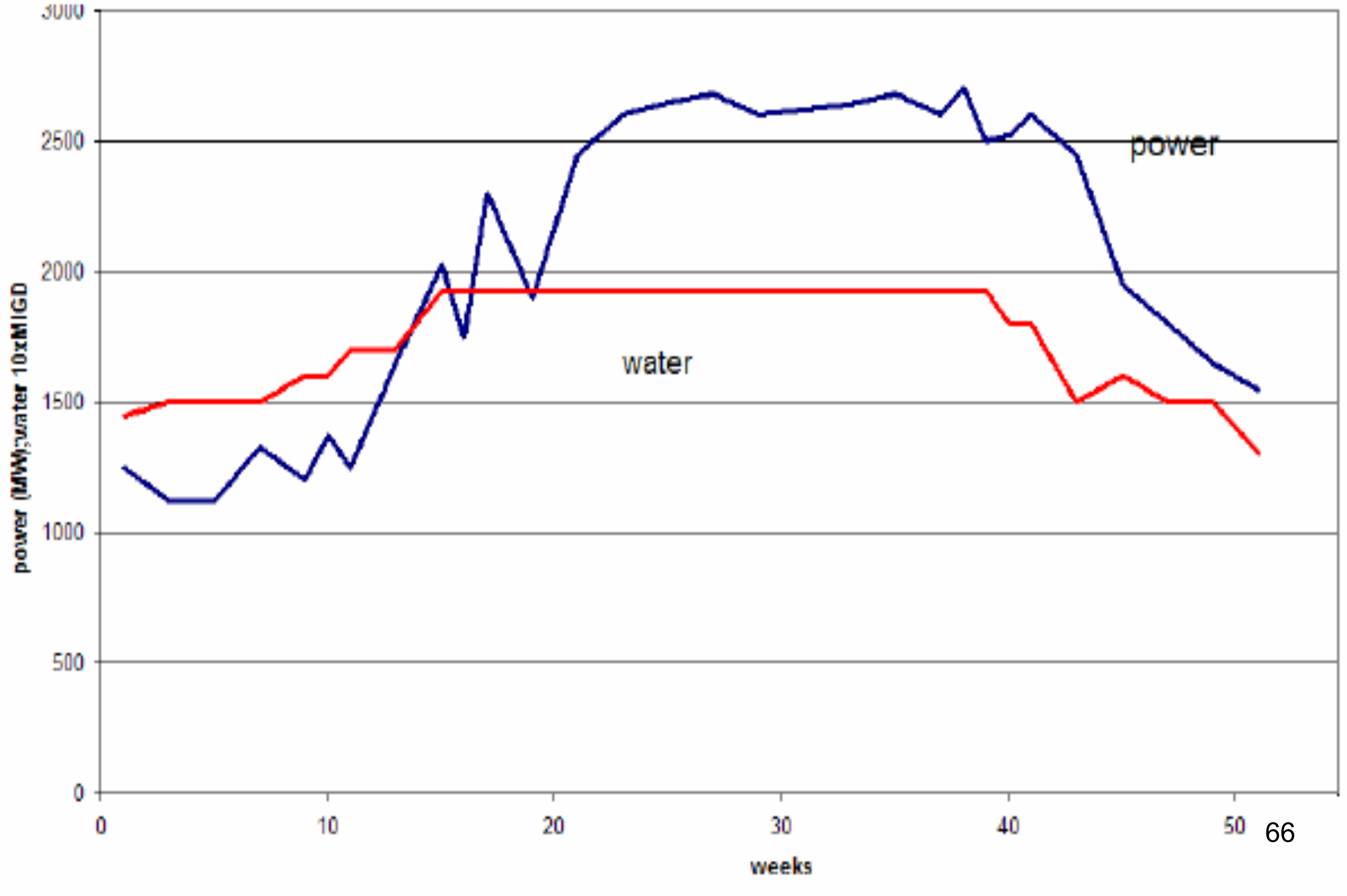
HYBRID SYSTEMS

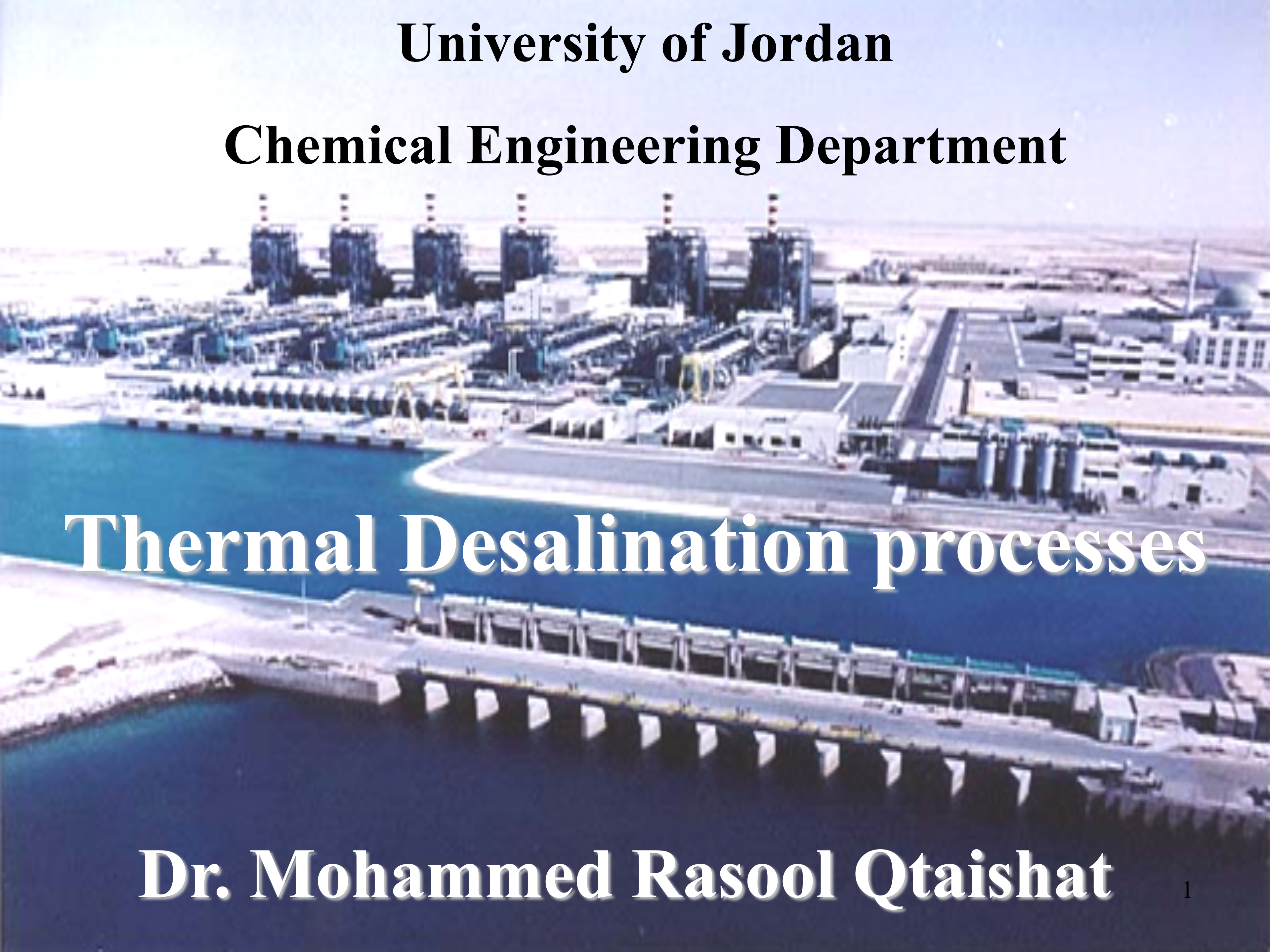
In winter operation different power and water network requirements occur:

Limited flexibility is allowed by CHP system in winter configuration

Operational configurations becomes very inefficient

Annual power and water requirements





University of Jordan

Chemical Engineering Department

Thermal Desalination processes

Dr. Mohammed Rasool Qtaishat

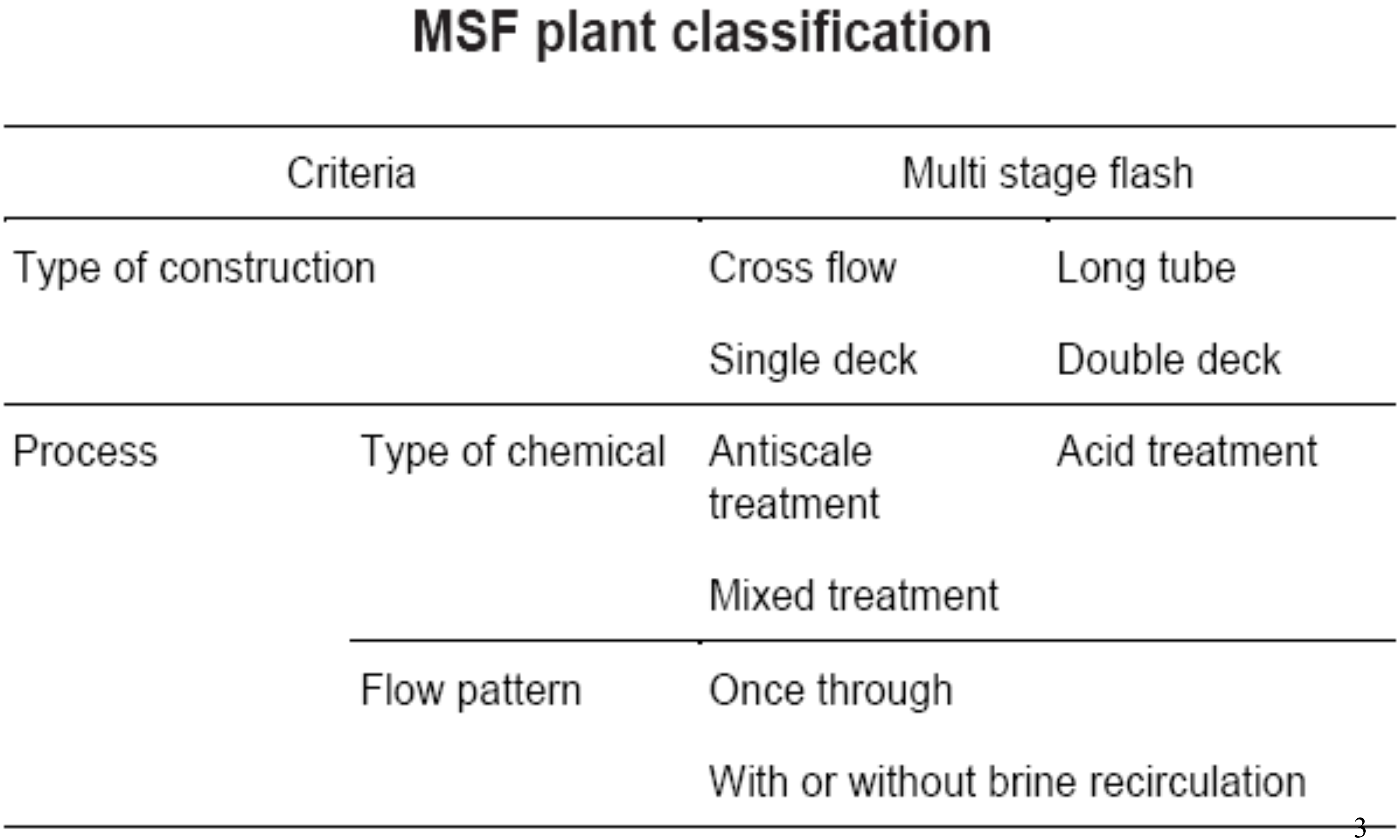
MULTISTAGE FLASH TECHNOLOGY (MSF)

- **Process description**
- **Process thermodynamics**
- **Stage simulation model**

Terminology

TBT — Top brine temperature
BBT — Bottom brine temperature
TBT-BBT — Flashing range
TDS — Total dissolved solids

MSF Plant Classification



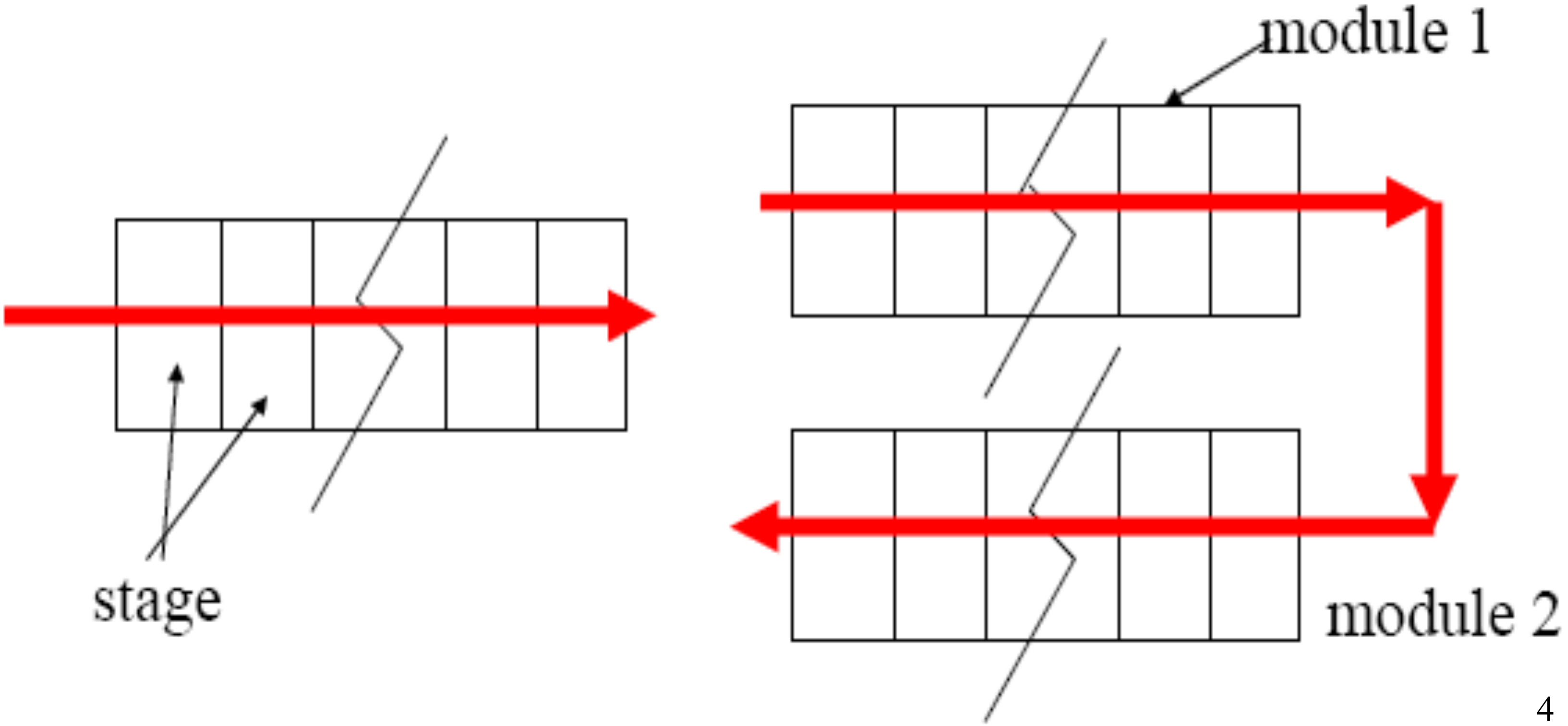
MSF plant classification

Criteria		Multi stage flash	
Type of construction		Cross flow	Long tube
		Single deck	Double deck
Process	Type of chemical	Antiscale treatment	Acid treatment
		Mixed treatment	
	Flow pattern	Once through	
		With or without brine recirculation	

