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Water Recovery Using Waste Heat from Coal Fired Power Plants

Charles W. Morrow, Brian P. Dwyer, Stephen W. Webb, Susan J. Altman

Prepared by

Sandia National Laboratories

Albuquerque, New Mexico 87185 and Livermore, California 94550

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Charles W. Morrow[1]
Brian P. Dwyer[2]
Stephen W. Webb[2]
Susan J. Altman[3]

[1]Advanced Nuclear Fuel Cycle Technologies Department
[2]Geotechnology & Engineering Department
[3]Geochemistry Department
Sandia National Laboratories
P.O. Box 5800
Albuquerque, NM 87185–0735

ABSTRACT

The potential to treat non-traditional water sources using power plant waste heat in conjunction with membrane distillation is assessed. Researchers and power plant designers continue to search for ways to use that waste heat from Rankine cycle power plants to recover water thereby reducing water net water consumption. Unfortunately, waste heat from a power plant is of poor quality. Membrane distillation (MD) systems may be a technology that can use the low temperature waste heat (<100 °F) to treat water. By their nature, they operate at low temperature and usually low pressure. This study investigates the use of MD to recover water from typical power plants. It looks at recovery from three heat producing locations (boiler blow down, steam diverted from bleed streams, and the cooling water system) within a power plant, providing process sketches, heat and material balances and equipment sizing for recovery schemes using MD for each of these locations. It also provides insight into life cycle cost tradeoffs between power production and incremental capital costs.

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EXECUTIVE SUMMARY

The goal of this project was to test the feasibility of reducing the makeup water requirements for a power plant by treating non-traditional water sources (e.g., saline groundwater, boiler blow down, cooling tower blow down, oil and gas production water, municipal/industrial produced water, and seawater) using power plant waste heat in conjunction with membrane distillation (MD). The potential benefits are the production of high quality water for cooling tower and boiler makeup utilizing non-traditional water sources. Membrane distillation technology uses a lower amount of energy as the driving force to clean water in comparison to more traditional reverse osmosis or nanofiltration membranes.

A thermodynamics based membrane distillation model has been developed to integrate into a power plant model in order to determine the amount of water produced using the different energy sources listed above. This model is described in more detail in Section 2.

The power plant model has been developed to evaluate three different energy streams for the use driving membrane distillation:

1) boiler blow down,

2) steam diverted from bleed streams, and

3) the cooling water system.

These streams are schematically shown in Figure 17 and more details are given in Section 3.2.

A comparison between these different options is presented in Table ES-1 and a more detailed description in the paragraphs below. Option 2 (steam diverted from bleed streams) is considered not viable because of the significant impact on plant power production. Options 1 and 3, which truly use waste heat, differ from each other in that significant capital expenditures are necessary for Option 3. With these expenditures the operator will get the advantage of almost an order of magnitude greater production of distillate (that can be used for cooling tower make up).

Boiler blow down is a high energy waste stream that can be used for water recovery. A system to treat the boiler blow down was designed with a series of membrane distillation units in series (Figure 20). The system is designed such that a heat exchanger cools the boiler blow down to a temperature range works with membrane distillation system. After each membrane distillation unit, the water is reheated via the boiler blow down and then treated by the next unit. The effluent has a total dissolved solid level of 1040 ppm temperature of 91 °F and pressure of 20 psia. Total extracted flow is 3.8 lb/s (30.5 gpm). In general, the capital costs for installing a membrane distillation unit is low with the largest piece of equipment being an aerial cooler. An analysis was conducted to determine the impact of the initial temperature of the boiler blow down when it is entering the membrane distillation unit versus the amount of water extracted by the units (as too high temperature may damage the membranes). It was found that the fraction of water extracted by membrane distillation ranged from 0.29 to 0.38. The higher the initial

Table ES-1: Comparison of Different Membrane Distillation Configuration Options.

Location	Advantages	Disadvantages	Capital Equipment Needed	Water Production	Other Notes
Boiler blow down (1 in Figure 17)	High energy per pound Energy is otherwise waste MD can be used to cool boiler blow down, helping the cooling system Low capital costs to implement	Small mass flow rate of water (amount of water will not be high)	Heat exchangers Cooling water pumps Aerial cooler Membrane distillatin unit	3.8 lb/s (30.5 gpm)	Treating water internal to the power plant (boiler blow down)
steam diverted from bleed streams (2 in Figure 17)	High energy per pound Low capital costs to implement	Lead to significant power production loss of plant Small mass flow rate of water (amount of water will not be high)	Heat exchangers Cooling water pumps Aerial cooler Membrane distillatin unit	10.4 lb/s (93.3 gpm)	Treating water external to the power plant
Cooling water system (3 in Figure 17)	High energy quantity Energy is otherwise waste MD unit acts as a heat exchanger as well as a water purifier. Generating makeup water for the cooling tower	Significant capital costs (all new equipment for brackish water loop and increase size of condensing system and cooling water system)	Condenser Cooling water pump Cooling tower duty Brackish water pump Membrane distillatin unit	370 lb/s (3,670 gpm)	Treating water external to the power plant

temperature the higher the fraction of water extracted. The lower the initial temperature the greater number of heat exchangers and membrane distillation units needed, thus the higher the capital costs.