Construction Criteria

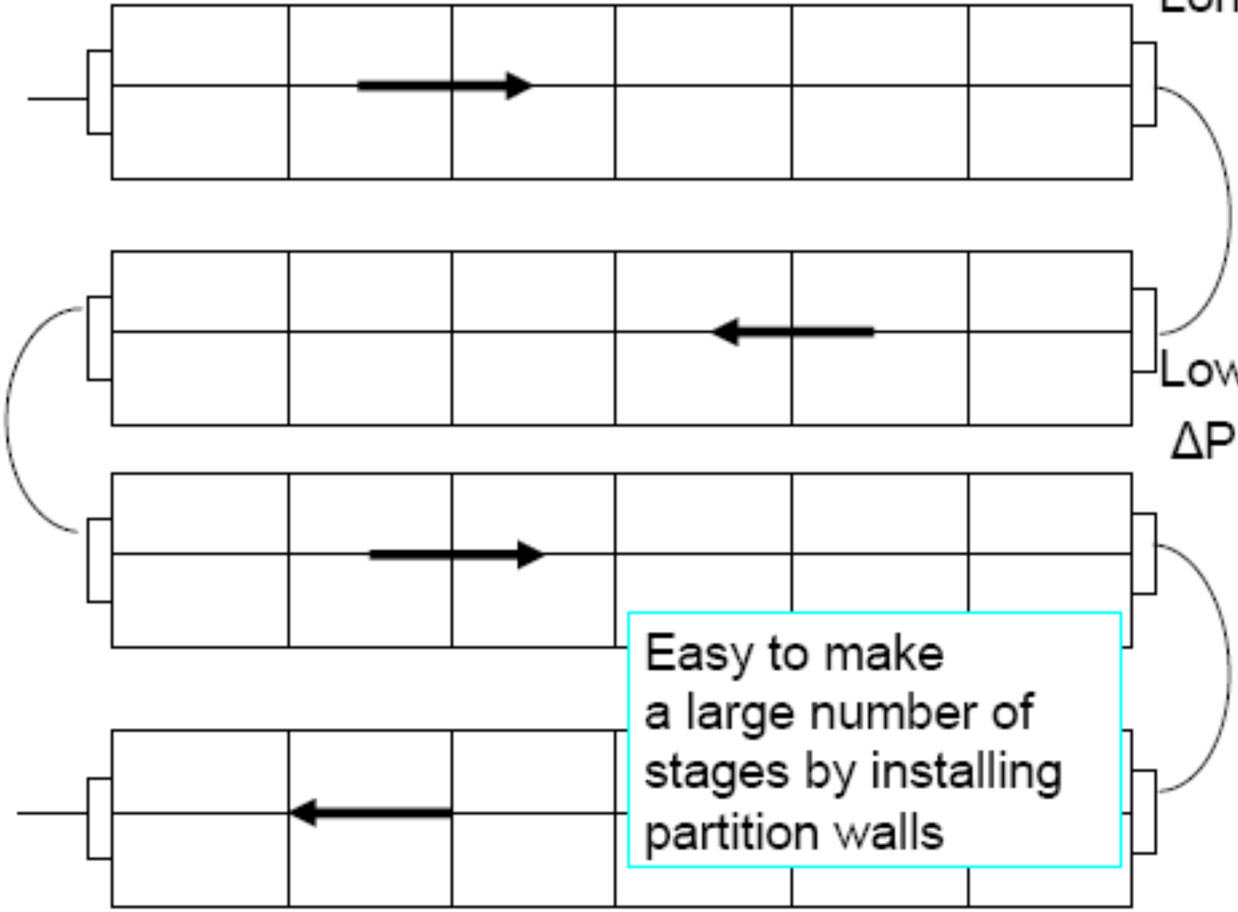
Construction criteria

Long tubes: distinguish between the stages and the module



MSF Plant Classification

MSF plant classification



Long tube

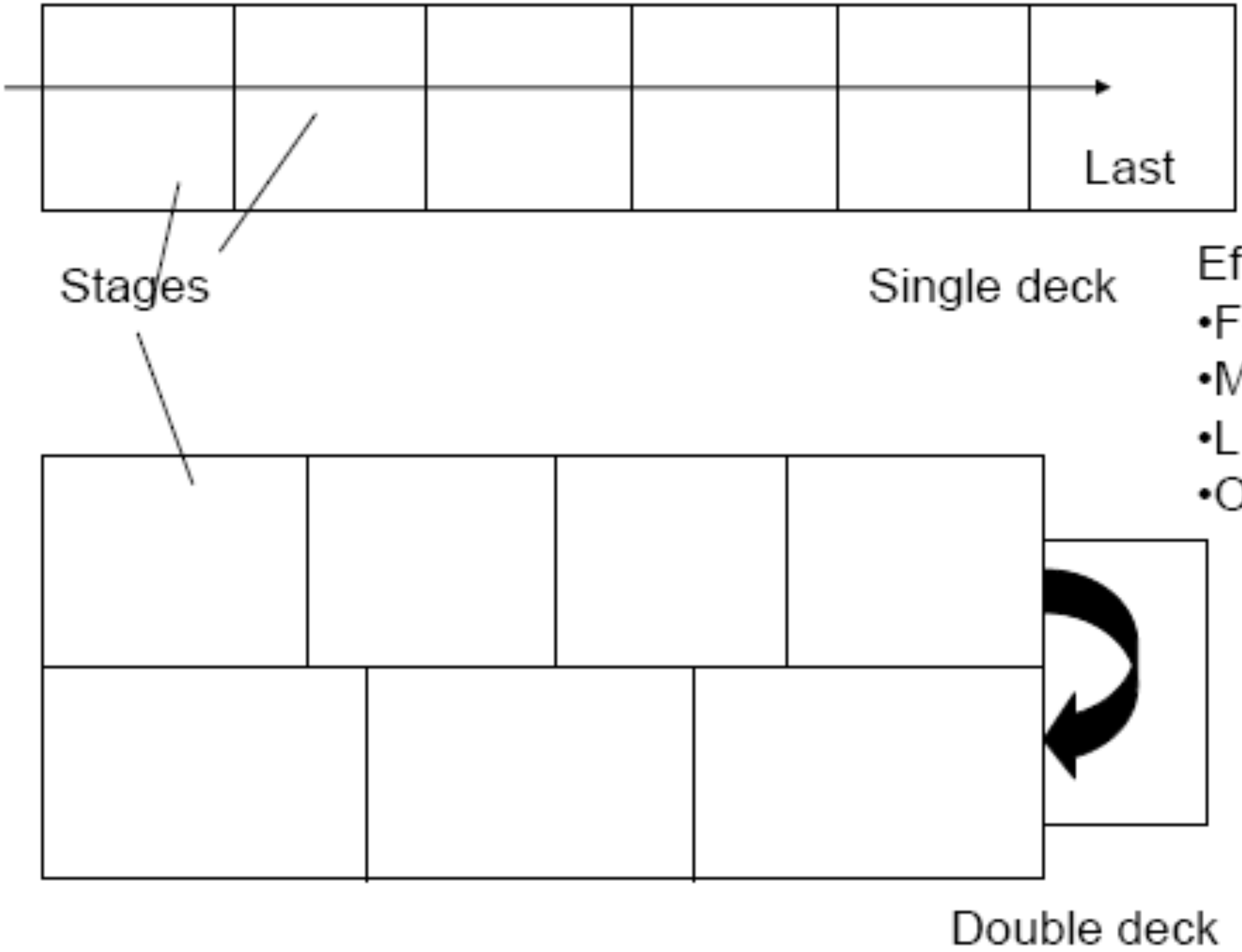
Lower ΔP

$\Delta P \infty v^2 1/d$

Easy to make
a large number of
stages by installing
partition walls

Construction Criteria

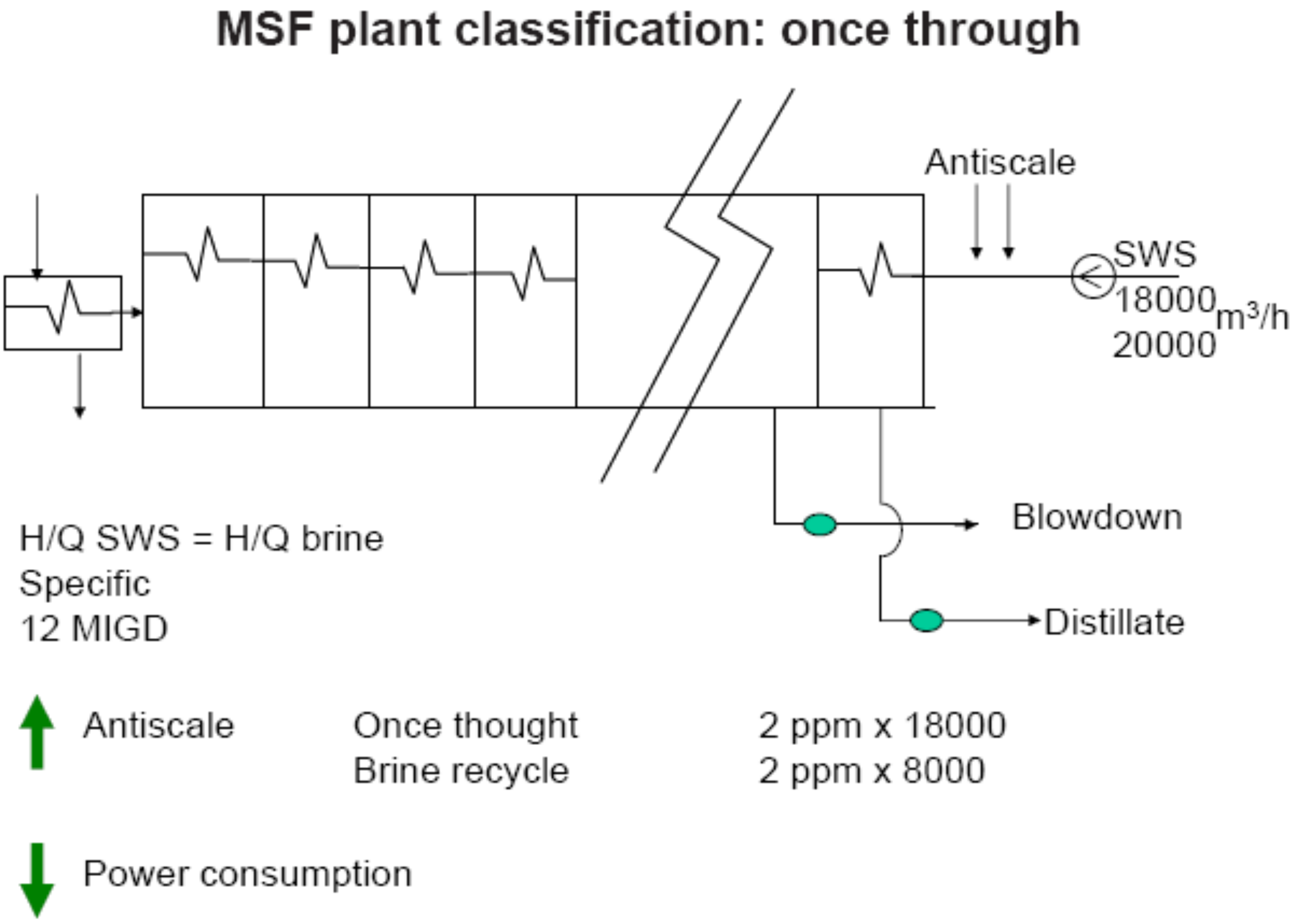
Construction criteria



Effects:

- Footprint
- Maintenance
- LIT 1st last stage
- Overall size

MSF Plant Classification: once through



MSF plant classification: once through

H/Q SWS = H/Q brine
Specific
12 MIGD

↑	Antiscale	Once thought Brine recycle	2 ppm x 18000 2 ppm x 8000

↓ Power consumption

MULTISTAGE FLASH TECHNOLOGY (MSF)

- **MSF what do we know?**

- Highly reliable operation
- Scalable up to very large sizes 18 MIGD
- Readily coupled with steam turbine generating stations in “dual purpose plant” configuration
- Good water to power ratio
- A big and well-deserved success since the 1960s

Process description

Process description: How did it begin?

- It had long been known that water could be heated above its normal boiling point in a pressurized system.

- If the pressure was released, a portion of the water would boil off or “flash”. The remaining liquid water would be cooled as the issuing vapor took with it its heat of vaporization.

- Since evaporation occurred from the bulk fluid rather than at a hot heat exchange surface, opportunities for scaling would be reduced.

MULTISTAGE FLASH TECHNOLOGY (MSF)

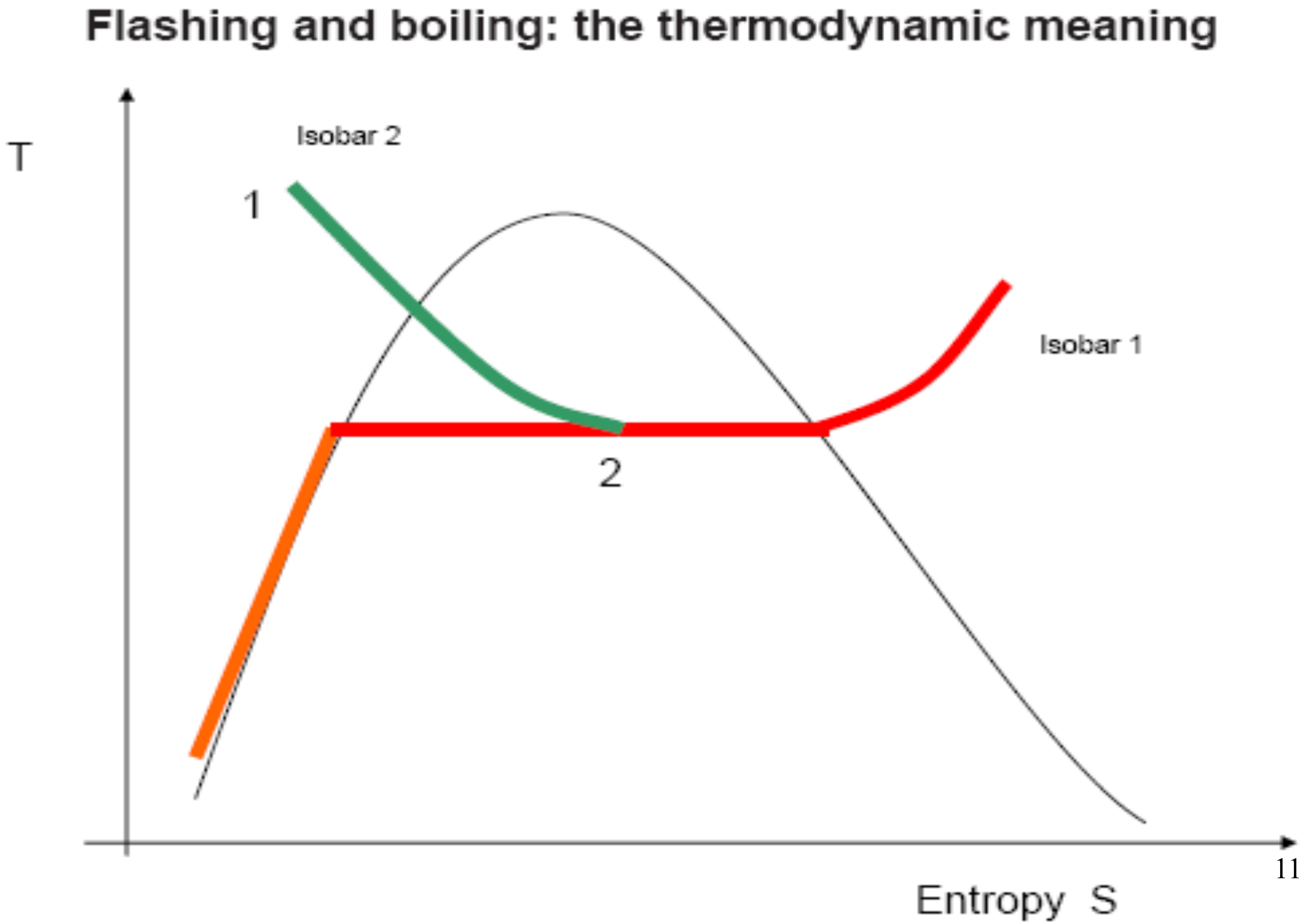
- **What flashing looks like**

- Hot brine from the previous stage enters through slot at lower temperature and pressure stage

- It senses the new lower pressure environment, and

- Flashes!

Flashing and Boiling



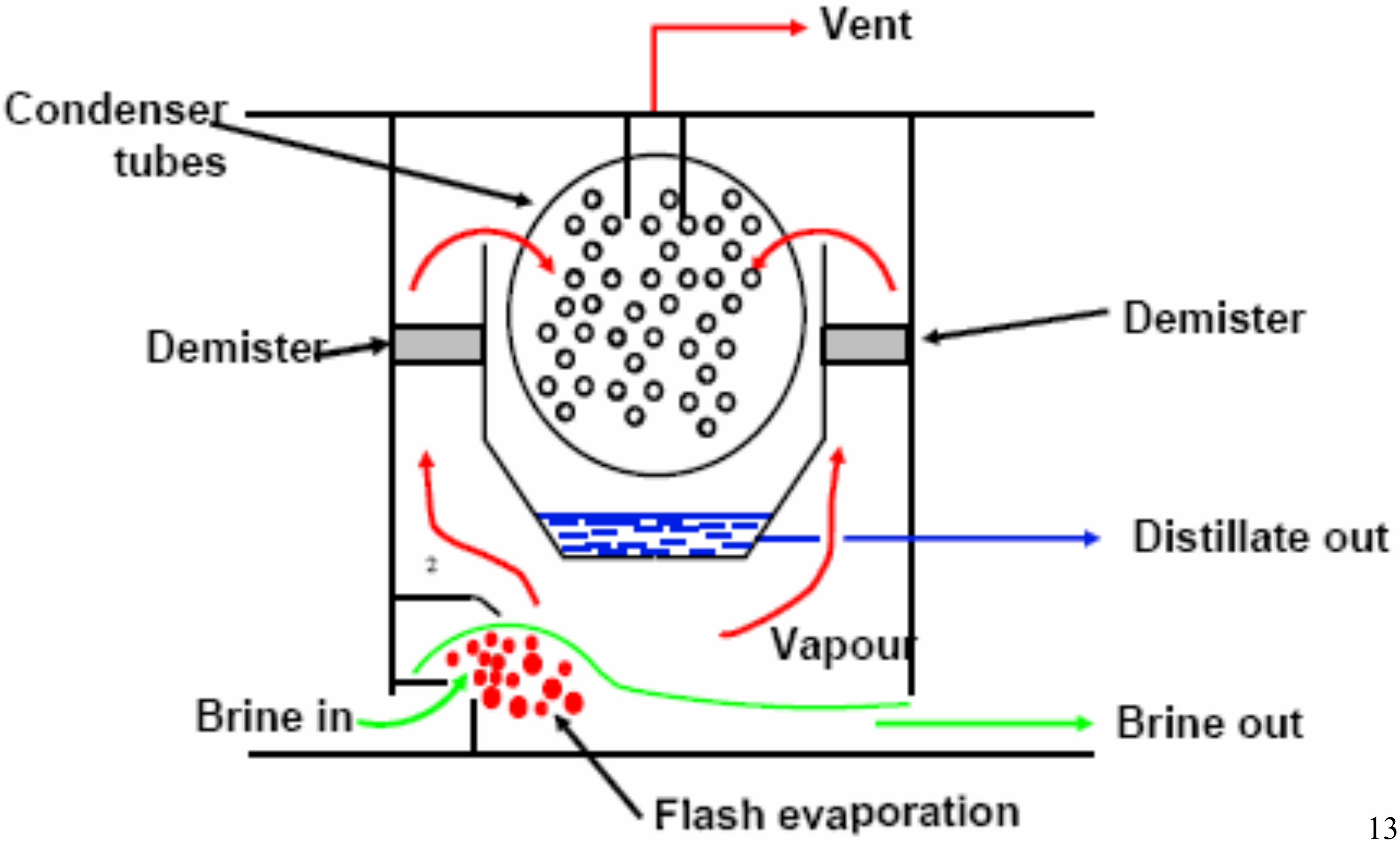
MSF Development

- **MSF development**

- Cross tube design — tube length limitations
- Long tube design
- Once through process
- Optimise structural design to reduce shell plate thickness and weight
- Solid stainless steel shell construction
- Thinner heat transfer tubes

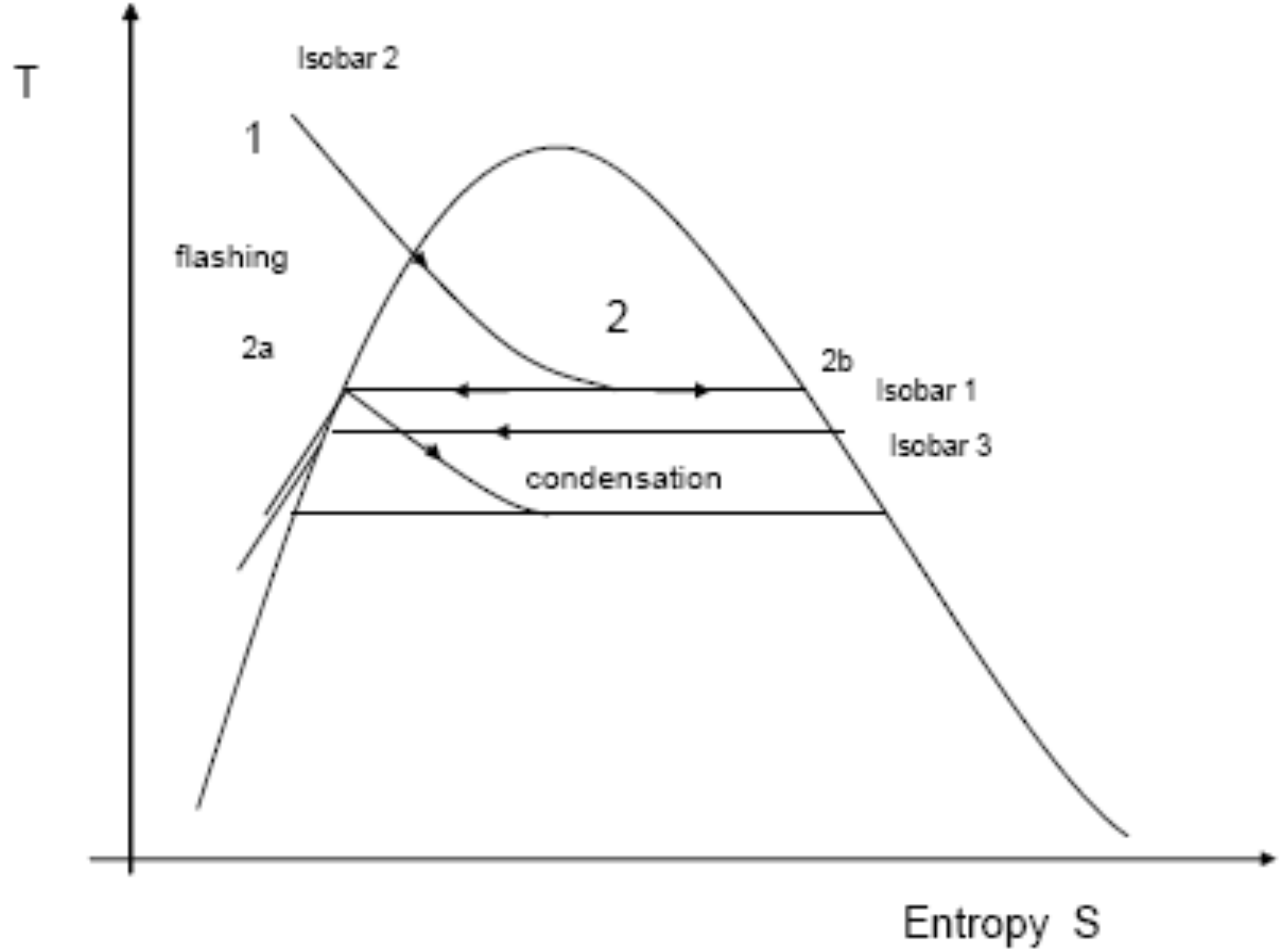
MSF Desalination Plant

Typical stage arrangement of a large MSF plant



MSF Desalination Plant

Stage modeling thermodynamic ideal case



Dissolved organic

- **The influence of minor constituents of seawater and brackish waters**

A. Dissolved inorganic

If seawater consisted of only H_2O and NaCl, life would be simple. But natural waters are often close to saturation in many inorganic compounds ($CaSO_4$, $Mg(OH)_2$, $Ca(HCO_3)_2$, etc.).

What is worse, their solubility may be inverse functions of temperature.