Option 2 examines the use of a bleed stream that would normally be used to preheat boiler feed water. Note that use of the energy of the bleed stream to run a membrane distillation unit will negatively impact the power production of the power plant. Therefore this option is not using waste heat, but heat that would otherwise be used in plant operation. The system designed to use the bleed stream was similar to that using the boiler blow down in that it used a heat exchanger/membrane sequence to cool down the steam to a temperature appropriate for membrane distillation. The exampled modeled used four heat exchanger/membrane pairs, though the design is relatively insensitive to the number of pairs used. The membrane distillation units produce a total of 10.4 lb/s (93.3 gpm) of distillate at a temperature of 90.6 °F. TDS of the concentrate would be approximately 7,300 ppm.

The cooling water system (option 3) appears to be an ideal source of energy for membrane distillation. It is the largest source of waste heat in a plant. Because of the low temperature (101.7 °F for the plant used in this study) of this waste heat source, it is difficult to use. However, this temperature is ideally suited for membrane distillation. The model used to assess this option added a loop, referred to as the brackish water loop, between the steam condenser loop and the cooling water loop. Brackish water from an external source is treated in this loop and the distillate added to the cooling tower system as make up water. This system produces 370 lb/s (3,670 gpm) of treated water. For this option, a much more significant change must be made to the power plant in order for the membrane distillation operations to work. A choice must be made between significant expenditures for capital equipment needed or sacrificing some power output by increasing condenser temperature and pressure.

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1 INTRODUCTION

This report presents an assessment of the feasibility of recovering water from non-traditional impaired water sources using power plant waste heat and membrane distillation technology. More specifically, the primary objective of this study is to reduce the makeup water requirements for a power plant boiler, wet cooling towers, and flue gas desulfurization (FSD) by developing the capability to treat non-traditional water sources (saline and brackish groundwater, boiler blowdown, cooling tower blowdown, surface water, oil and gas hydraulic-fracture and production water, municipal/industrial produced water, and sea-water) using power plant waste heat in conjunction with membrane distillation (MD). The potential benefits include: (1) production of high quality water for cooling tower and boiler makeup utilizing non-traditional water sources and (2) replacement of a portion of steam condenser/cooling tower capacity with heat exchanger/MD units.

1.1 Problem Statement

In the United States, approximately 89% of the energy produced in power plants is generated by thermoelectric systems, which evaporate water during the cooling of the condenser water [1]. Eighty to ninety percent of the power plant raw water usage is through a combination of cooling tower evaporation and blowdown. [2] Figure 1 shows water use by power plant type. Raw water usage is defined as the total amount of water to be supplied from local water resources to

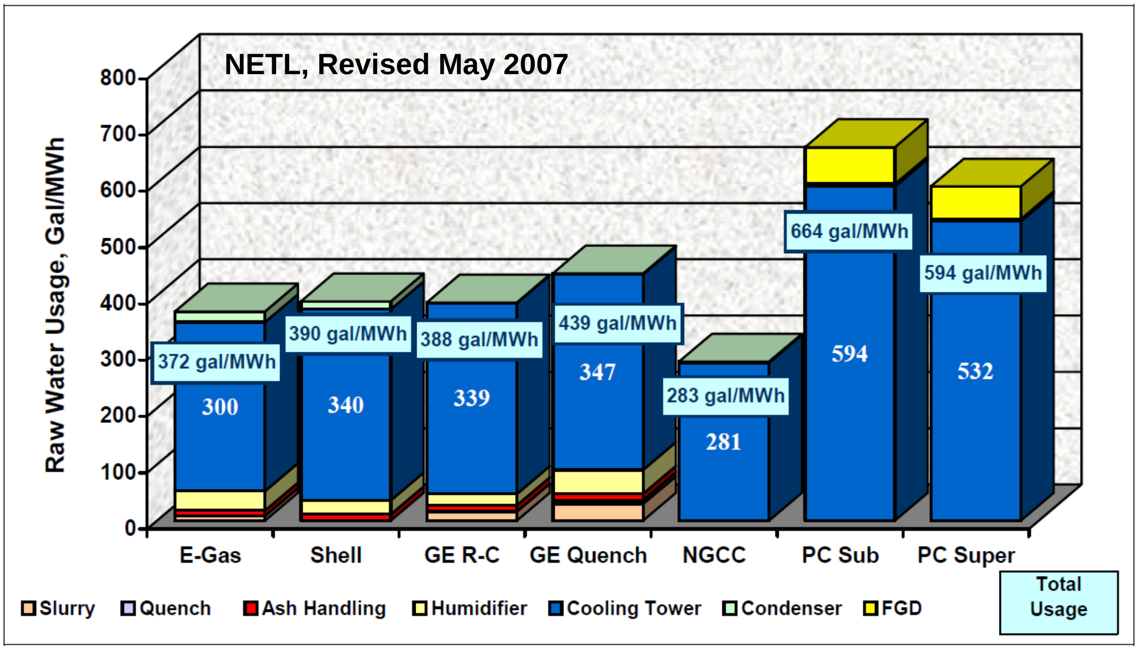

Figure 1: Comparison of Raw Water Usage for Various Fossil Plants, gallons per MWh.

provide for the needs of the plant. For example, the pulverized coal super-critical power plant example (PC Super) shows 594 gallons of water are used per MWh of electricity produced. If we assume 80% of that is lost to evaporation, that is 475 gallons/ MWh electricity produced.