This involves the following aspects to be considered:

– scaling

– venting

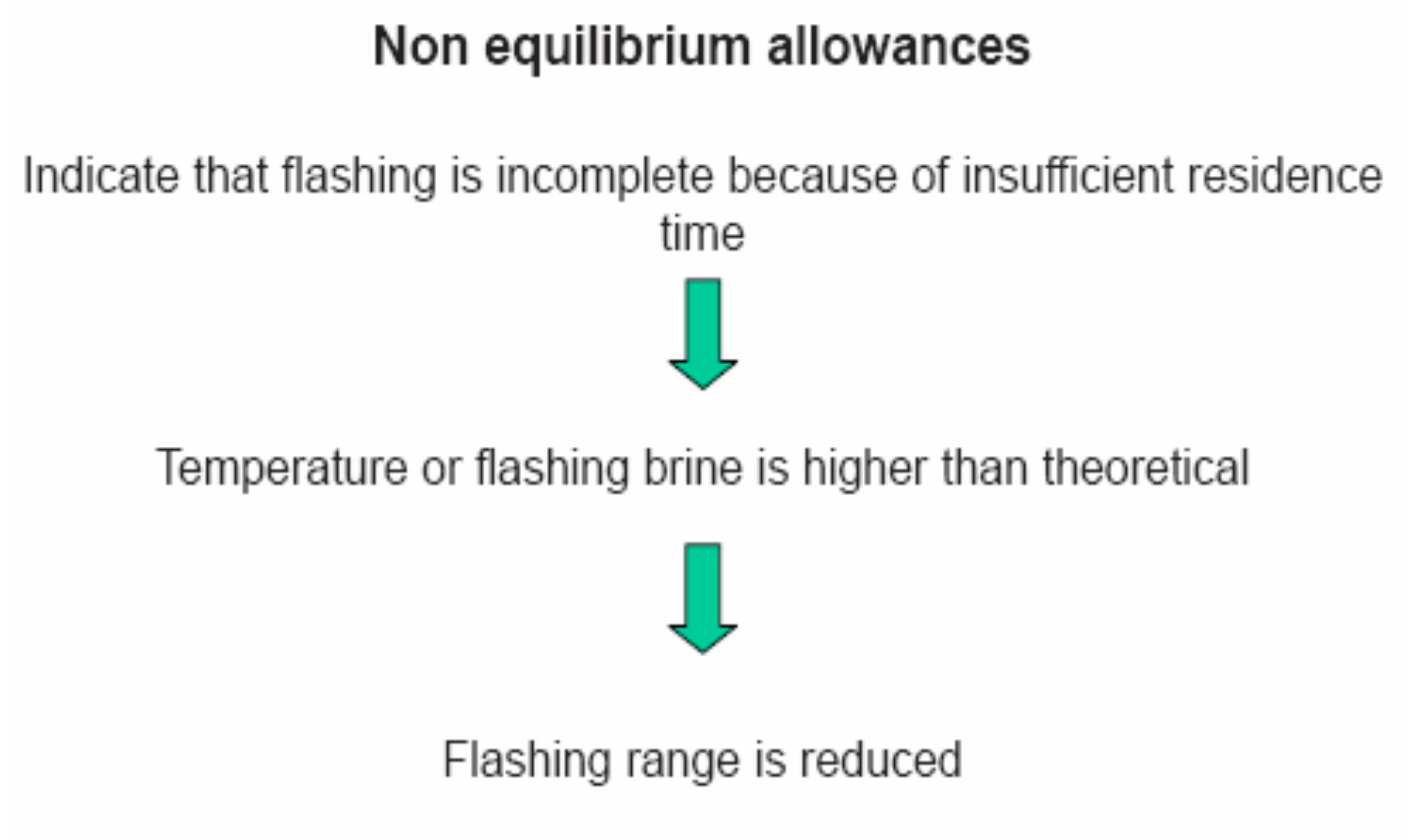
Parasitic losses

➢ Parasitic losses

- Demisters ΔP
- Demisters ΔT
- Tube bundle (ΔT)
- Distribution ΔT
- Non-equilibrium allowances
- Others, i.e. incondensable losses

Parasitic losses + Stage! High top temperature
Low bottom temperature

Non Equilibrium allowances



Non equilibrium allowances

Indicate that flashing is incomplete because of insufficient residence time

↓

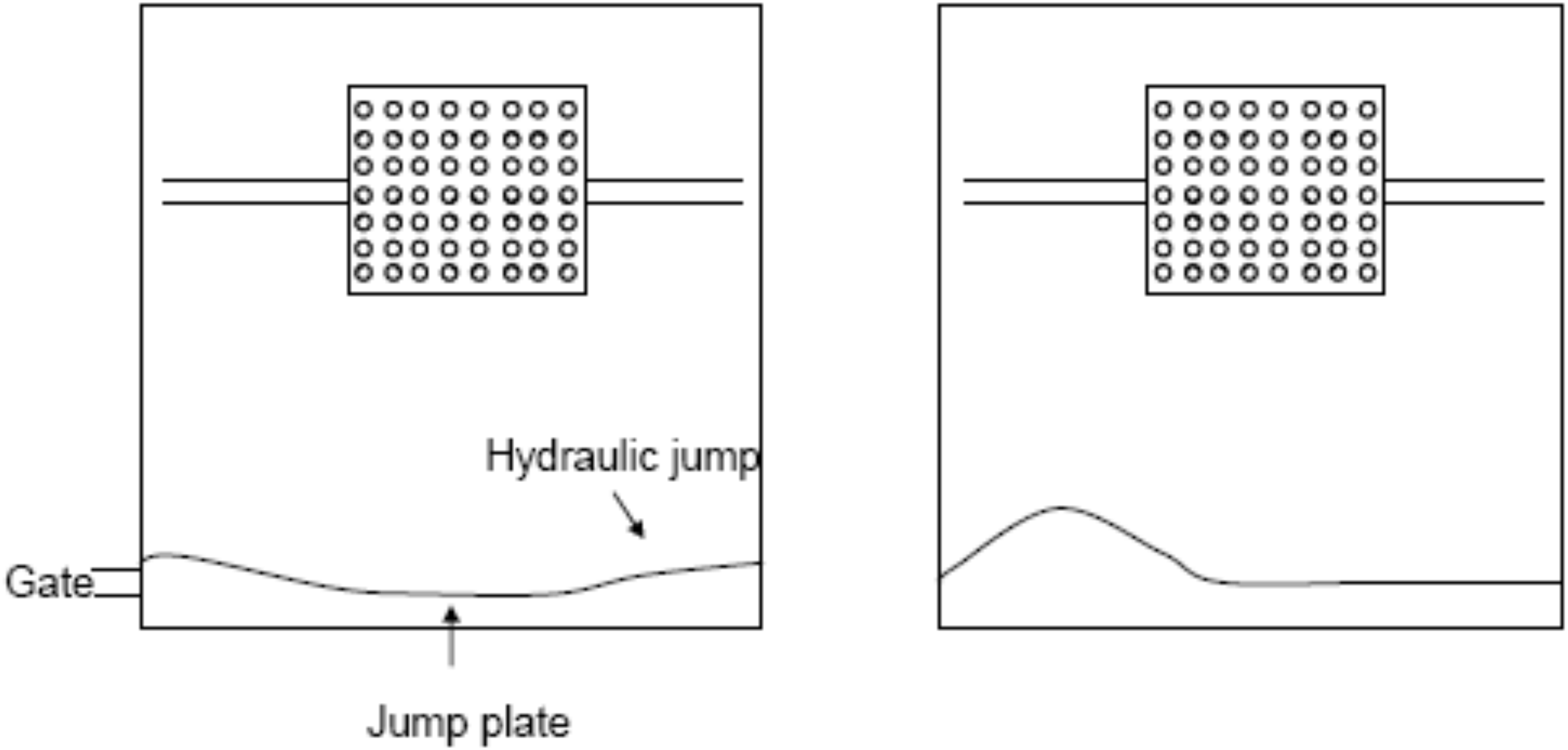
Temperature or flashing brine is higher than theoretical

↓

Flashing range is reduced

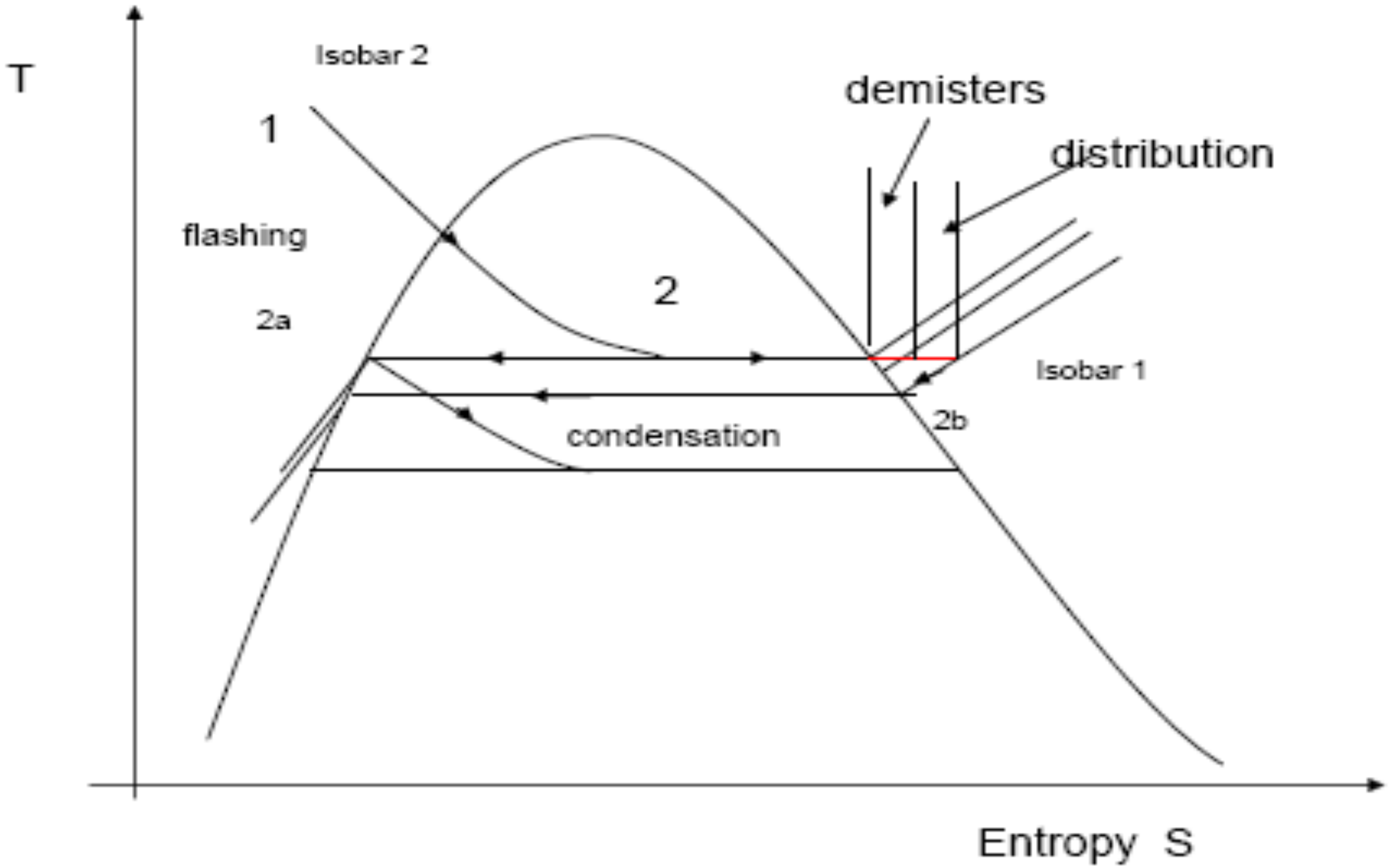
Non Equilibrium allowances

Different designs to minimise brine levels in the evaporation



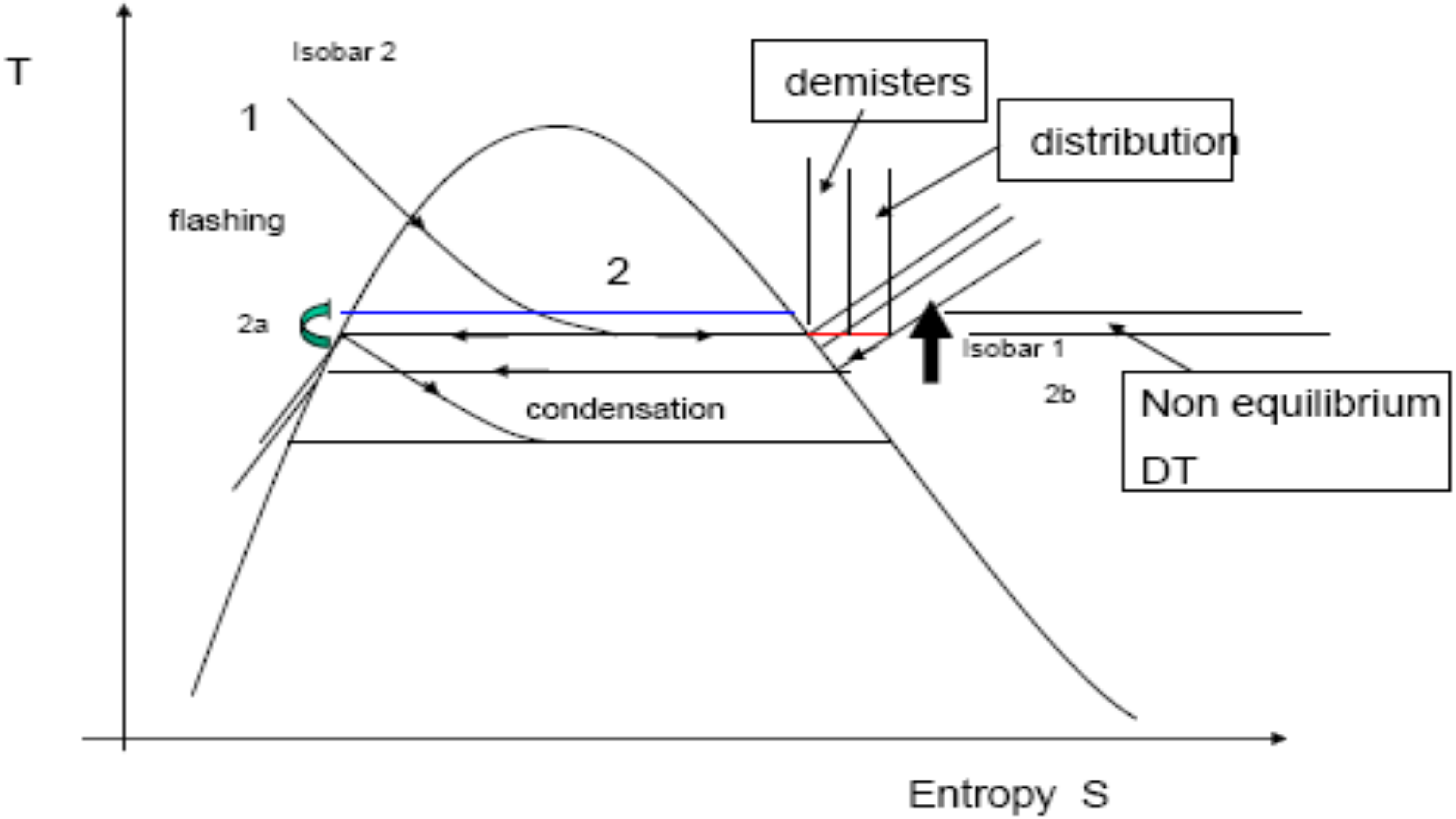
Stage Modeling Thermodynamic

Stage modeling thermodynamic: understanding parasitic losses



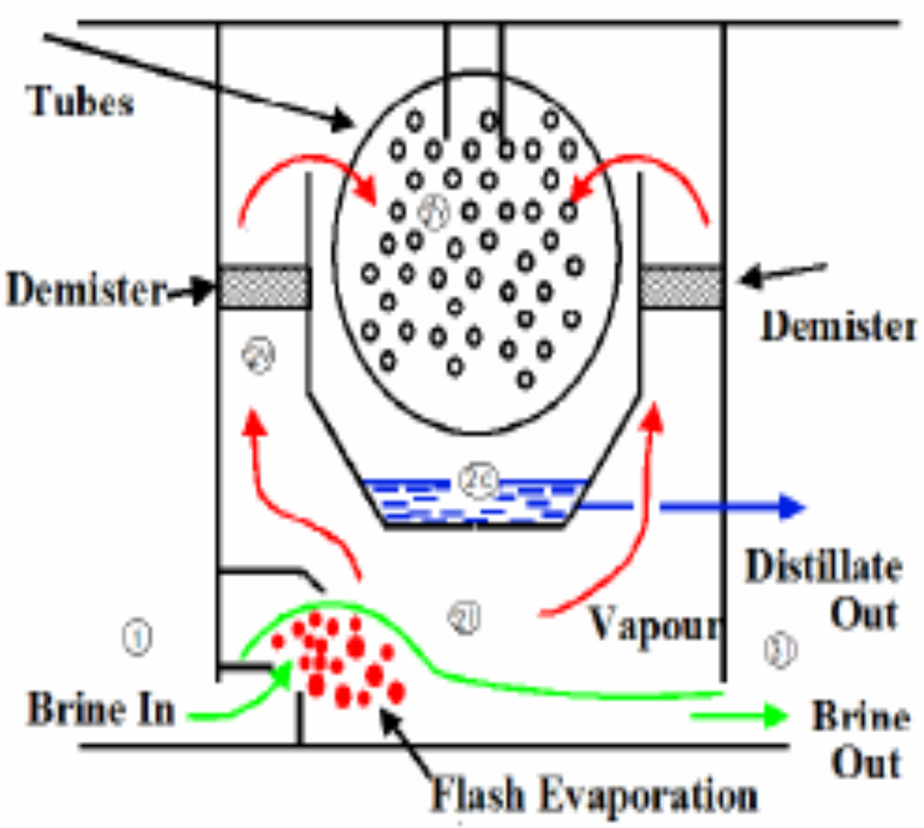
Stage Modeling Thermodynamic

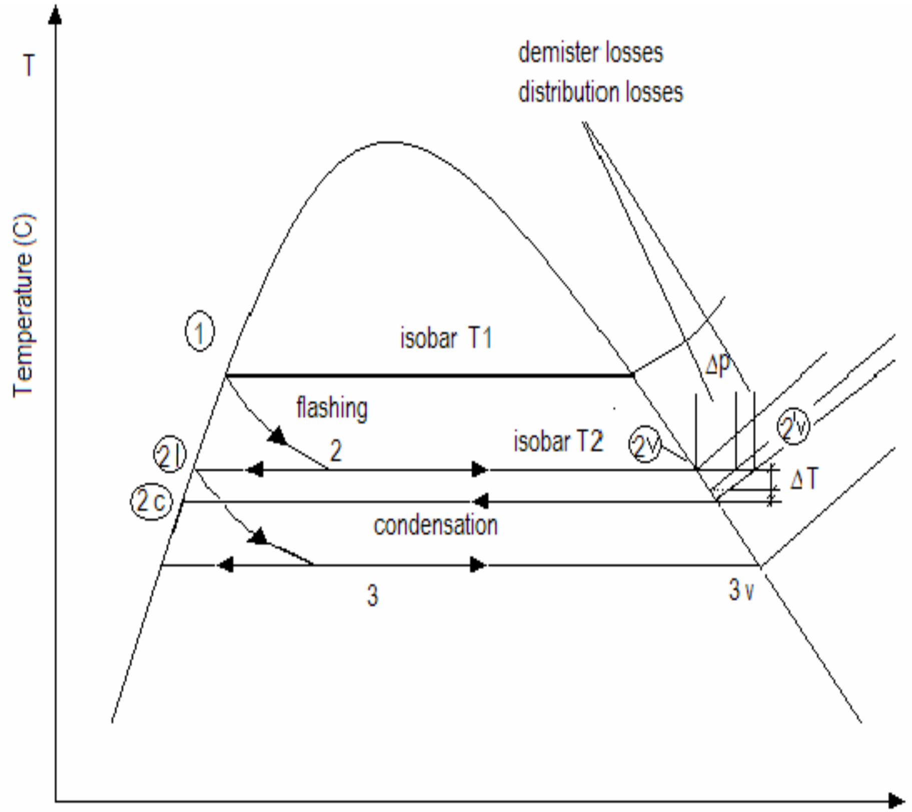
Stage modeling thermodynamic: understanding non equilibrium allowances



Stage Modeling Thermodynamic

Stage modeling thermodynamic: real case

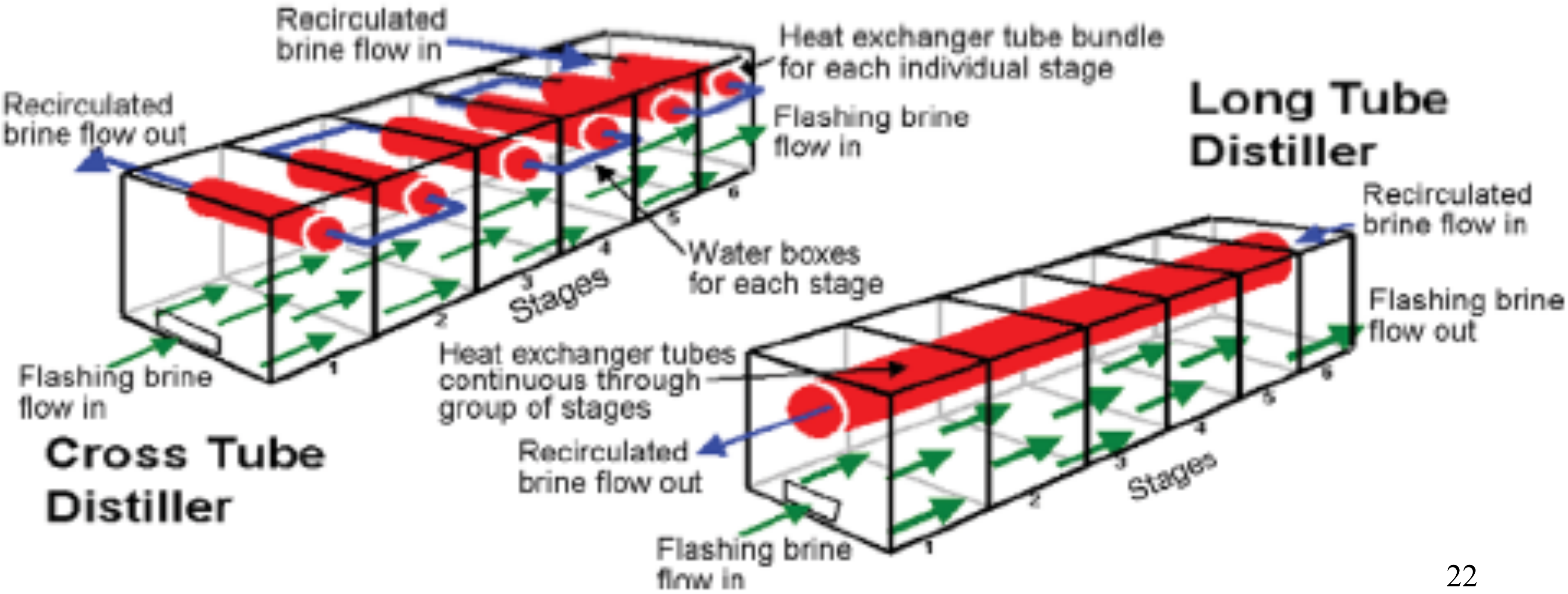




MULTISTAGE FLASH TECHNOLOGY (MSF)

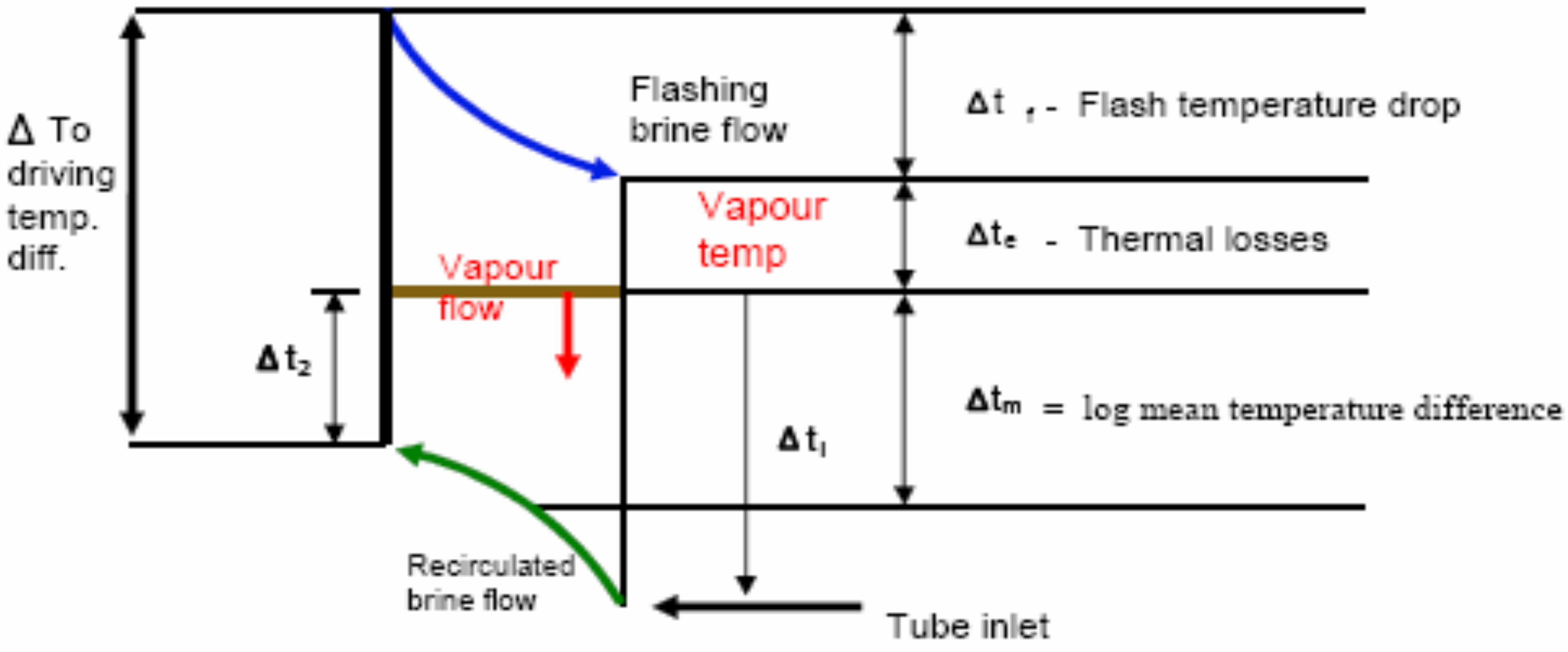
Multi stage flash

Cross tube and long tube MSF distillers



MSF Desalination Plant

Single stage temperature diagram



Δt_1 = Inlet temperature difference - Vapour to brine in

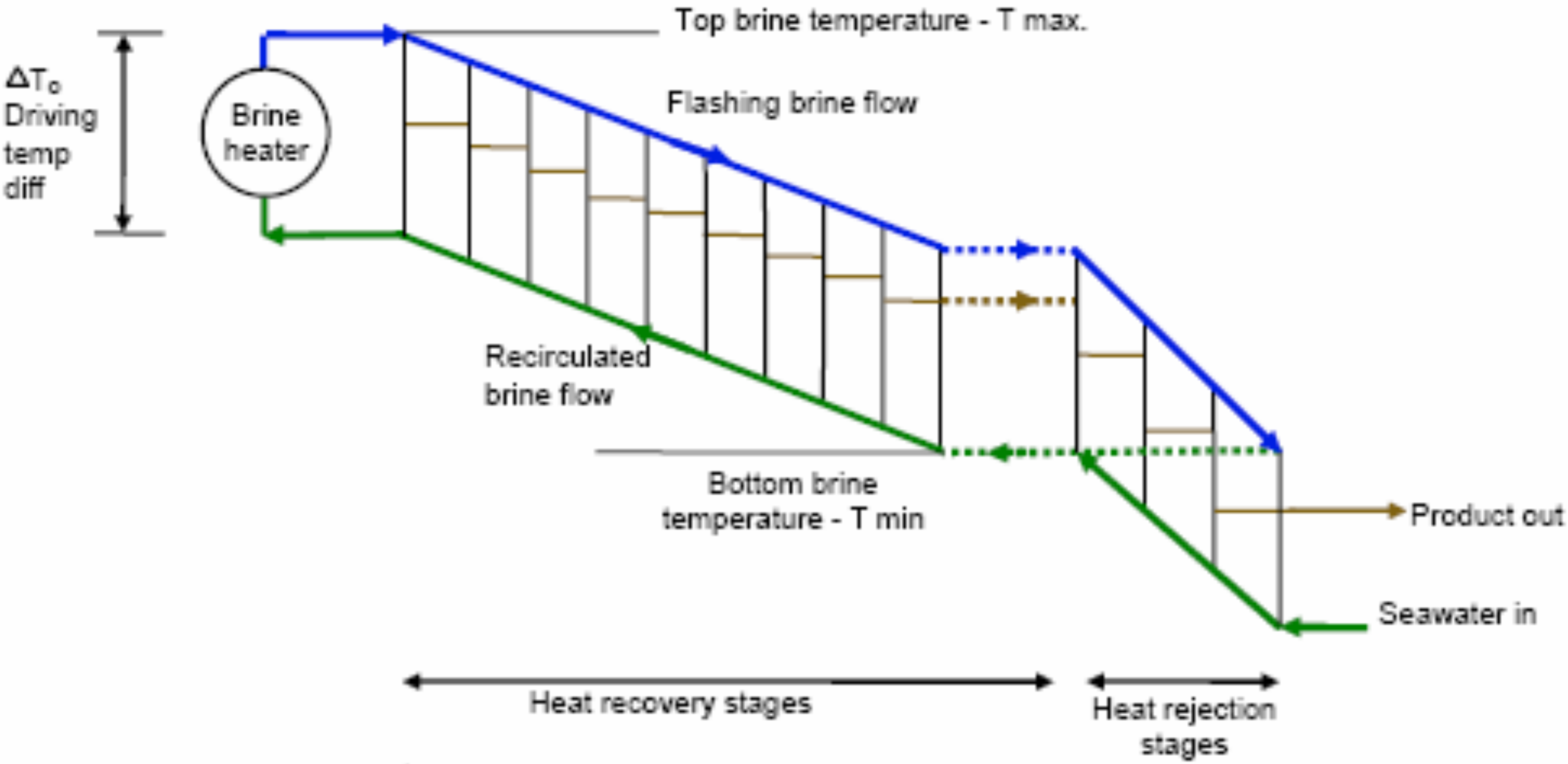
Δt_2 = Outlet temperature difference - Vapour to brine out

Δt_m = log mean temperature difference (LMTD)

$$\Delta t_m = \frac{\Delta t_1 \quad \Delta t_2}{\log\left(\Delta t_1 / \Delta t_2\right)}$$

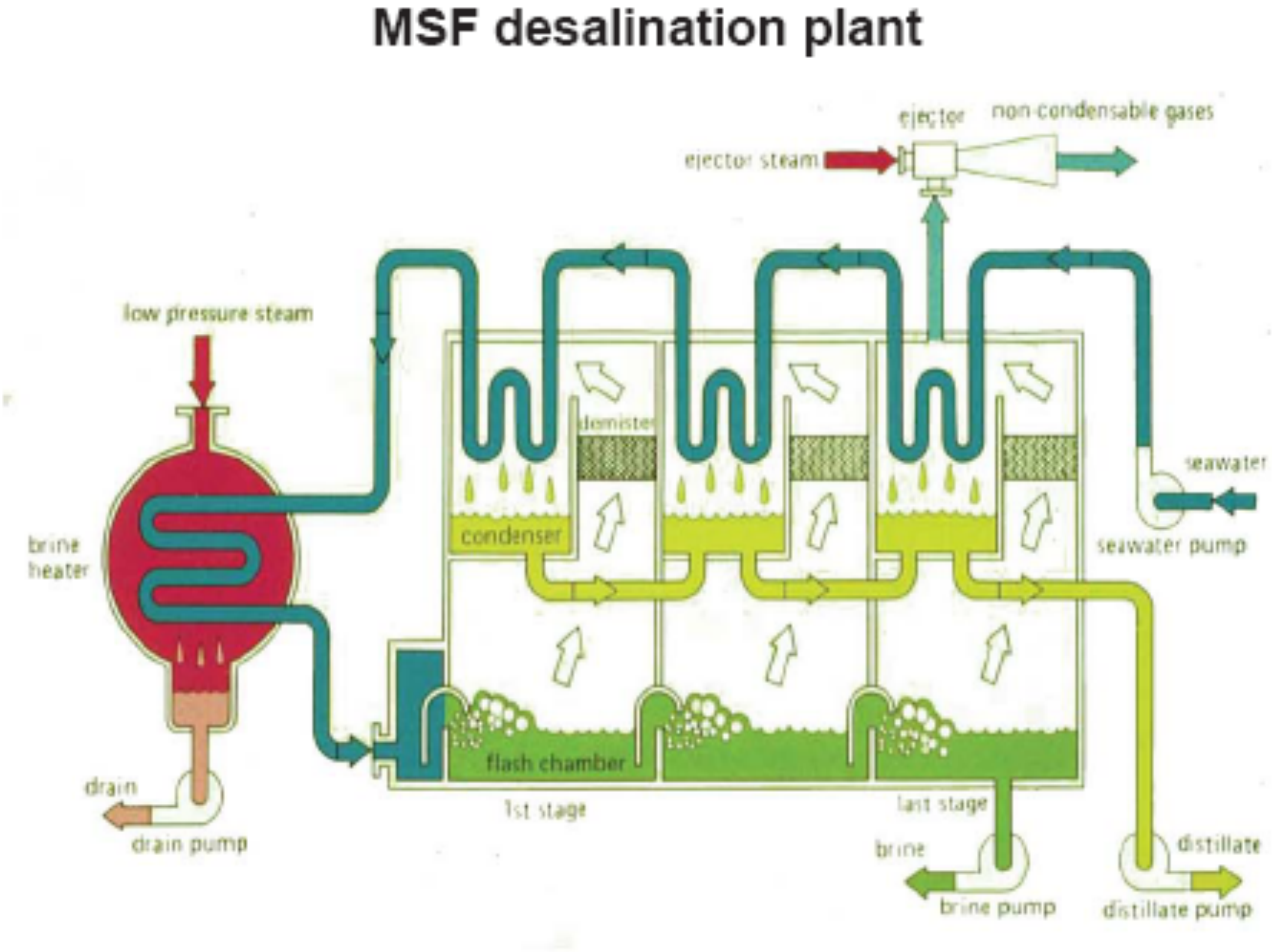
Stage Temperature Diagram

Stage temperature diagram
Complete plant (brine recirculation type)

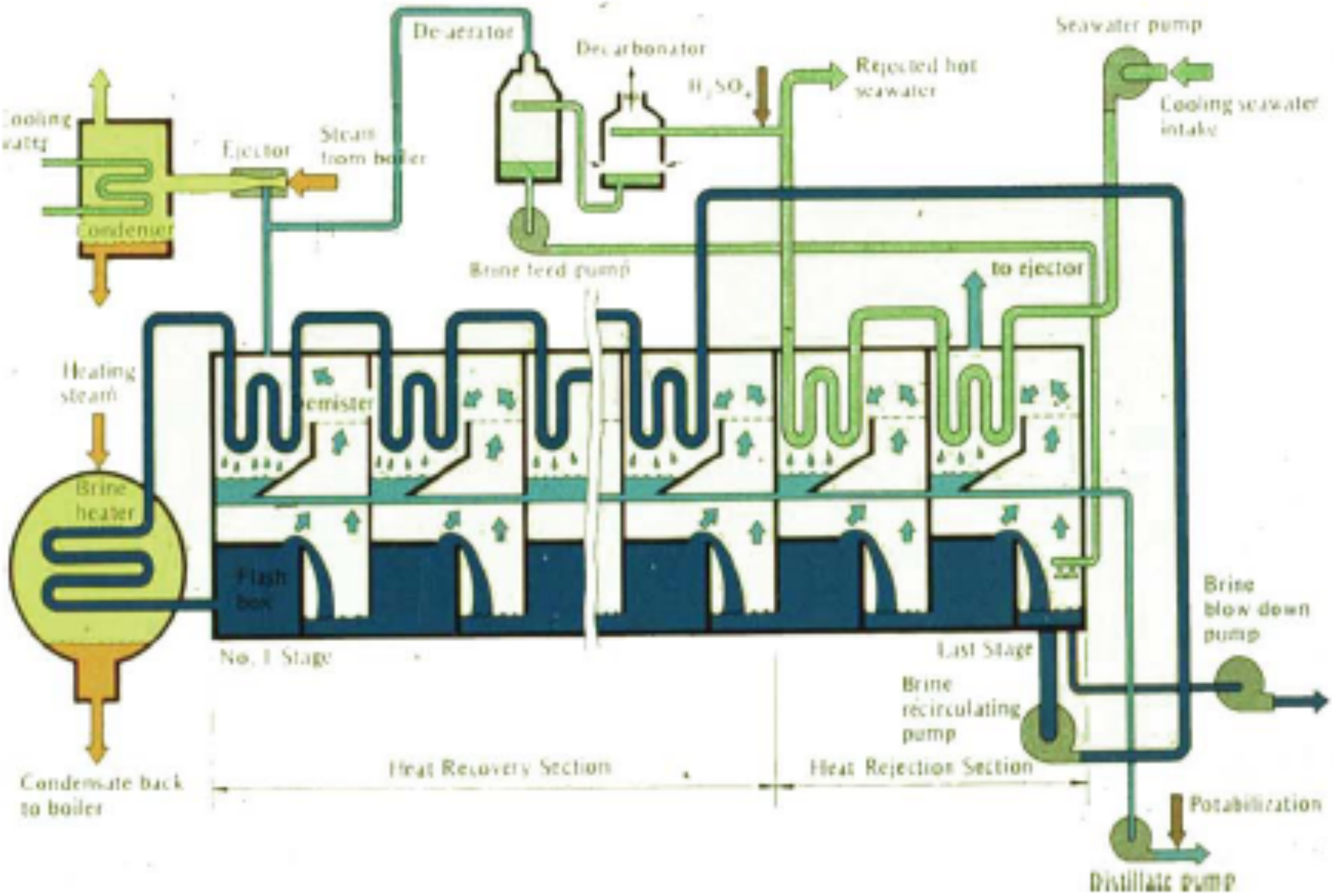


ΔTo = Driving temperature difference
(practically constant through heat recovery stages)

MSF Desalination Plant



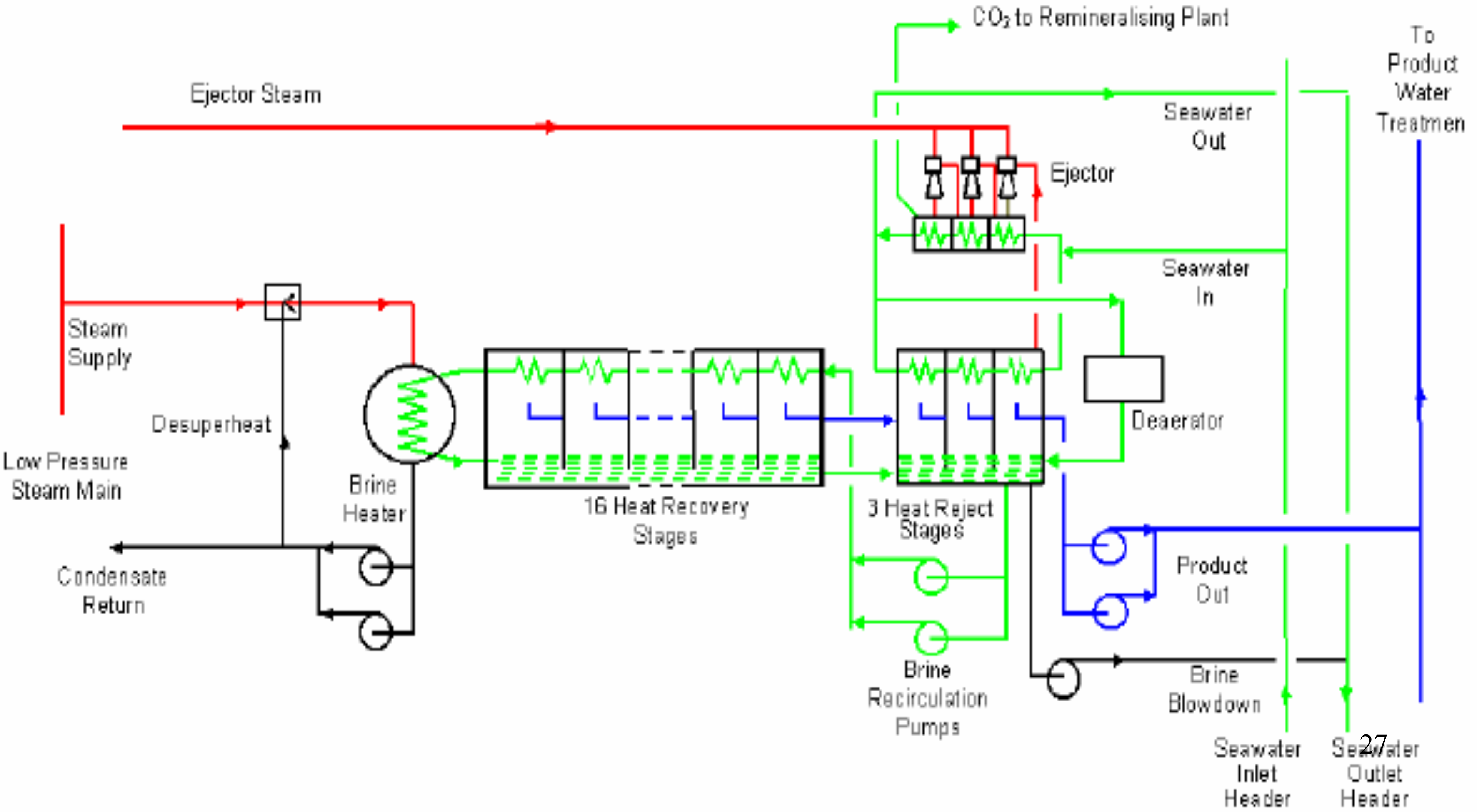
MULTISTAGE FLASH TECHNOLOGY (MSF)



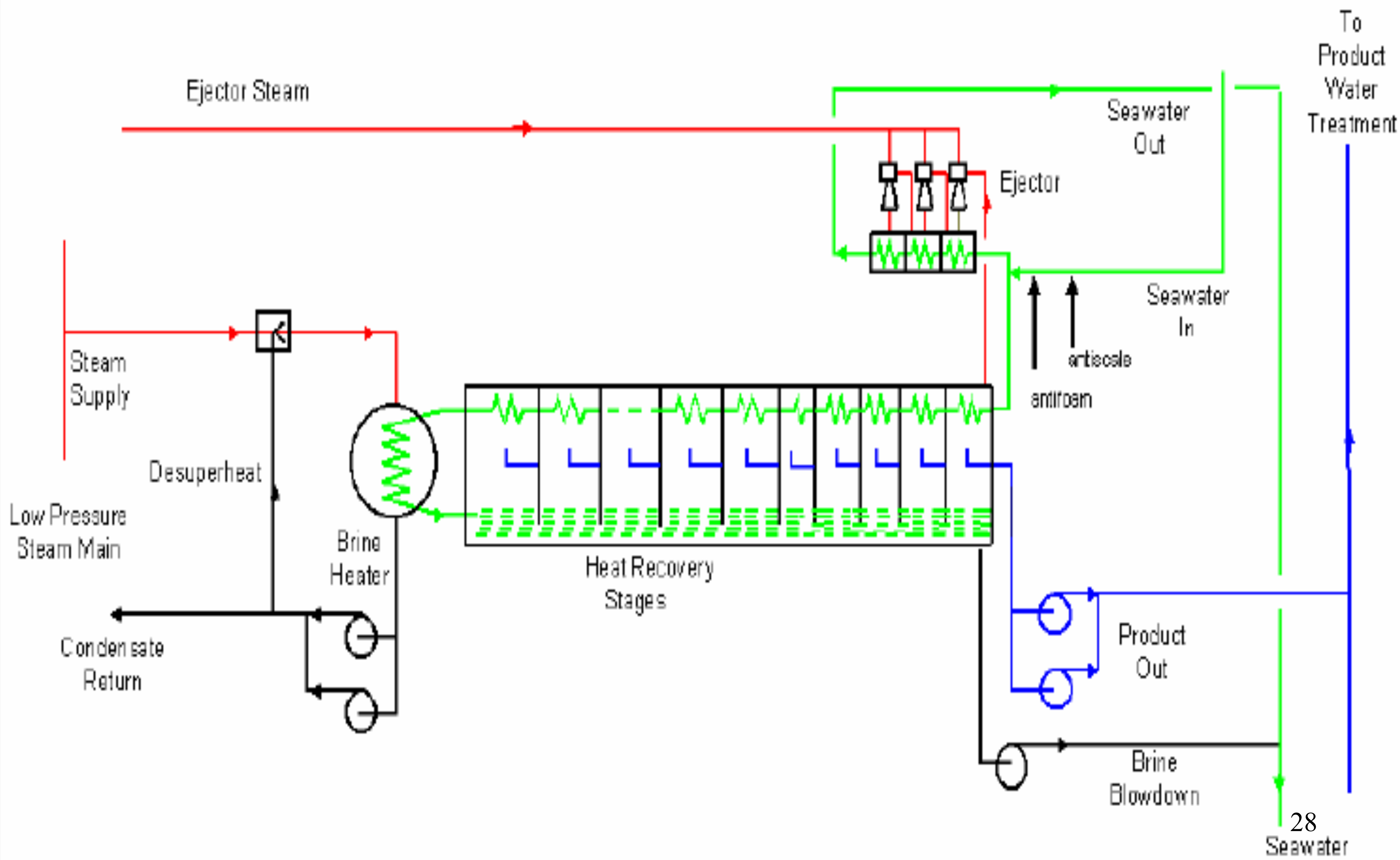
MSF what process?