The interdependence of water and energy production becomes more clear as water becomes more scarce. As the United States population grows, so does the need for water and energy. At present, freshwater withdrawals already exceed precipitation in many areas across the country, as illustrated in Figure 2 below. The figure shows the ratio of total freshwater withdrawals in all counties in the U. S. divided by available precipitation (precipitation minus evapotranspiration) shown as a percentage. The figure provides an indication of the areas where current water demands are being met with significant groundwater pumping or transport of surface water from other locales.[3]

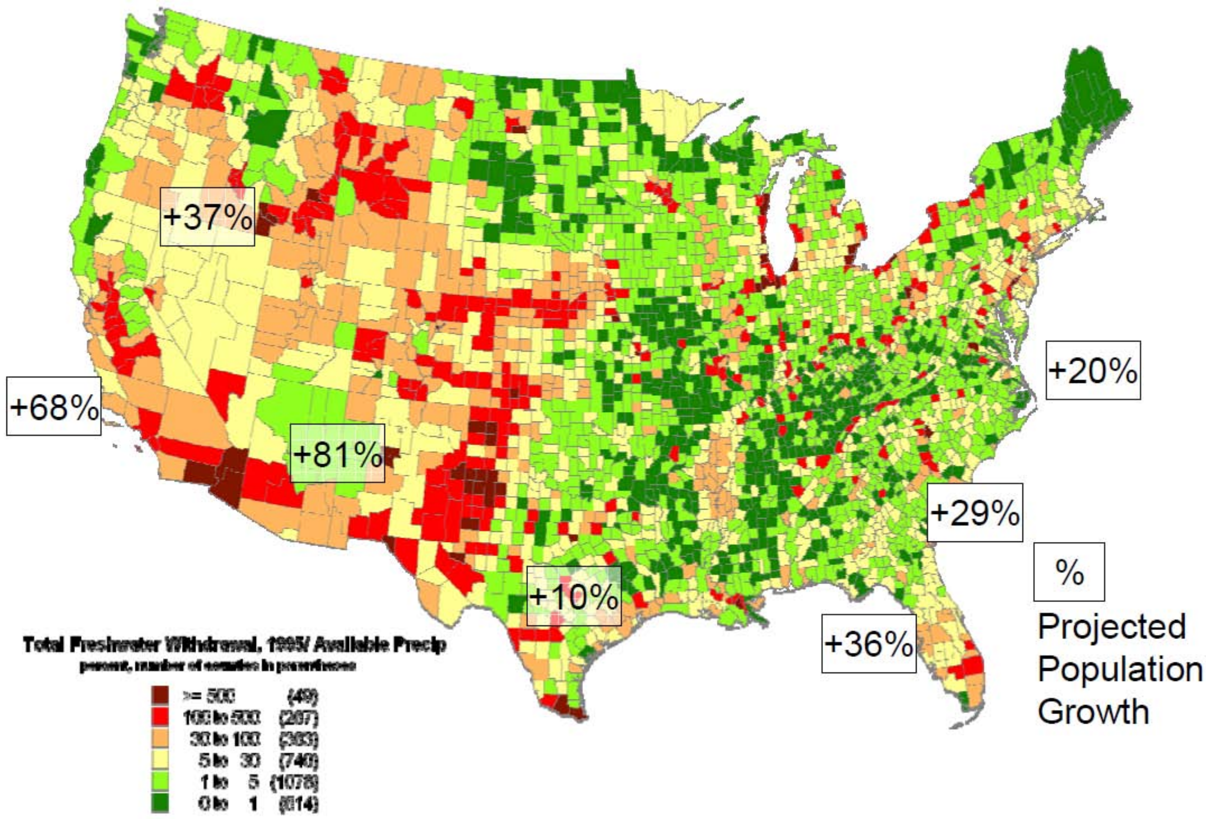

Figure 2: Water Shortages and Population Growth [3]

Future projections indicated water shortages in the coming decades (Figure 3). As the population grows the impact of power generation on future water demand depends on the type of generation installed and the rate at which existing plants are retired. Under a business-as-usual case, where most new power is provided by water-cooled thermoelectric power plants, the most dramatic changes will occur if old plants using seawater or freshwater open-loop plants are