Flow Sheet

Flow sheets: cross flow brine recirculation

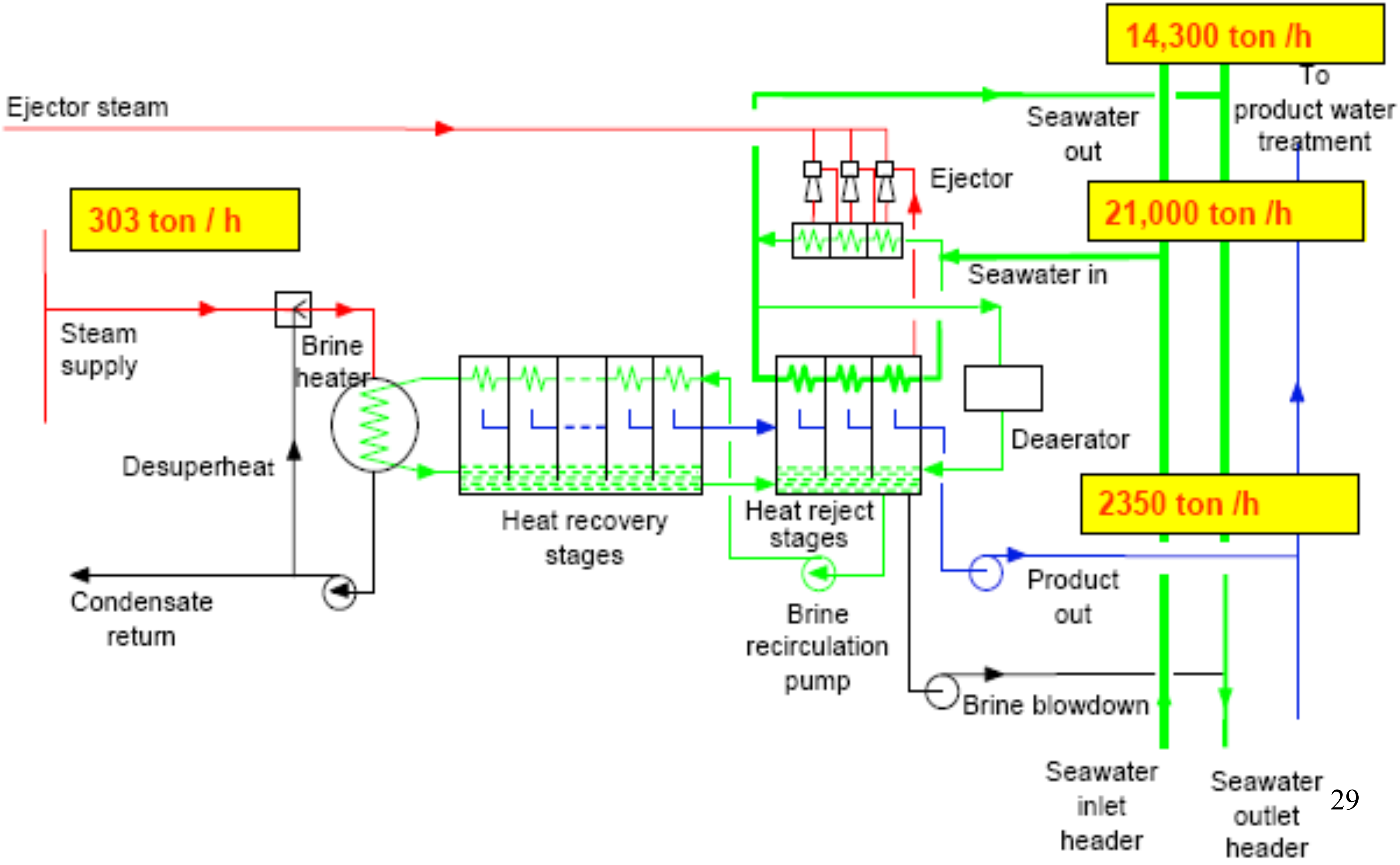


Flow Sheet: once through



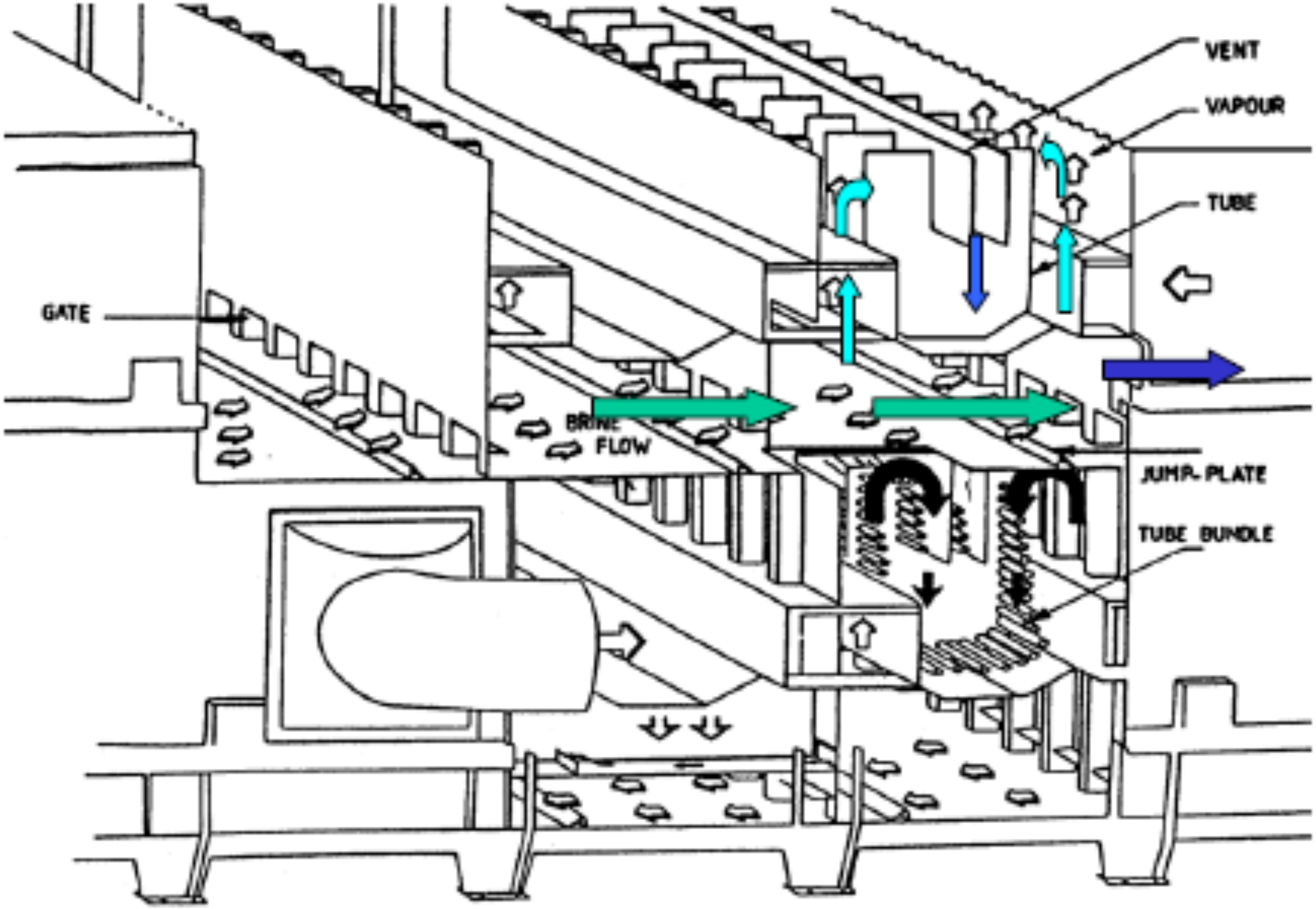
Stream Mass Balance

Main flow stream mass balance



MSF cross flow plant internal layout

MSF cross flow plant internal layout



MSF cross flow plant internal layout

How it really looks like - low side flash chamber

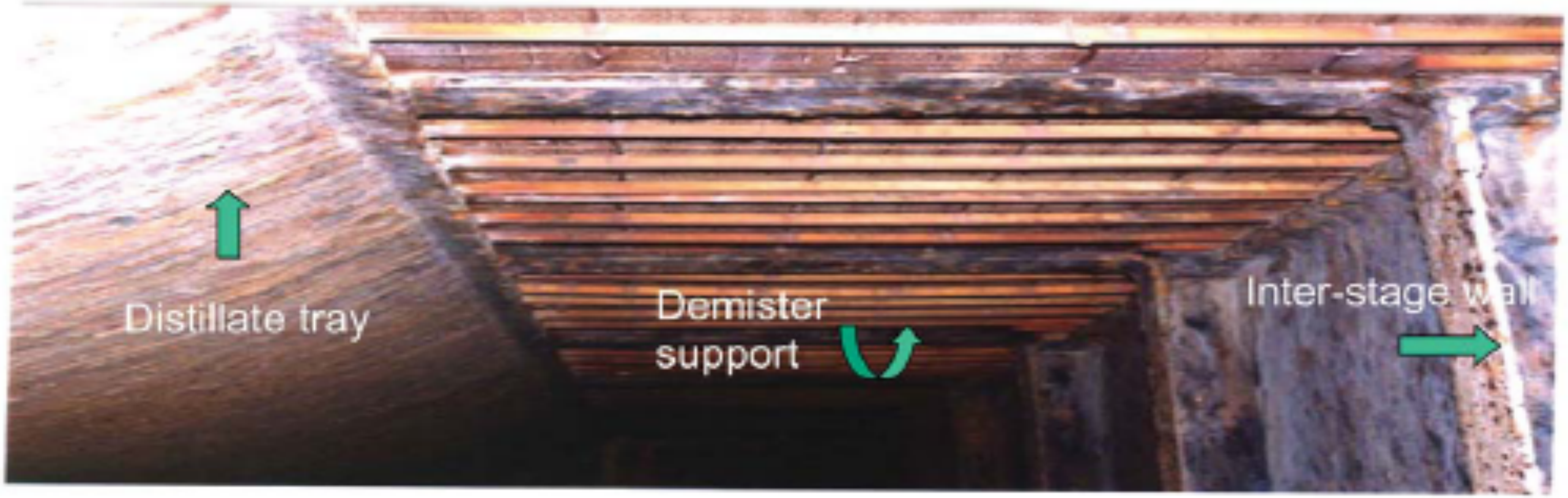
MSF cross flow plant internal layout

How it really looks like - upper side

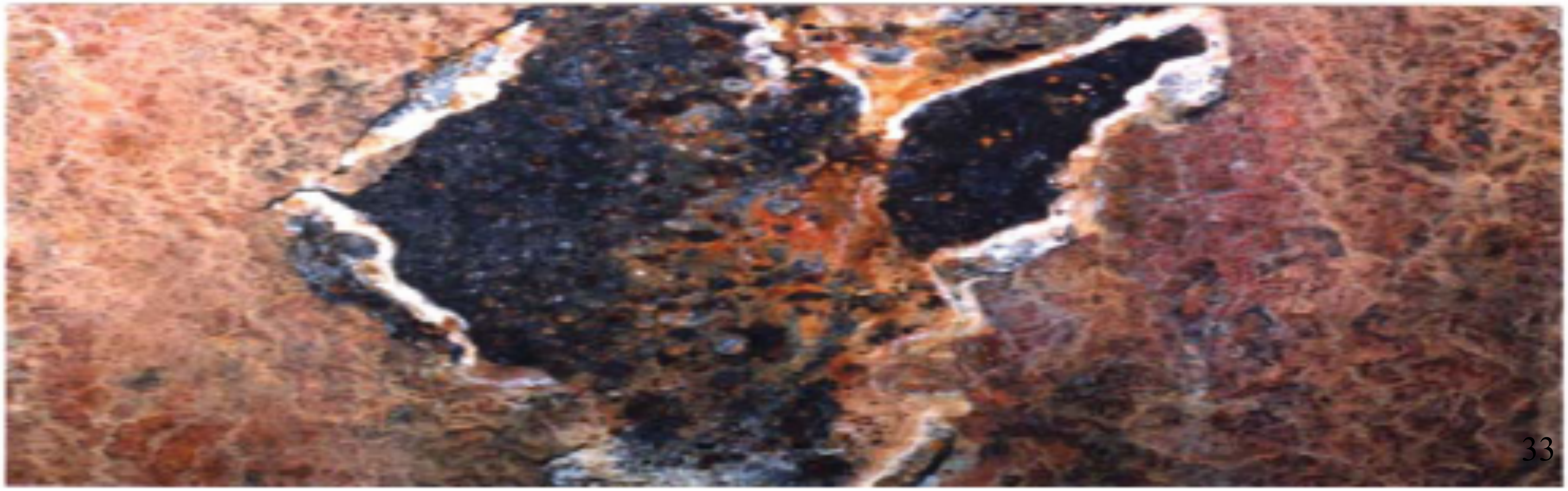
Tube bundle tube supports roof plates and incondensable extraction pipes

Details of tube bundle and tube support

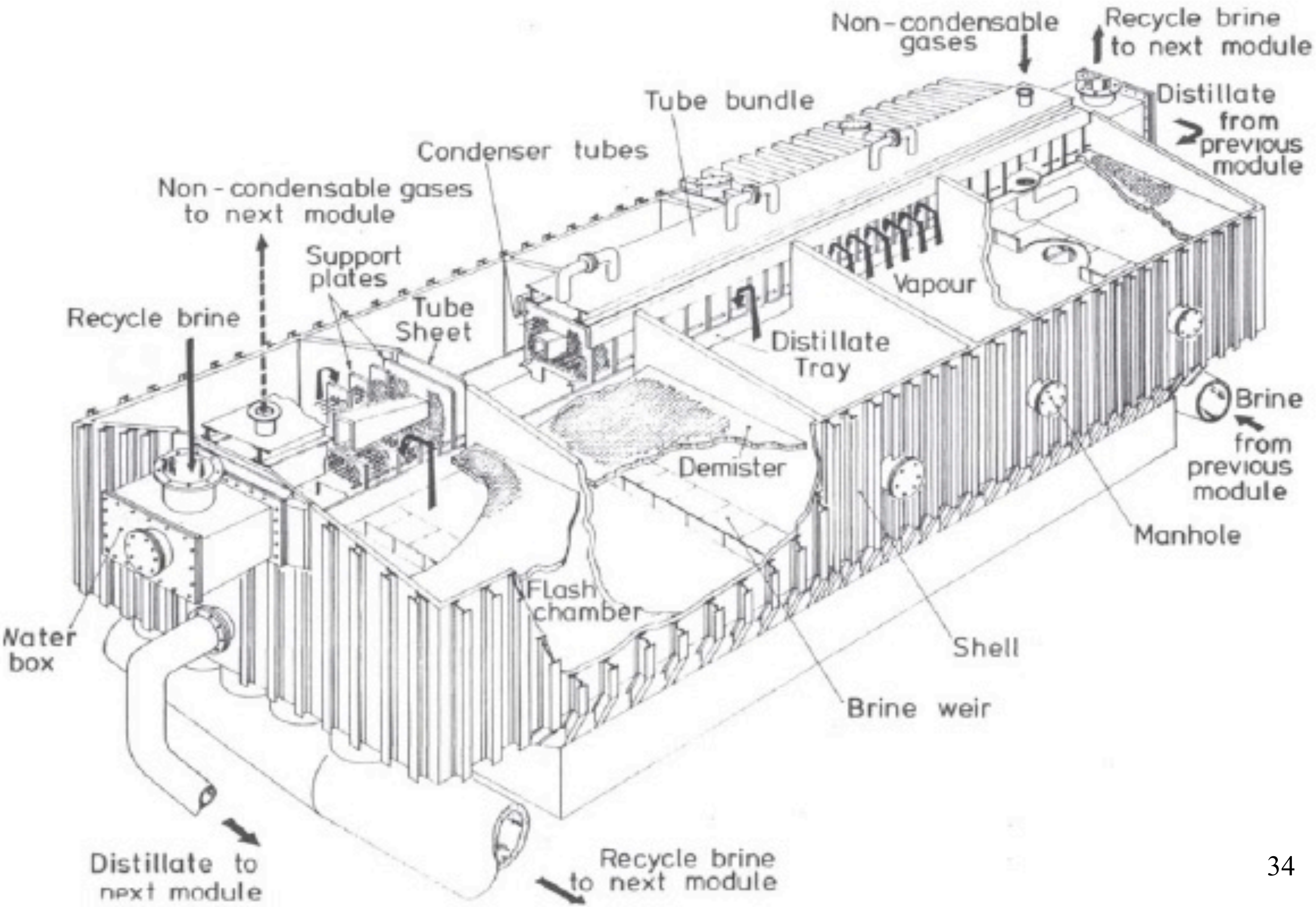
MSF cross flow plant internal layout



Distillate tray, demister supports and interstage walls



MSF long flow plant internal layout



MSF long flow plant internal layout

MSF long flow plant internal layout:
how it really looks like

Process constraints & Venting

- **Process constraints?**

Experience with all systems indicated need for scale control, because

- Hot brines easily reached saturation with inorganic species ($Mg(OH)_2$, $CaCO_3$, $Ca(HCO_3)_2CaSO_4$, etc.)
- Scale restricted flow paths, reduced heat transfer, caused outages

- **Venting**

In evaporative systems, non-condensable gases accumulate in vapor space, reducing the partial pressure of water. This in turn retards condensation rates and can even lead to a “blanketing” effect on heat exchange surfaces.

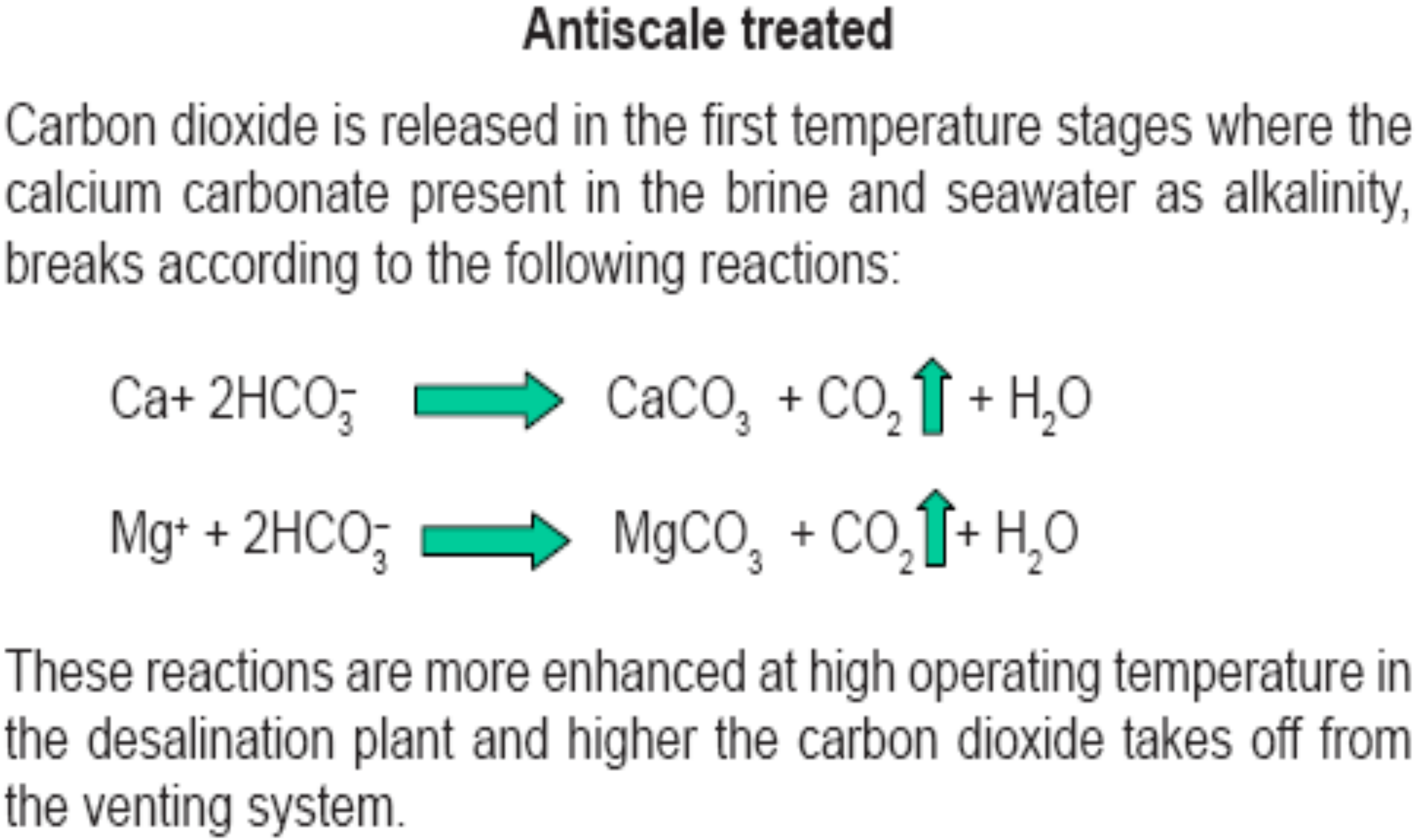
Needs for a incondensable gases extraction system

Dissolved gases

- **Dissolved gases**

- Chemically active gases include:
- Oxygen
- Carbon dioxide
- Chlorine
- Hydrogen sulfide

Antiscale Treated



Antiscale treated

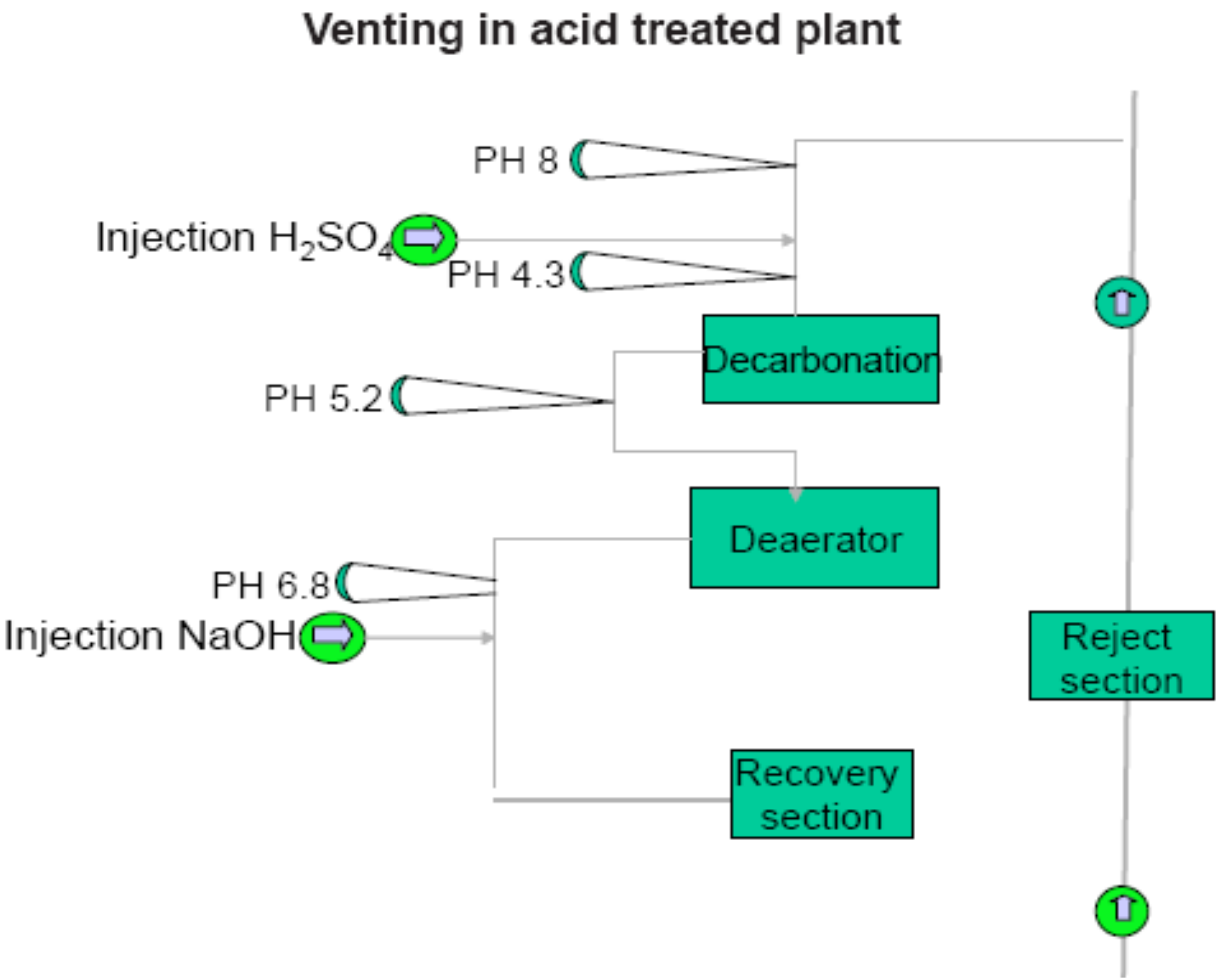
Carbon dioxide is released in the first temperature stages where the calcium carbonate present in the brine and seawater as alkalinity, breaks according to the following reactions:

$$Ca^{+} + 2HCO_3^{-} \rightarrow CaCO_3 + CO_2 \uparrow + H_2O$$

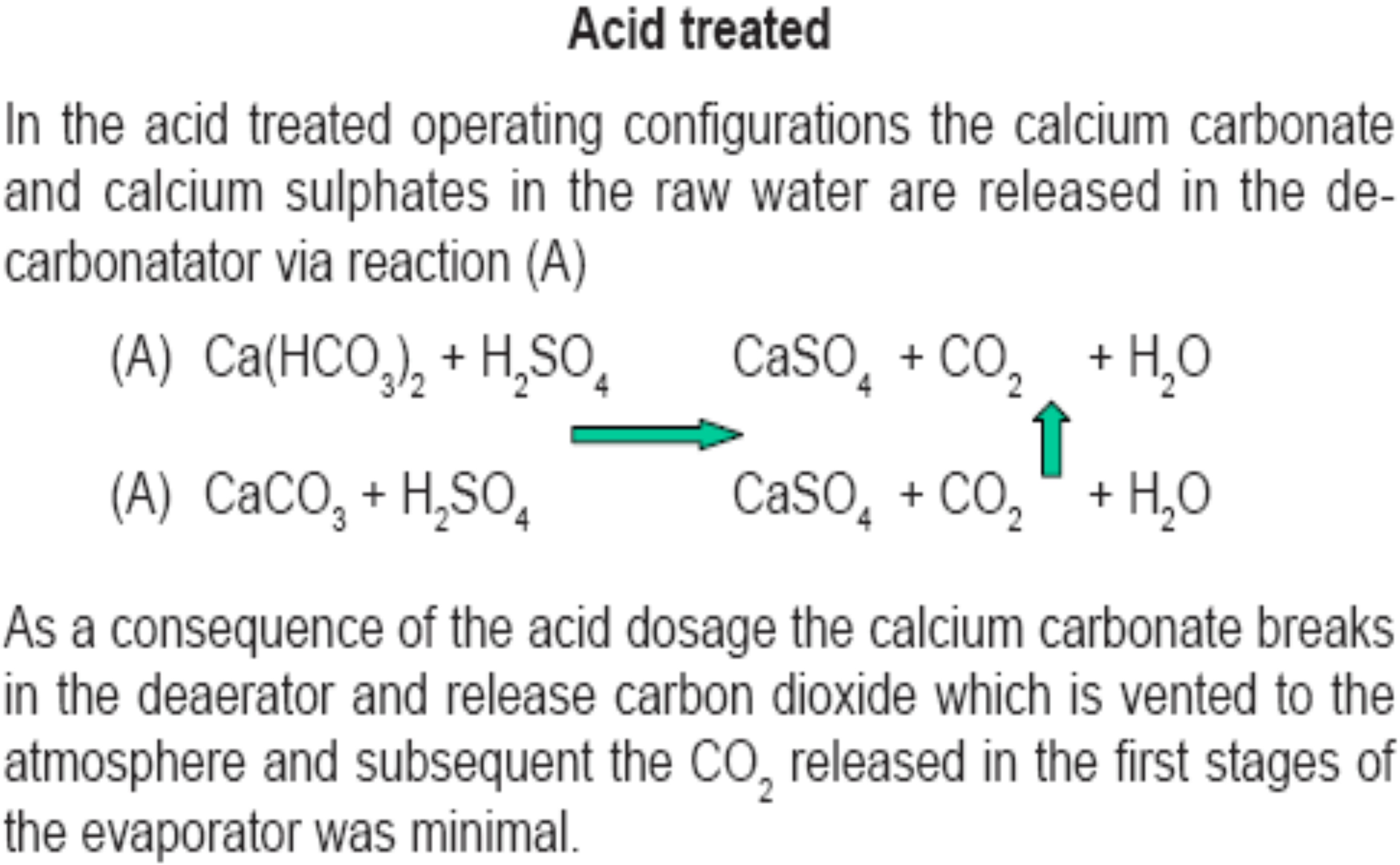
$$Mg^{+} + 2HCO_3^{-} \rightarrow MgCO_3 + CO_2 \uparrow + H_2O$$

These reactions are more enhanced at high operating temperature in the desalination plant and higher the carbon dioxide takes off from the venting system.

Venting in acid treated plant



Acid Treated



Acid treated

In the acid treated operating configurations the calcium carbonate and calcium sulphates in the raw water are released in the de-carbonatator via reaction (A)

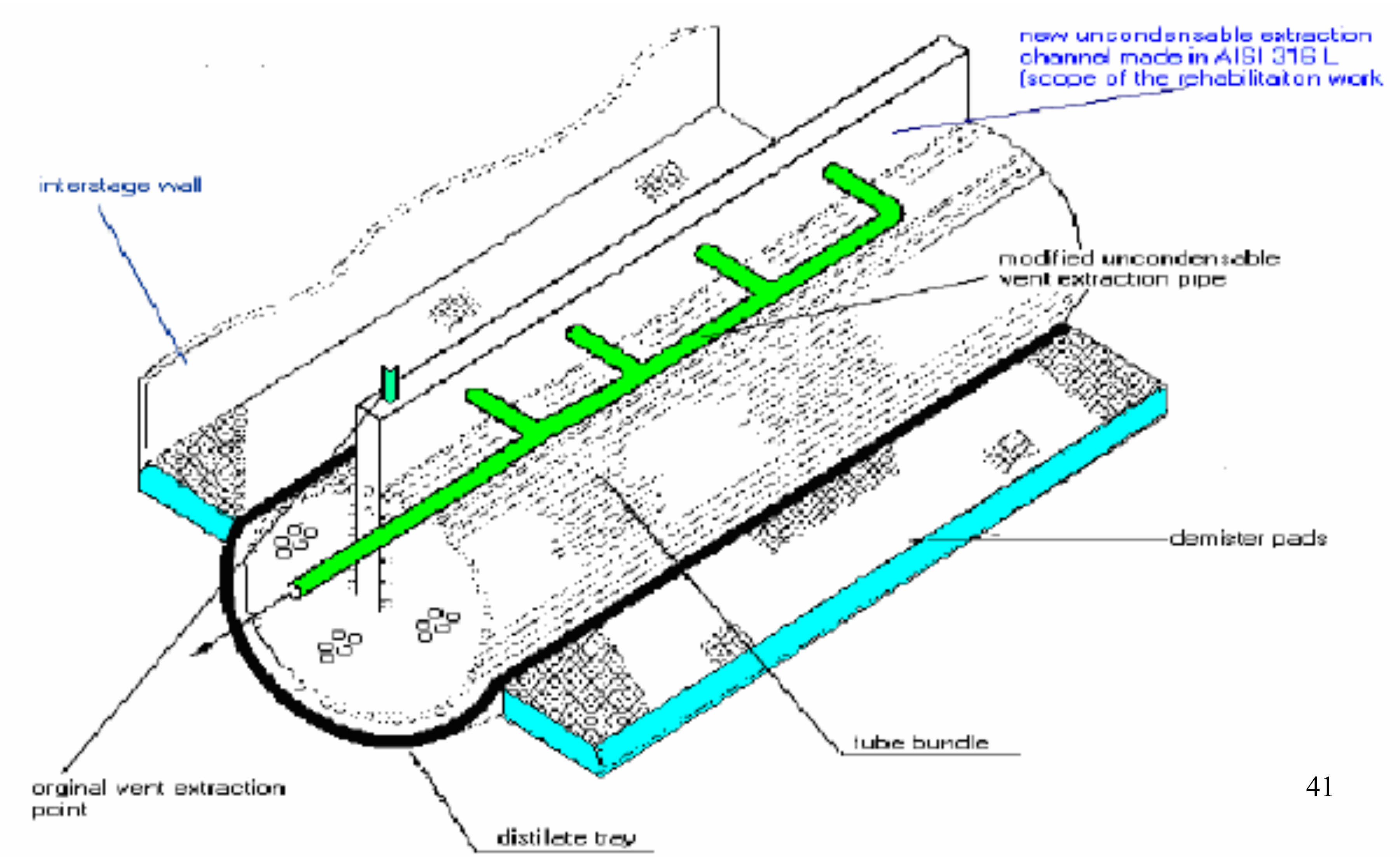
(A) $Ca(HCO_3)_2 + H_2SO_4 \rightarrow CaSO_4 + CO_2 \uparrow + H_2O$

(A) $CaCO_3 + H_2SO_4 \rightarrow CaSO_4 + CO_2 \uparrow + H_2O$

As a consequence of the acid dosage the calcium carbonate breaks in the deaerator and release carbon dioxide which is vented to the atmosphere and subsequent the CO_2 released in the first stages of the evaporator was minimal.