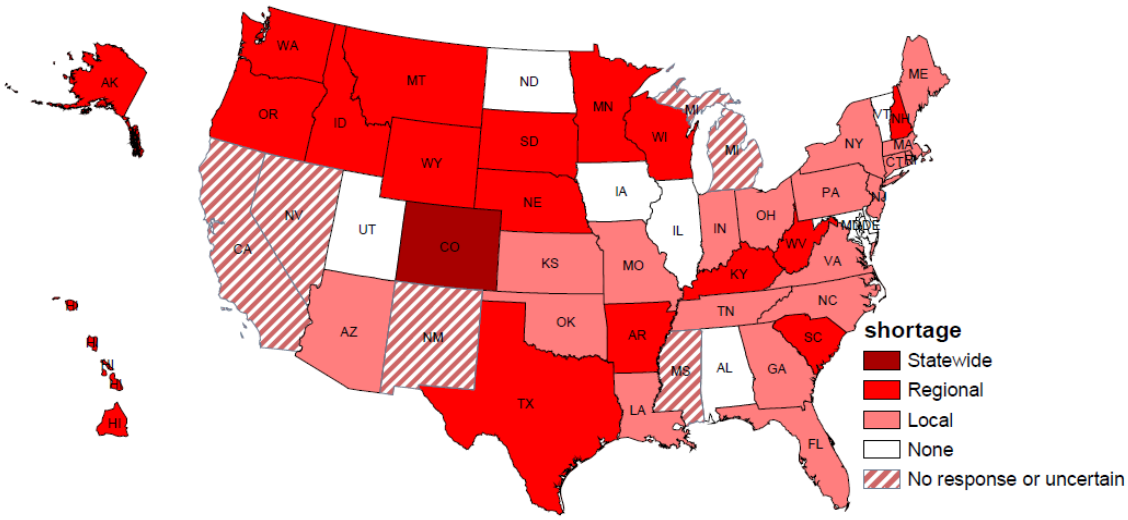

Figure 3: Survey of Likely Water Shortages over the Next Decade under Average Conditions [4]

retired, and replacement plants as well as capacity additions are installed with evaporative closed-loop cooling.[4]

The earth has plenty of water available, but at what cost? The water is either not in the right location, or is not clean enough. Herein comes the cost. Sandia National Laboratories proposes that the most economical solution is to clean non-traditional water sources and recycle power plant blowdown waters using membrane distillation. The primary costs savings comes from using power plant waste heat as the energy source to drive the treatment process.

This study included the construction of a power plant model was built based upon Case 11 from a DOE/NETL report – Cost and Performance Baseline for Fossil Energy Plants. Case 11 is a design for a supercritical coal fired steam plant without CO_2 removal.[5][1] Next a membrane model based upon Direct Contact Membrane Distillation technology was built and integrated into the power plant model to evaluate the potential water recovery outcomes.

1.2 An Introduction of Steam Power Plants

Figure 4 contains a schematic of a modern steam power plant. The extraction process used forms a loop or cycle described by the following numbered paragraphs. The paragraph numbers are repeated on Figure 4 to assist the reader.

1. Warm high pressure water enters the boiler.
2. The boiler heats the water further so that it exits as high pressure superheated vapor to be expanded through a train of turbines losing pressure and temperature along the way.

3. The process used in this study reheats the steam after the first turbine to increase the overall energy extraction.
4. On leaving the lowest pressure turbine, steam enters the condenser where it condenses to liquid.
5. The cooling water system takes this latent heat away for rejection to atmosphere.
6. A system of boiler feed pumps re-pressurize the water for another lap around the cycle.
7. Feed pre-heaters re-warm the water to reduce demand in the boiler. The model used for this study bleeds steam off of the turbine train to provide the heating media for these heaters. These bleed streams were omitted from this drawing to avoid complexity.

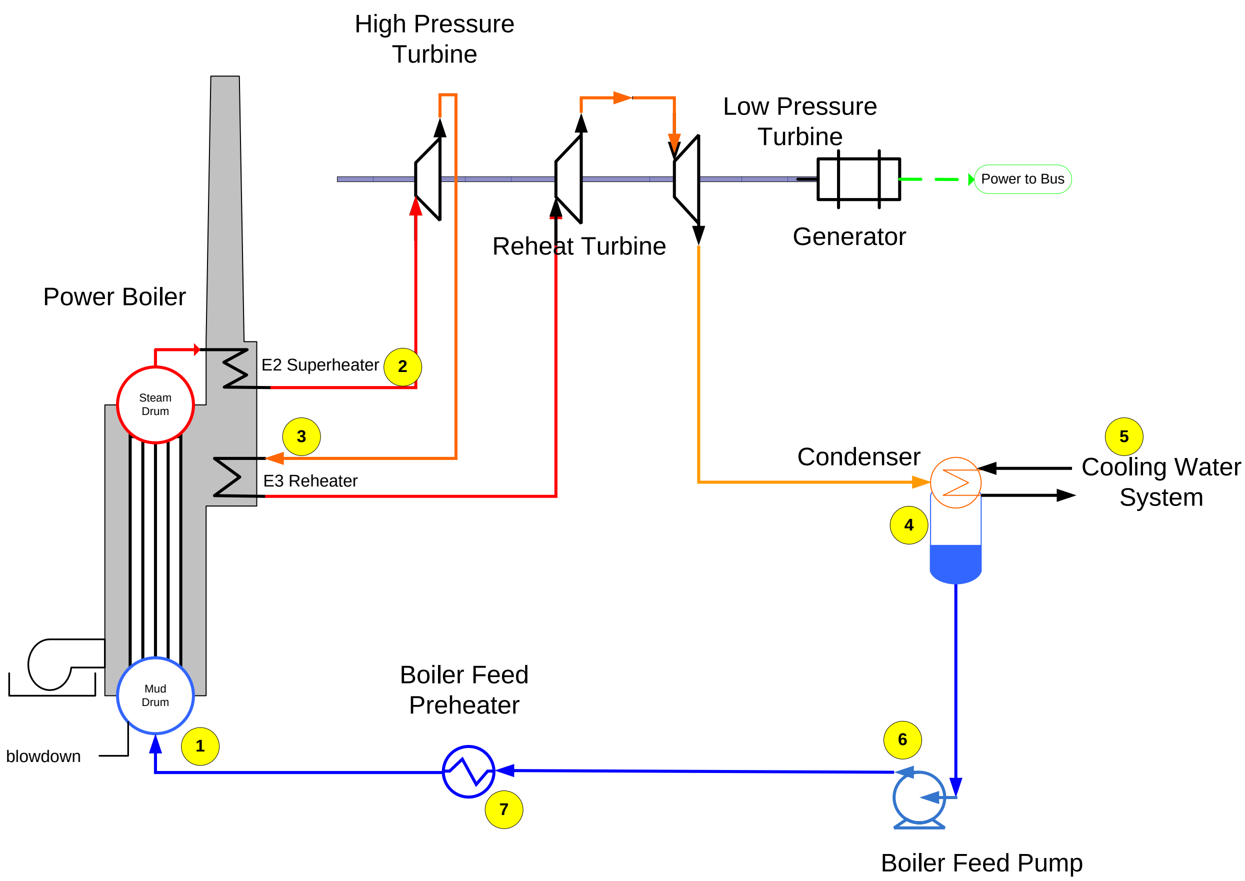

Figure 4: Simplified Schematic of Rankine Cycle Power Plant

The power plant model used for this study has an overall efficiency of 39.1% and produces 550 MW of electrical power for sale. This relatively high efficiency is typical of modern plant designs. Nonetheless, it must reject 589 MW to the cooling water system. At the same time, it consumes 78,000 gallons per day of water.

1.3 Membrane Distillation Technology

Membrane Distillation (MD) is an emerging separations technology that is traditionally accomplished by conventional distillation or reverse osmosis (RO). In contrast to RO where water is forced via high pressure through a membrane, MD uses low-grade thermal energy to heat water; thereby increasing the vapor pressure causing evaporation of clean water across a porous hydrophobic membrane. This process requires much less energy input than RO or multi stage flash (MSF) because it operates at near atmospheric pressure and is not subject to the osmotic pressure driven limitation of RO. It is for this reason that a low grade thermal energy source such as power plant waste heat is potentially enough energy for MD to treat non-traditional water. The physics of the MD technology is discussed in more detail in Section 2.1.

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2 MEMBRANE DISTILLATION MODEL

This Chapter describes the physics of membrane distillation and the thermodynamics and heat transfer used to model that physics. The physics of membrane distillation is described in Section 2.1. The chapter also includes sections that describe conversion of this general physics into a model. Section 2.2.1 presents the mathematics of the model. Section 2.2.2 then explains the numerical implementation. Section 2.2.3 finally discusses the thermodynamic water properties used by the model. Sections 2.2.4 introduces the issue of sensible heat transfer in membranes. Finally, Section 2.3 describes in some detail the two membrane filter schemes used almost exclusively in the water recovery processes described in subsequent chapters.

Several variants exist regarding membrane distillation designs. This study uses the Direct Contact Membrane Distillation (DCMD) scheme. This process is the simplest to model. It is also the most straightforward with the fewest parts. It therefore lends itself most easily to waste recovery schemes. .[6][7]

2.1 Membrane Distillation Technology

The overwhelming advantage of a membrane distillation process over other water treatment methods is its low operating temperature. Most boilers or other distillation processes involve boiling water at 212 °F or higher. Not so membrane distillation. These units are constrained to operate <u>below</u> water's boiling point.

Consider Figure 5, a plot of water vapor pressure as a function of temperature. As one would expect, at 212 °F, the vapor pressure is 14.7 psia. Of importance to membrane distillation units

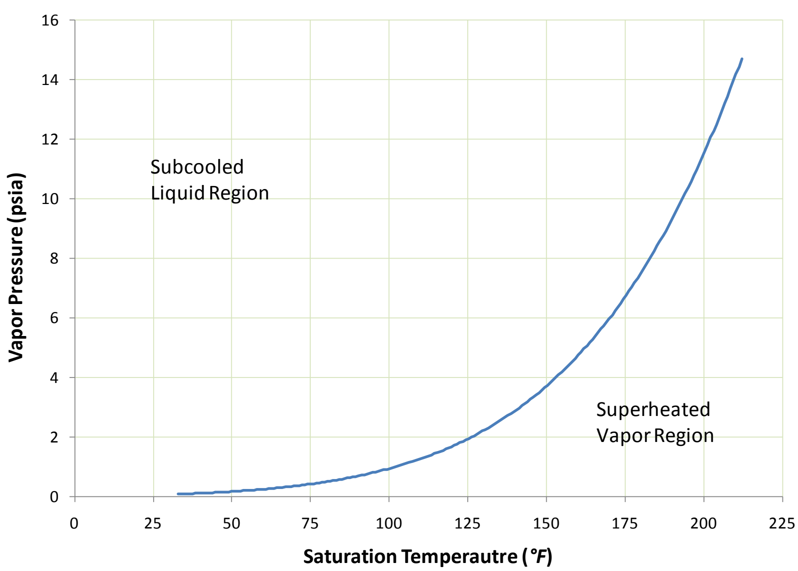

Figure 5: Vapor Pressure vs Temperature for Water

is the fact that water exerts a vapor pressure even at lower temperatures. Equally important is the fact that the vapor pressure drops with dropping temperature. Therefore, a temperature difference means a vapor pressure difference.