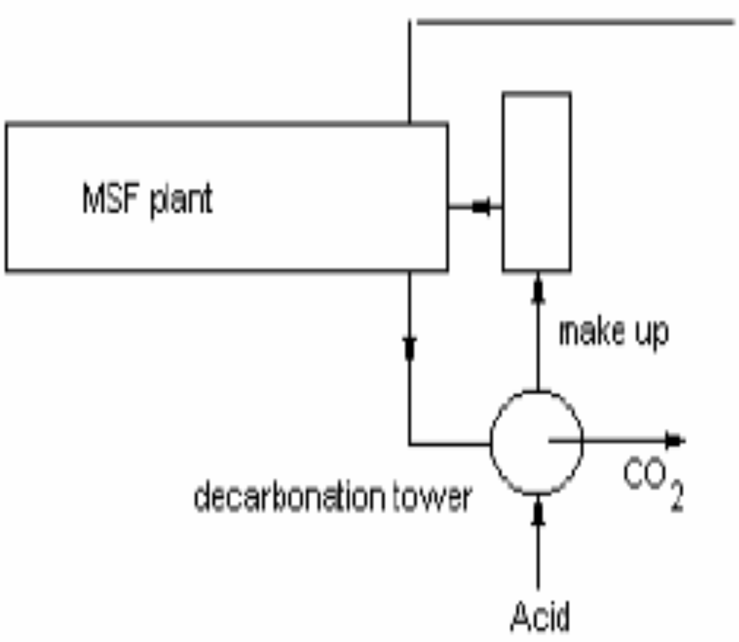
Venting System

Venting system

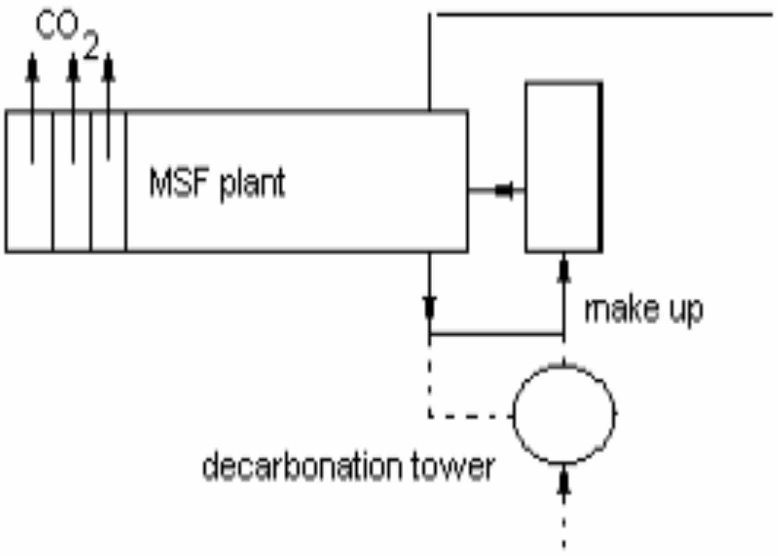


Incondensable gases in acid versus antiscale treatment system

Incondensable gases in acid versus antiscale treatment system



Acid

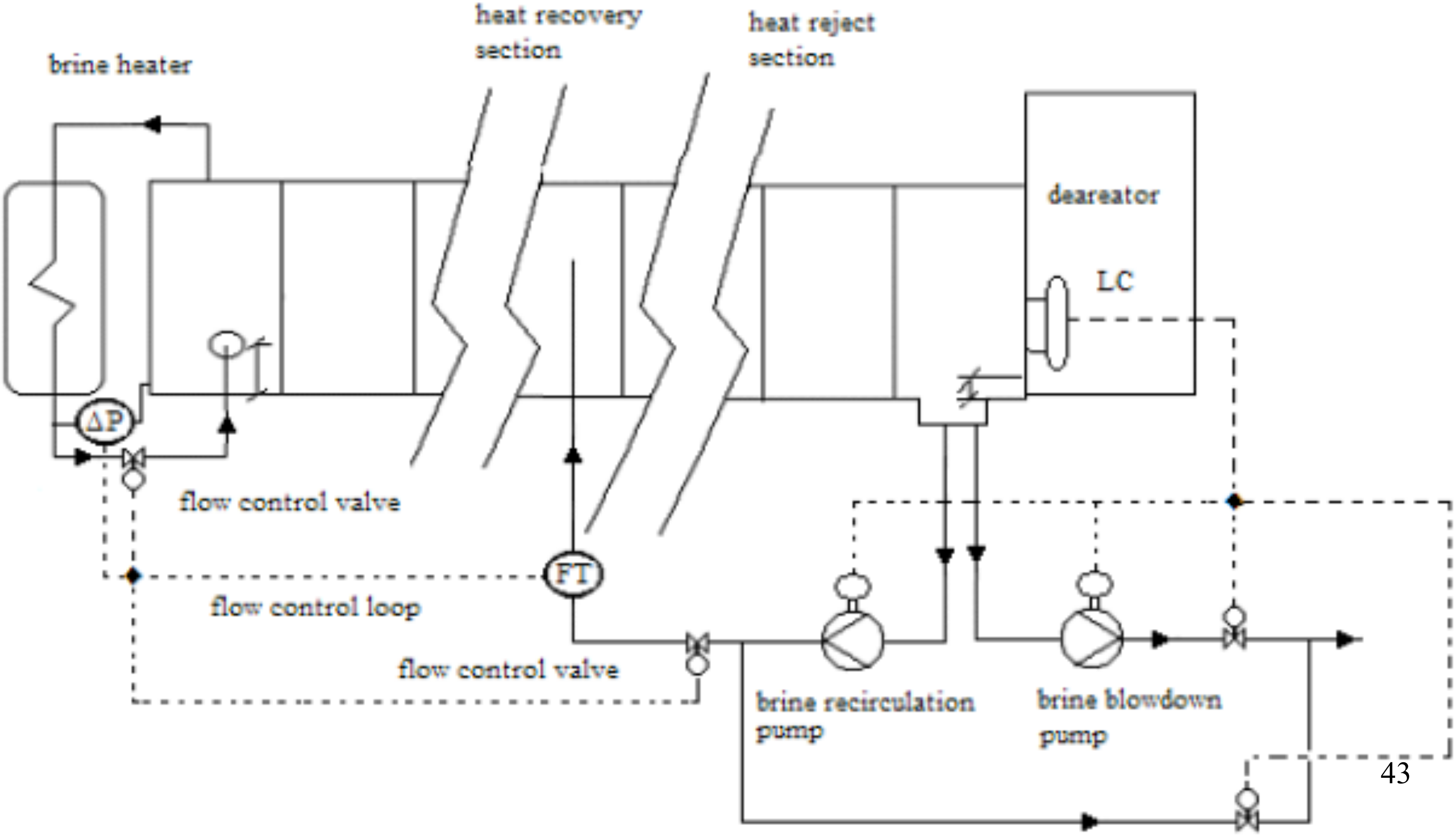


Anti-scale

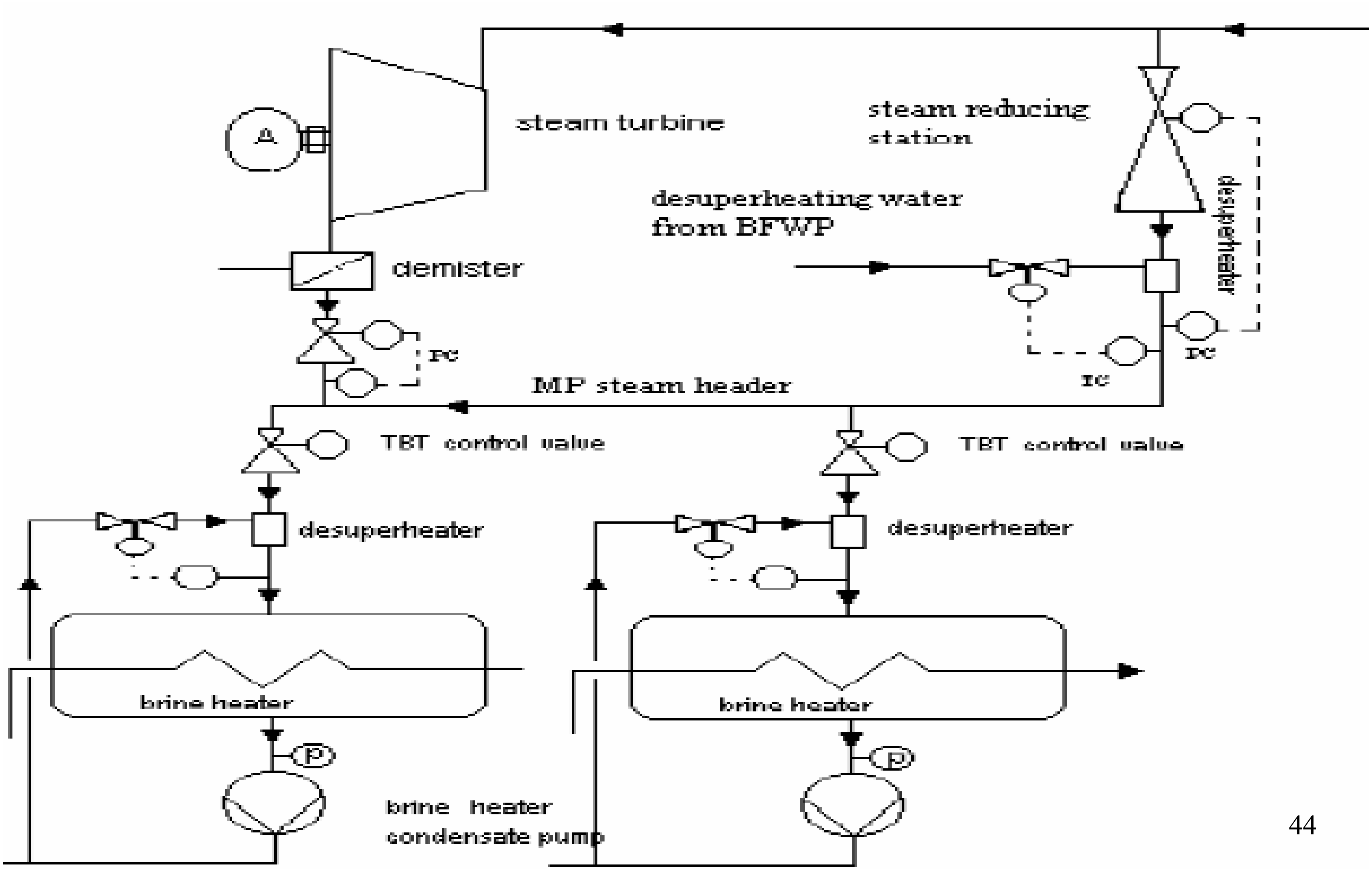
Other Control Loops

Other control loops

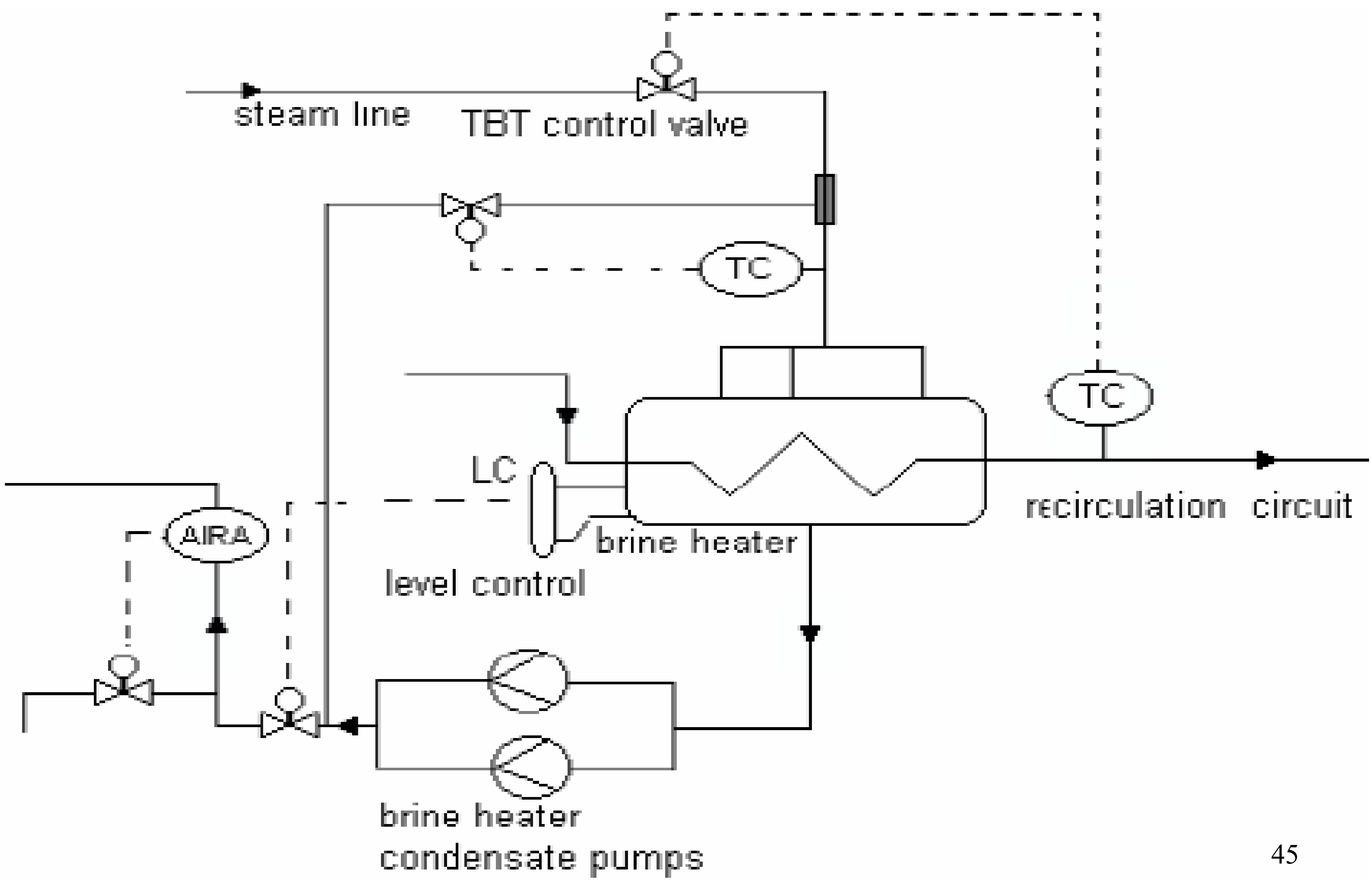
Brine recirculation-blowdown



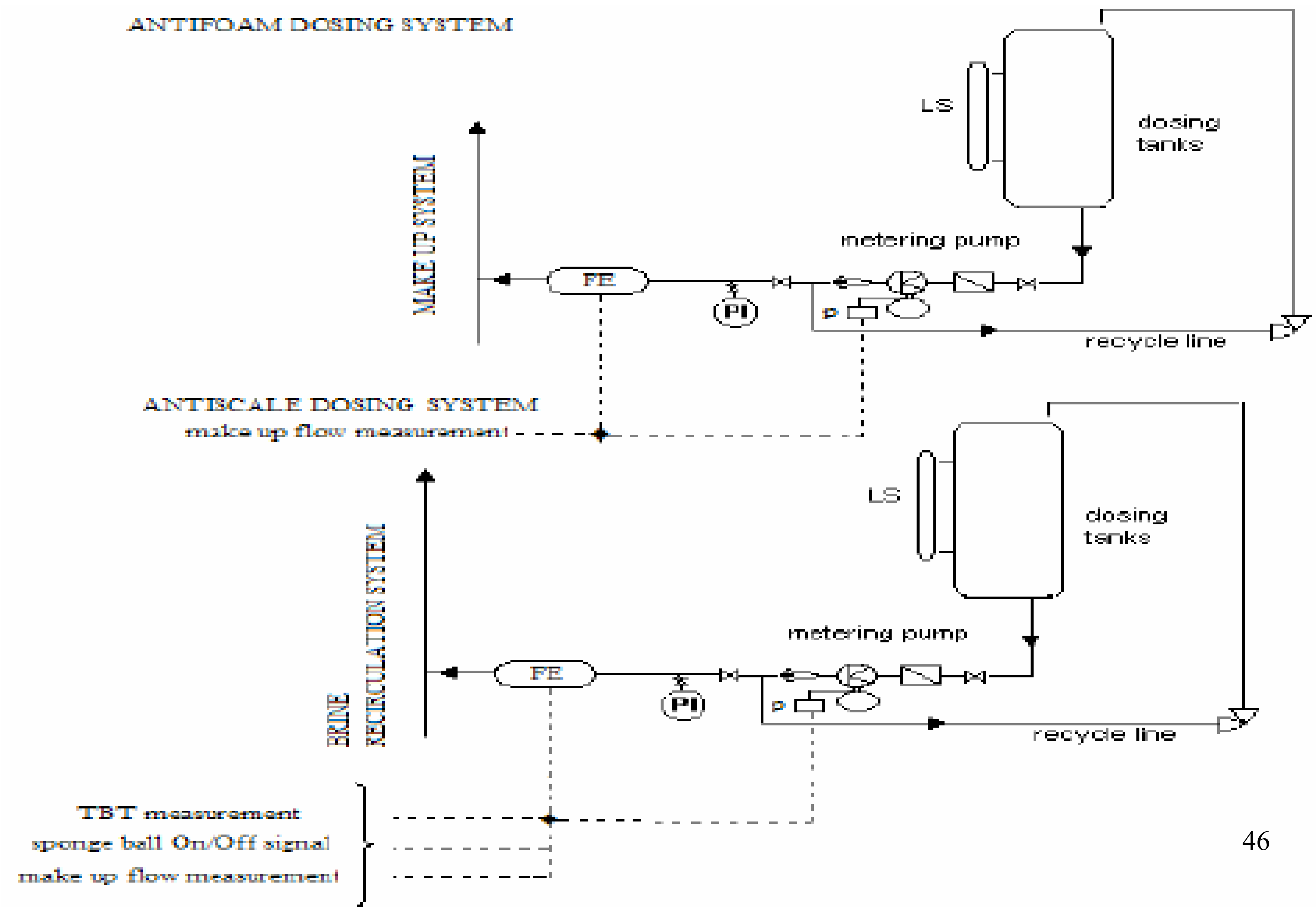
Steam feed arrangement to the plant



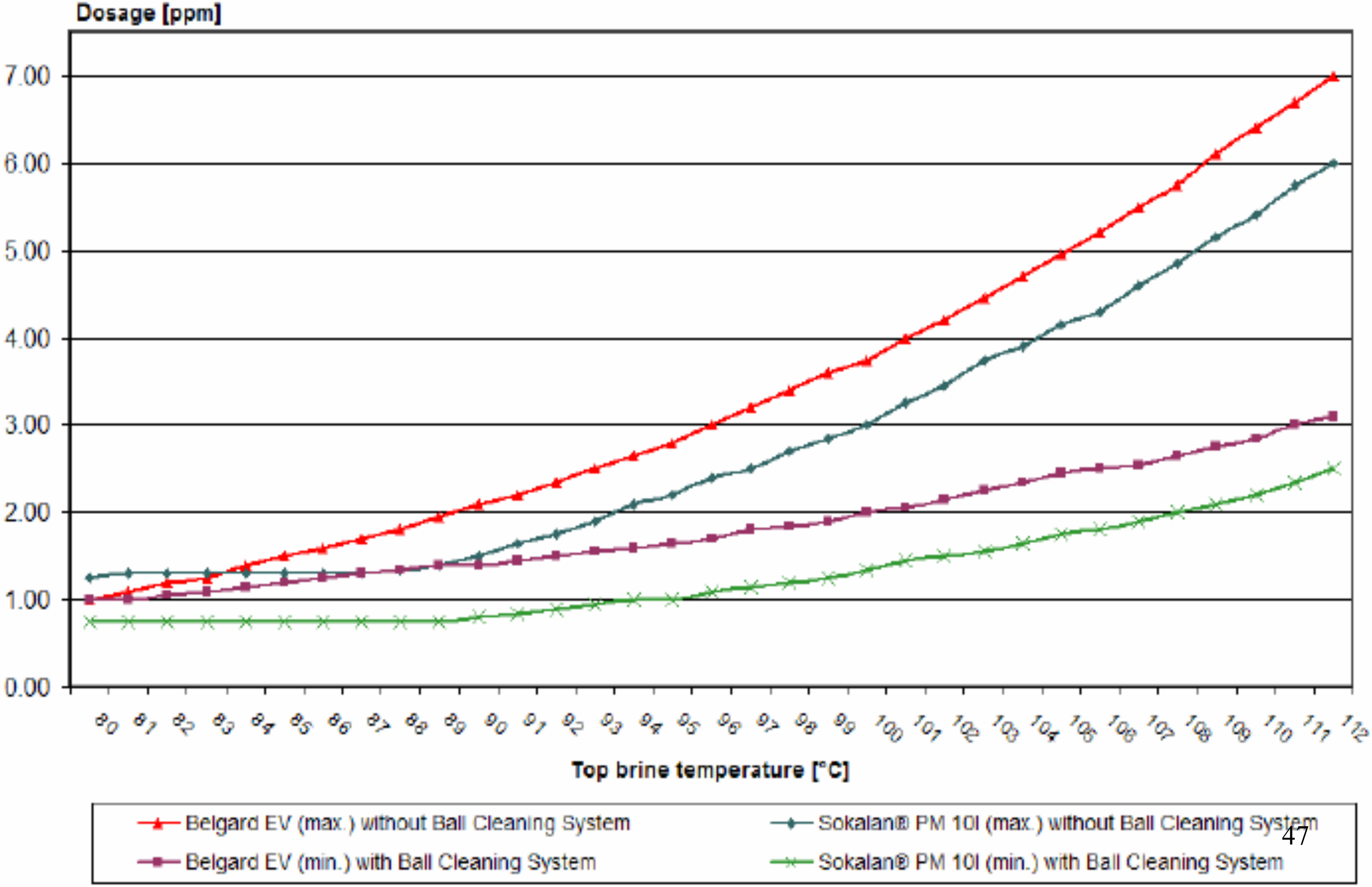
The most important control loop: TBT control loop



Variable controlled: steam to the brine heater resulting in variation of TBT



Dosage of Antiscalant in function of the top brine temperature



MSF plant: deaerator

MSF plant: deaerator

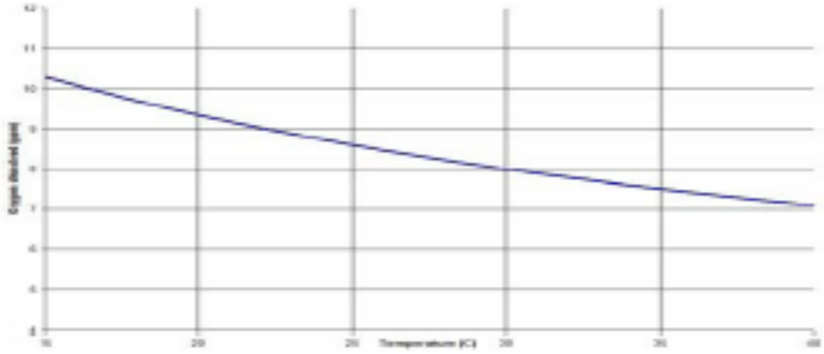
Solubility: a little bit of theory

In a ideal solution the vapor pressure of a gas is directly proportional to the molar fraction of that component in the solution (law of Raoult).

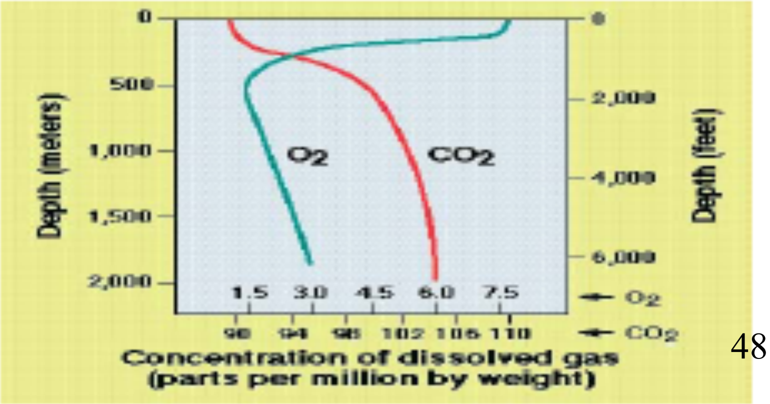
$$X_o(t) = \frac{0.21 \cdot PMO_2 \cdot p}{KH_o(t) \cdot PMH_2O}$$

where KH_o is the Henry constant variable with temperature.

$$KH_o(t) = \left(2.439 + \frac{9.086}{10^2} \cdot t - \frac{4.48}{10^4} \cdot t^2\right) \cdot 10^4 = (\text{bar})$$