For membrane distillation units, the other two important properties of water are its surface tension and its propensity not to wet some surfaces, defined as hydrophobic surfaces. Consider the droplet of water shown on the left of Figure 6. The water's surface tension pulls the droplet away from the hydrophobic surface. Now consider the water droplet on the right in the figure. The hydrophobic nature of the solid surface pushes the water away from the hole. The water's surface tension holds the droplet together and allows the water to span a small diameter hole.

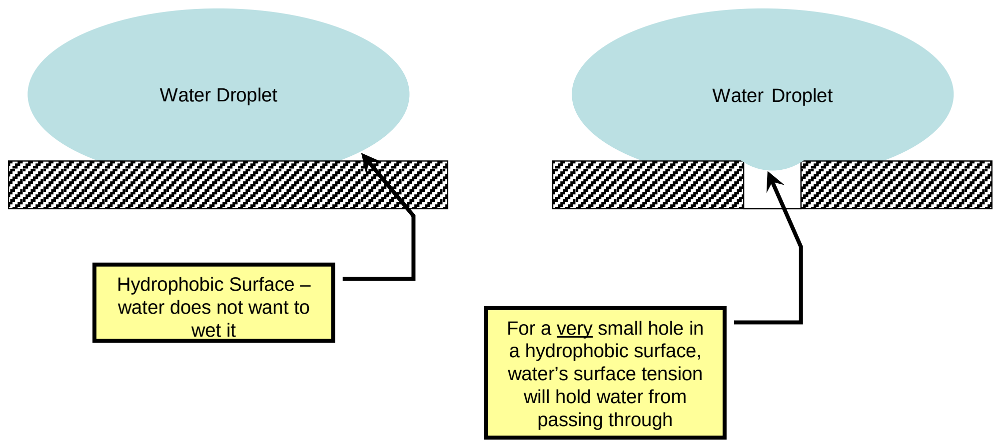

Figure 6: Concepts Behind Membrane Distillation

The dimple the water forms over the entrance to the hole is called a meniscus. It is important to note that the liquid phase must be present. Without it, we have no surface tension and the meniscus would not form.

Figure 7 shows how membrane distillation combines these phenomena to advantage. Start with a thin porous hydrophobic material with a lot of microscopic through passages. Teflon®, for example, works well. Put relatively hot but not boiling water on one side and colder water on the other. The hydrophobic material and the waters surface tension keep liquid phase interstitial pores of the material. However, a water vapor phase will form there. Because of its relatively higher vapor pressure, water on the hot side will pass into the vapor phase. The lower vapor pressure of the cold side means that vapor phase water will tend to condense into the cold side water. This constitutes a mass transport between the hot and the cold water streams. Because process is a distillation process, the transport will be of pure water. This mass transport, $\dot{m}_x$, across the membrane has been modeled successfully as a linear function of the difference between the vapor pressure of the hot fluid, Pv_h and the vapor pressure of the cold fluid, Pv_c:

$$\dot{m}_x = AK_J\left(Pv_h - Pv_c\right) \qquad 1$$

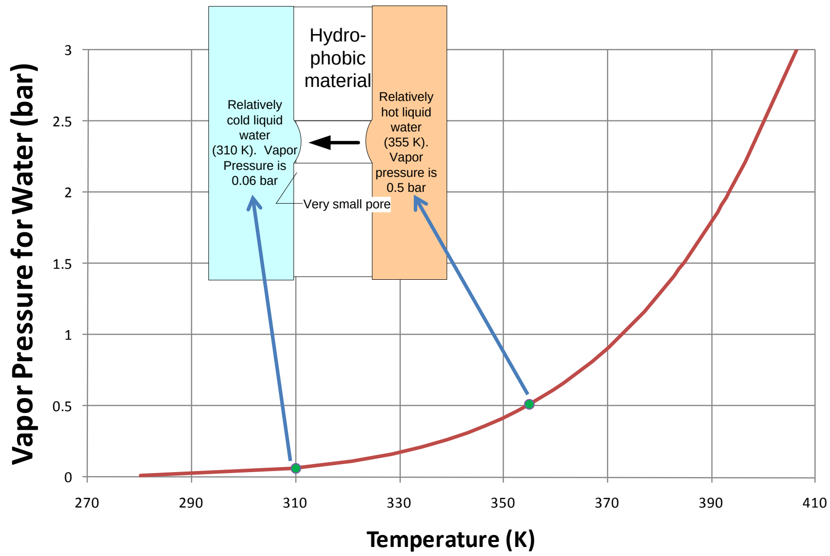

Figure 7: Schematic of Membrane Distillation unit Operation

By convention, the area, *A*, used in the above equation is the total area of the membrane instead of the area of the passages. The proportionality constant, K_J, depends on several factors including diffusivity of the water vapor through the membrane's pore volume, bulk temperature of the membrane, membrane porosity, total pressure within the membrane, membrane thickness, and average passage size. However, it tends to be relatively constant in the range of interest.[6][7][7][8]

2.2 Model Formulation

2.2.1 Membrane Distillation Unit Model

In practice, a membrane distillation unit looks like a counter-flow heat exchanger. However, in the case of the membrane distillation unit, the separating partition passes both energy and mass. The model assumes that both approaches are known and constant but not necessarily equal. The approach is defined as the absolute of the difference between exiting fluids in one stream and entering (or approaching) fluids in the other. In other words:

$$\begin{aligned} \Delta T_A &= T_{h,0} - T_{c,0} \\ \Delta T_B &= T_{h,n} - T_{c,N} \end{aligned} \qquad 2$$

Figure 8 defines the temperatures in the above equation. It also lays out the nomenclature used subsequent equations. For computational purposes, the exchanger is subdivided into *n* sequential segments identified by the subscript *i*. Each segment can be treated as two control volumes

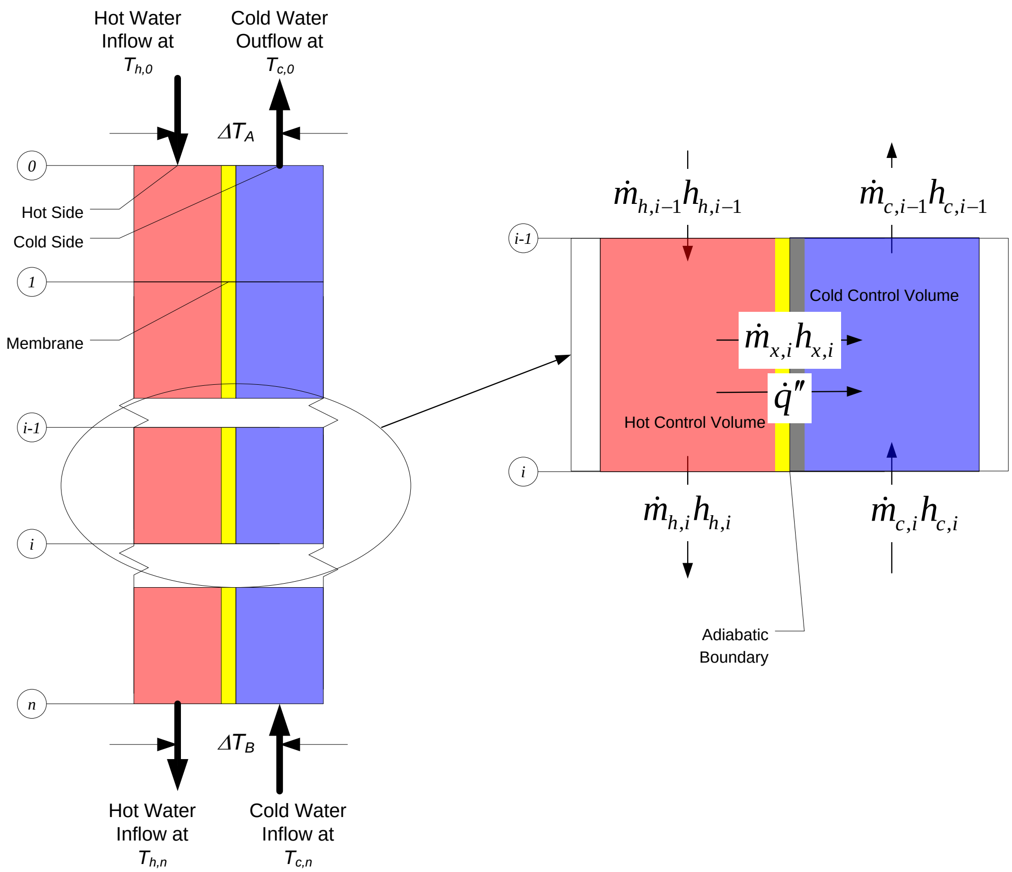

Figure 8: Thermodynamic Control Volumes

similar to the two control volumes shown on the right in Figure 8. The two control volumes are linked by the heat and mass transfer between them.

Figure 9 shows graphically temperature variation along the membrane. Make note of the following important items:

1. The counter flow design means that the temperature difference between the two streams stays as low as practicable.
2. The temperature drop through the unit is almost, but not quite, linear.
3. Approach temperatures for heat exchange equipment are typically set by economics. Reasonable values range from 5 ºF and up with values usually between 10 and 20 ºF [9].