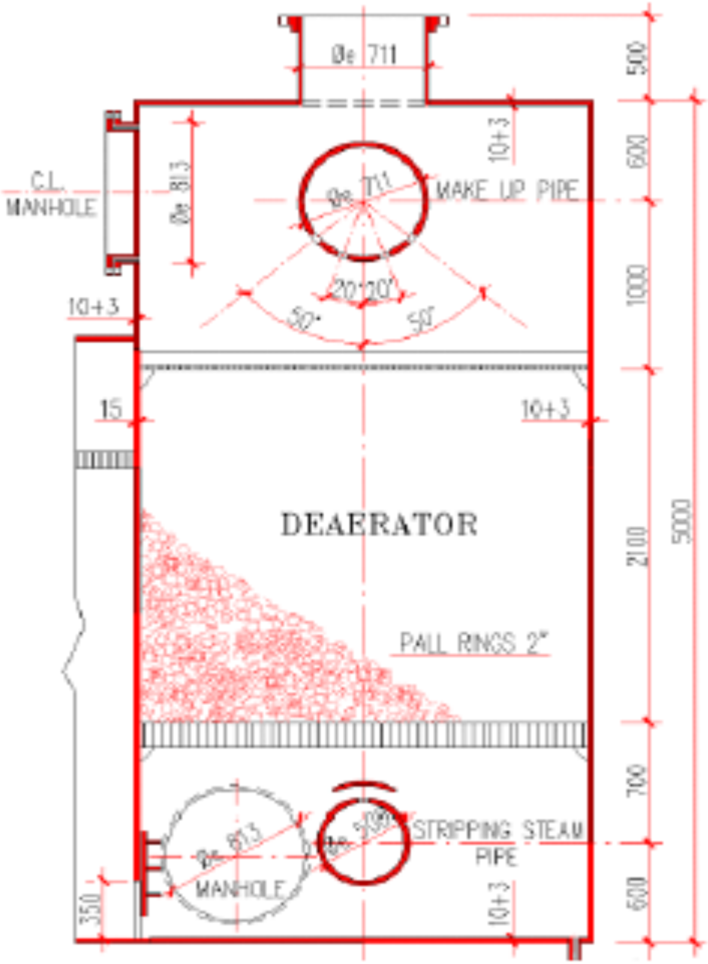


Solubility: oxygen

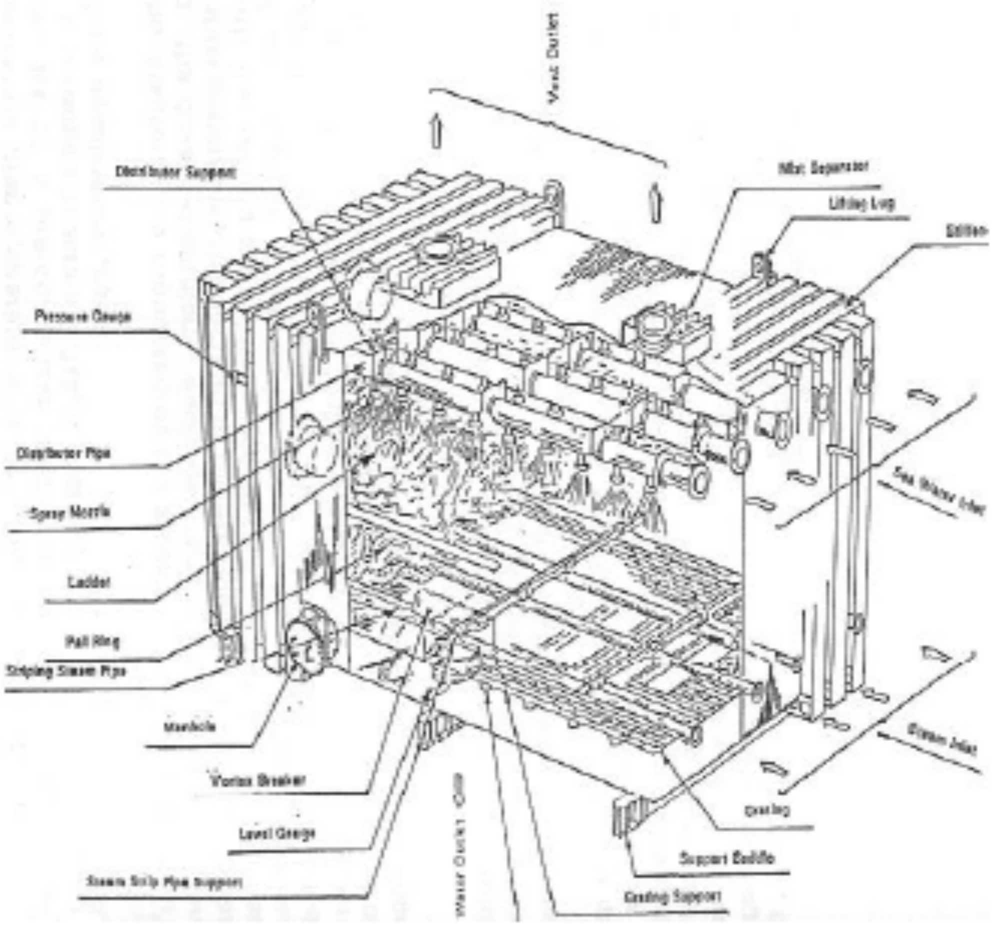


MSF plant: deaerator

MSF plant: deaerator



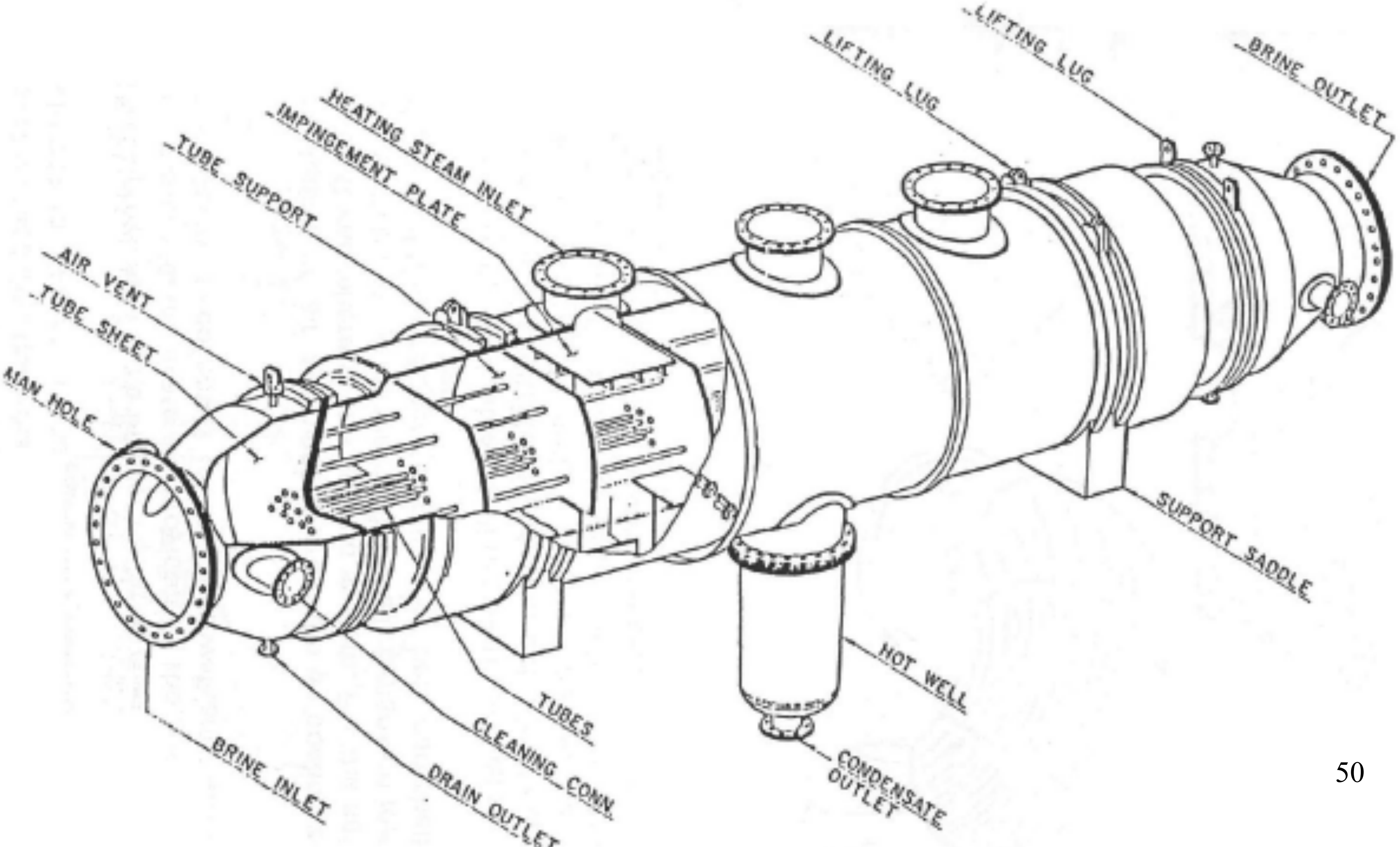
New design



Old design

MSF plant: brine heater

MSF plant: brine heater



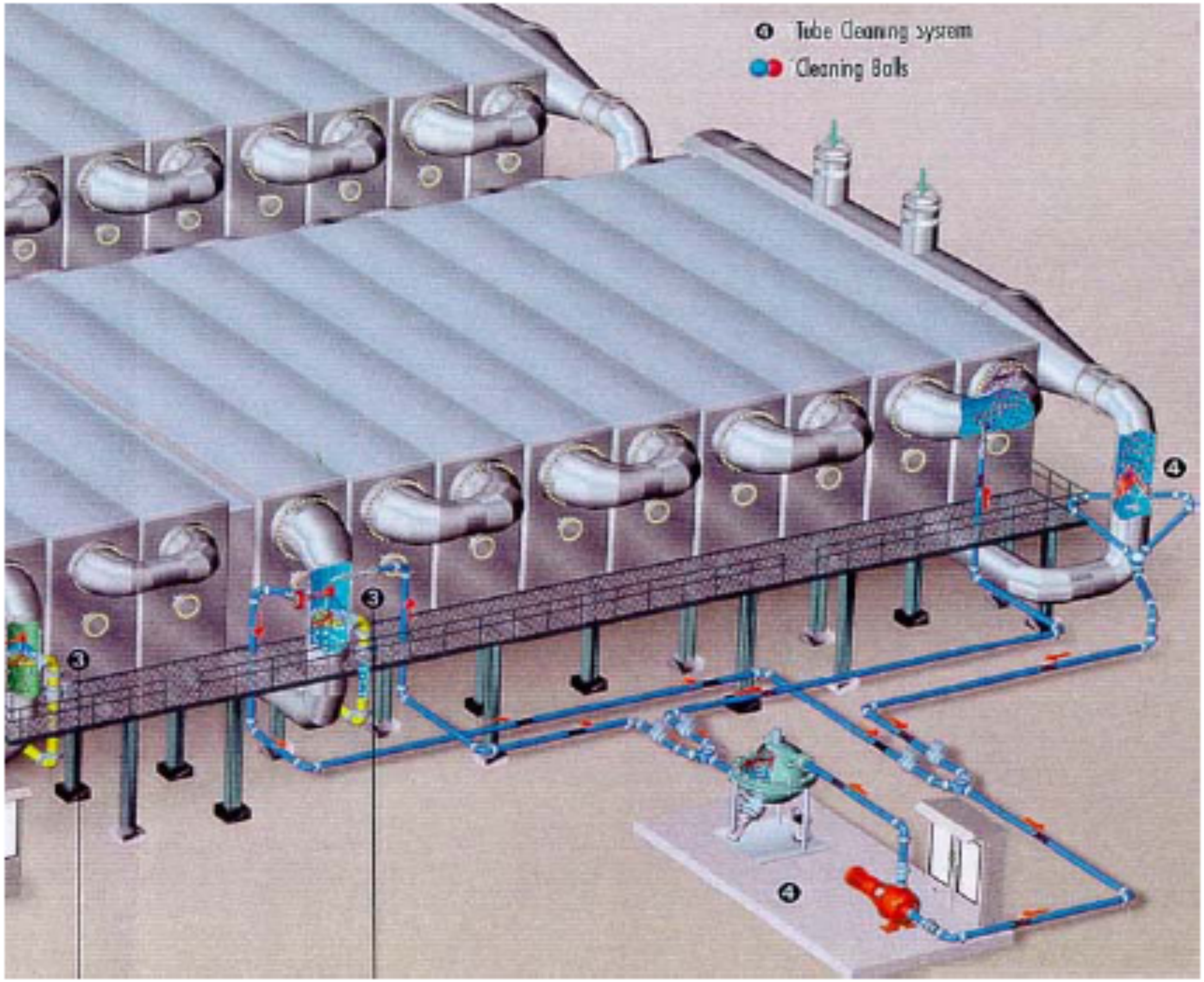
MSF plant: brine heater

- Shell and tube heat exchanger
- Single or double pass
- Tubes material: generally copper-nickel (90-10 or 66-30) sometime titanium
- Tubeplate material: generally copper-nickel (90-10) rarely titanium sometime aluminium-nickel-bronze

MSF plant: other essential components

MSF plant: other essential components

Sponge ball cleaning system



Some pictures

Some pictures

MODELLING MSF PLANTS

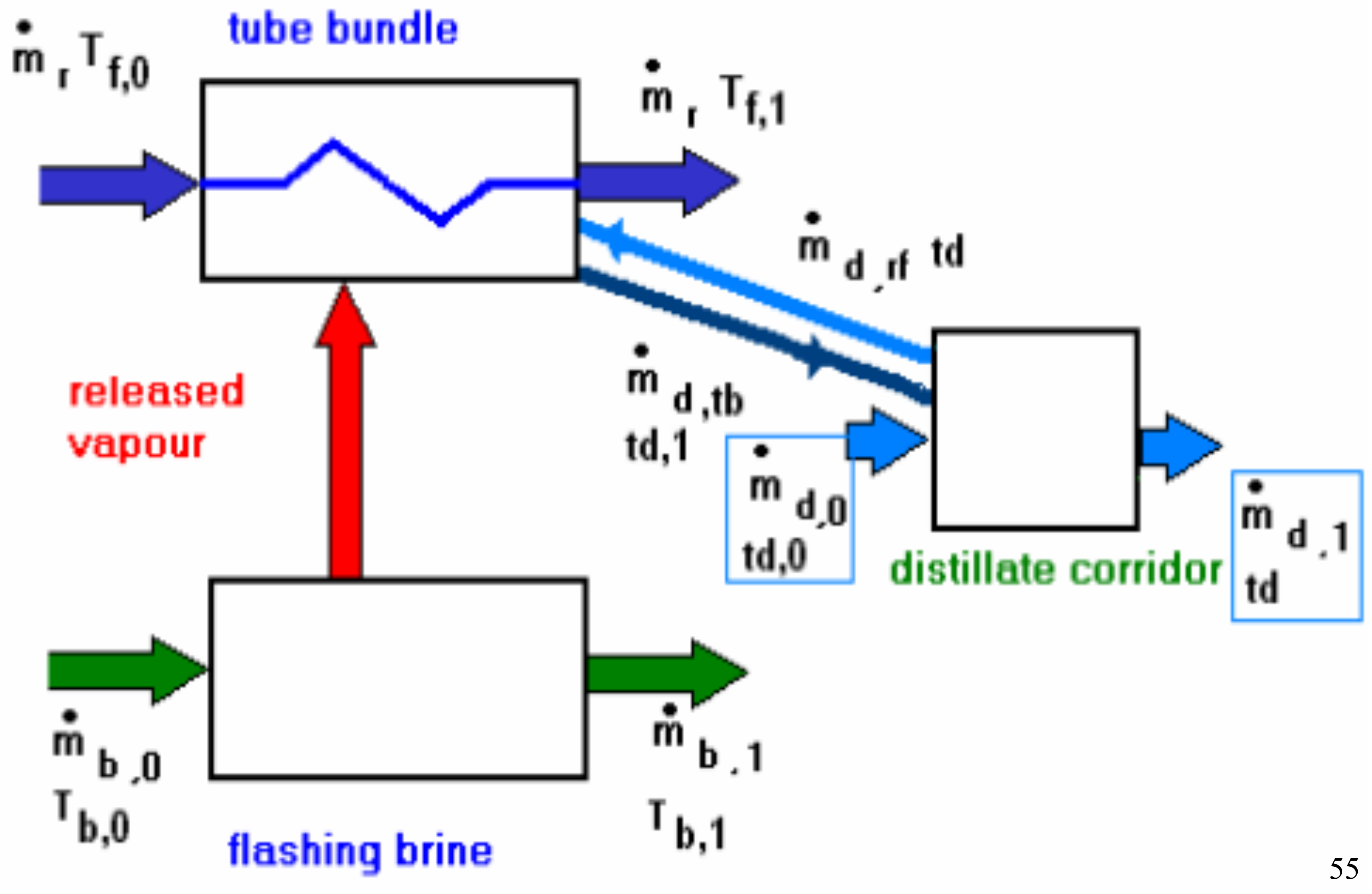
- **Stage model**

The conventional MSF desalination plant stage can be modelled as a system of three control volumes which interact with each other.

These control volumes are:

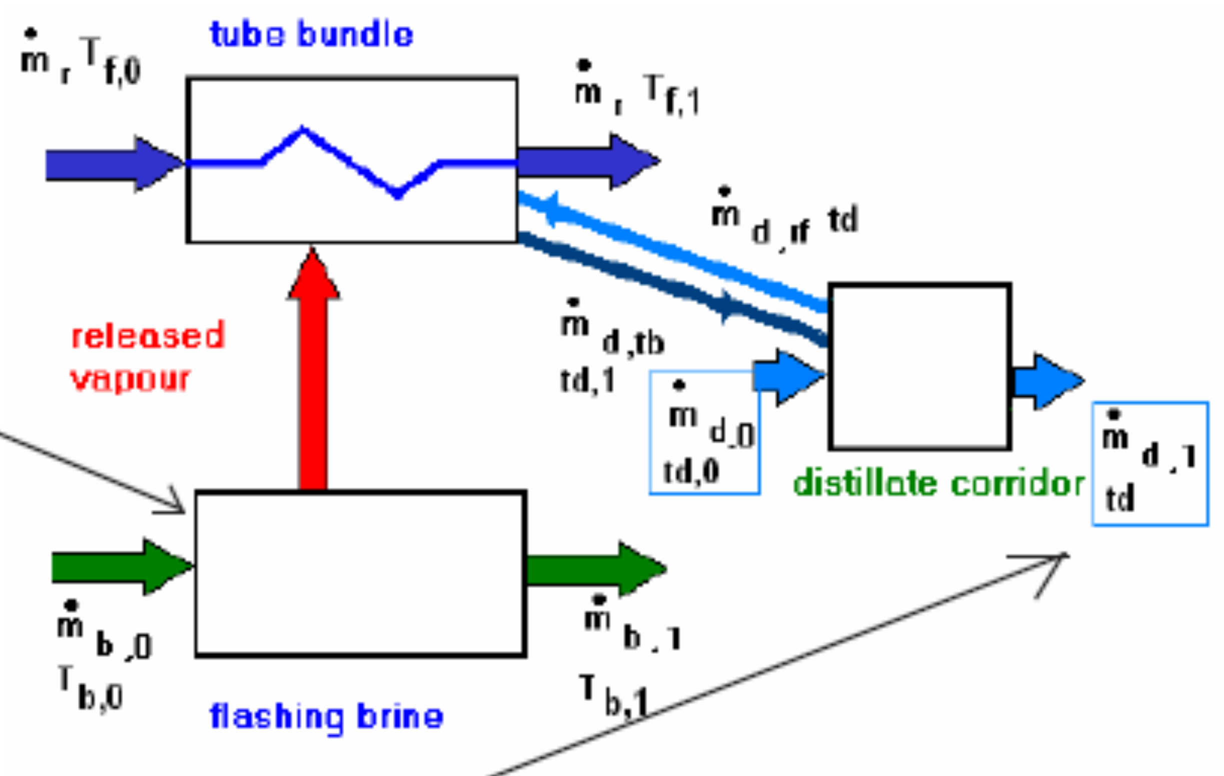
- Flashing brine
- Tube bundle
- Distillate corridor and tray

Stage Model


tube bundle
$\dot{m}_r, T_{f,0}$
$\dot{m}_r, T_{f,1}$
$\dot{m}_{d,rf}$ td
released vapour
$\dot{m}_{d,tb}$
td,1
$\dot{m}_{d,0}$
td,0
distillate corridor
$\dot{m}_{d,1}$
td
$\dot{m}_{b,0}$
$T_{b,0}$
$\dot{m}_{b,1}$
$T_{b,1}$
flashing brine

Stage model: mass balances

Stage model: mass balances



Flashing brine mass balance

$$\dot m_{b,0} = \dot m_{b,1} + \dot m_d$$

Distillate corridor mass balance

$$\dot m_{d,rb} = \dot m_{d,rf} + \dot m_d$$

Stage model: energy balances

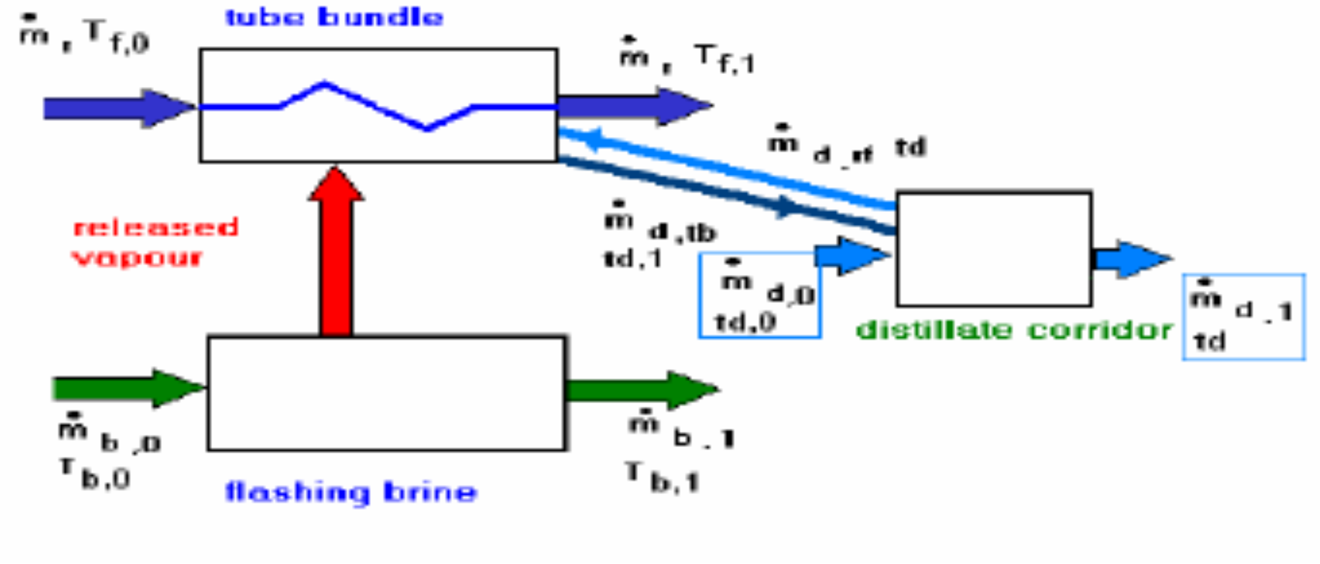
Stage model: energy balances

Flashing brine heat balance

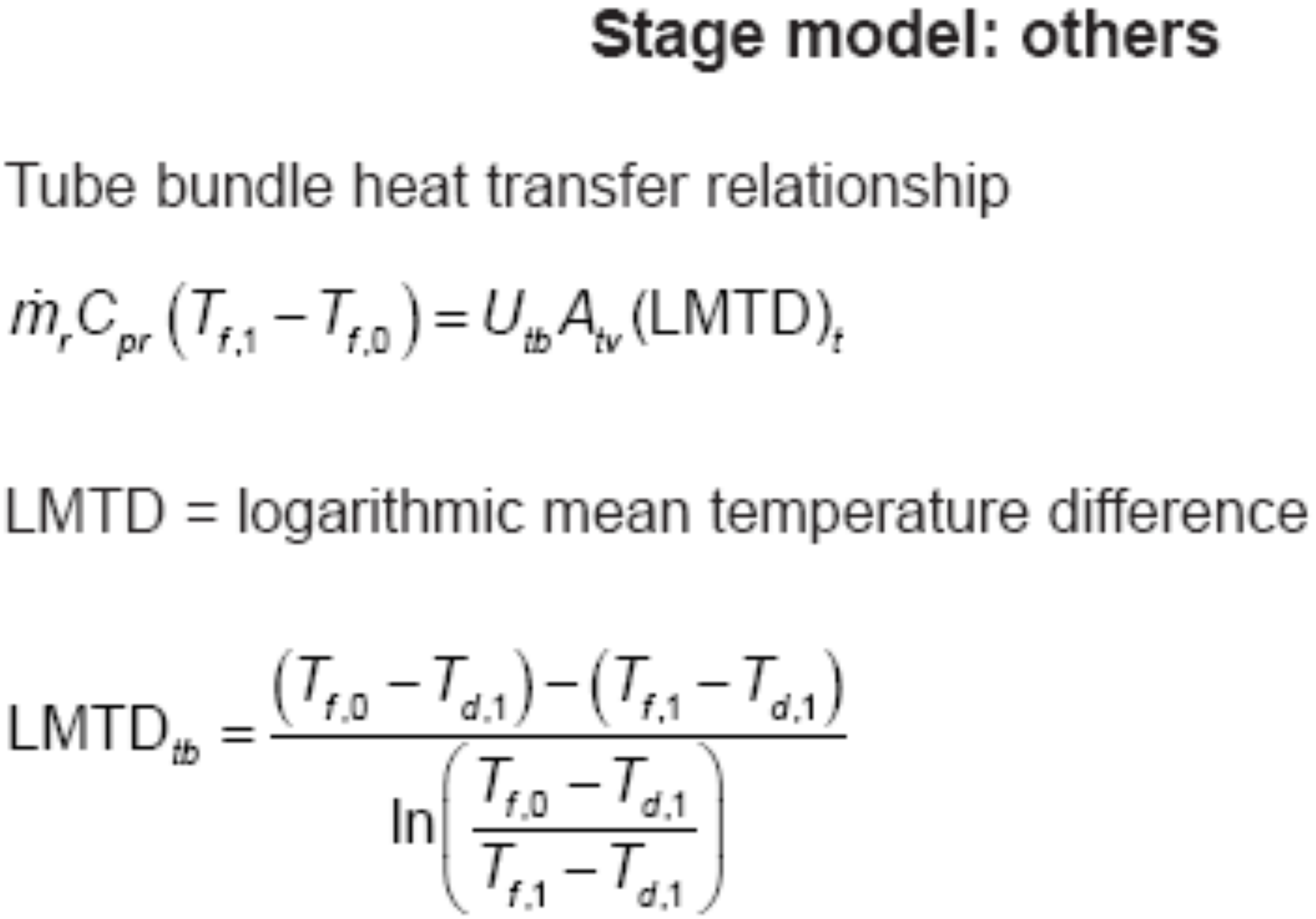
$$\dot{m}_{b,0}C_{p0}T_0 - \dot{m}_{b,1}C_{p1}T_1 = (\dot{m}_{b,o} - \dot{m}_{b,1})h_{f,g}$$

Distillate corridor heat balance

$$\dot{m}_{d,0}C_{pd0}T_{d,0} + \dot{m}_{d,1}C_{pd1}T_{d,1} = (\dot{m}_{d,0} + \dot{m}_{d})C_{pd}T_{d} + \dot{m}_{d,rf}h_{v,rf(Td)}$$



Stage model: others



Stage model: others

Tube bundle heat transfer relationship

$$\dot{m}_r C_{pr}\left(T_{f,1}-T_{f,0}\right)=U_{tb}A_{tv}(\text{LMTD})_t$$

LMTD = logarithmic mean temperature difference

$$\text{LMTD}_{tb}=\frac{\left(T_{f,0}-T_{d,1}\right)-\left(T_{f,1}-T_{d,1}\right)}{\ln\left(\dfrac{T_{f,0}-T_{d,1}}{T_{f,1}-T_{d,1}}\right)}$$

Stage model: others