The membrane distillation model will accept as input different approaches at each end. However, it defaults to equal approaches, as this typically results in the most compact design.[10]

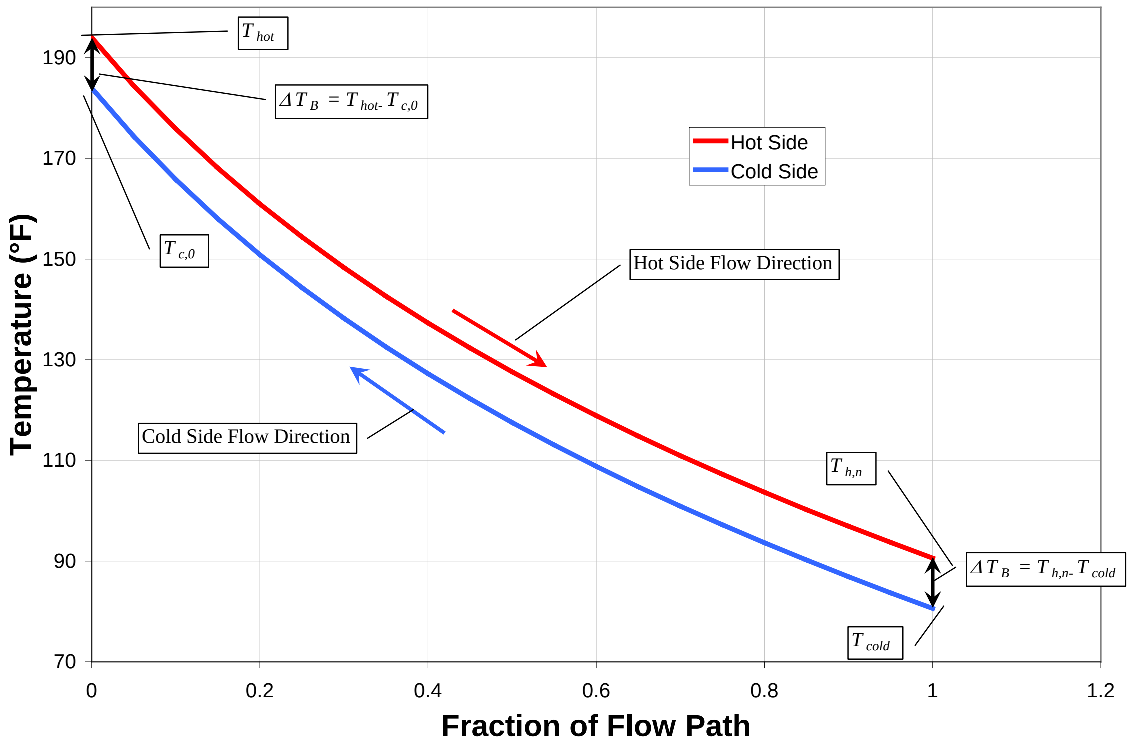

Figure 9: Temperature Variation Through a Typical Membrane Distillation Unit

Assume the thermodynamic systems to be modeled are steady state with no work. Assume further that all external surfaces are sufficiently insulated to be adiabatic. The energy balance for the i^{th} segment on the hot side would be written as

$$0 = -A_i \dot{q}_i'' + \dot{m}_{h,i-1} h_{h,i-1} - \dot{m}_{x,i} h_{hx,avg} - \dot{m}_{h,i} h_{h,i} \quad 3$$

The control volume boundary at the transfer plain occurs in the vapor phase of the membrane pores between the two control volumes. For that reason, $h_{hx,avg}$ is the enthalpy of the saturated vapor phase at the average temperature of the hot control volume. The area term, A_i, refers to the total area of the surface shared by the two control volumes. In this case,

$$A_i = w_i \frac{L}{n}$$

Where w_i is the dimension of the membrane orthogonal to flow and L is the total length of the membrane's flow passage. The model assumes w_i is constant. The integer n denotes the number of control volumes into which the membrane is subdivided. Mass balance for the hot side can be written as

$$\dot{m}_{h,i-1} = \dot{m}_{h,i} - \dot{m}_{x,i} \qquad 4$$

Energy balance for the i^{th} control volume on the cold side can be written as

$$0 = A_i \dot{q}_i'' - \dot{m}_{c,i-1} h_{c,i-1} + \dot{m}_{x,i} h_{cx,avg} + \dot{m}_{c,i} h_{c,i} \qquad 5$$

The control volume boundary at the transfer plain occurs in the vapor phase between the two control volumes. For that reason, $h_{cx,avg}$ is the enthalpy of the saturated vapor phase. This enthalpy is evaluated at the average temperature of the control volume. Mass balance for the i^{th} control volume on the cold side is

$$\dot{m}_{c,i-1} = \dot{m}_{c,i} + \dot{m}_{x,i} \qquad 6$$

Mass flux across the membrane depends on the vapor pressure difference across the membrane

$$\dot{m}_{x,i} = A_i K_j \left(Pv_{h,i} - Pv_{c,i} \right) \qquad 7$$

Vapor pressures are functions of the average temperature

$$T_{h,i,avg} = \frac{T_{h,i} + T_{h,i-1}}{2}, \; T_{c,i,avg} = \frac{T_{c,i} + T_{c,i-1}}{2} \qquad 8$$

Heat transfer from hot side to cold side is

$$\dot{q}_i'' = U \left(T_{h,iavg} - T_{c,iavg} \right) \qquad 9$$

The overall heat transfer coefficient, U, is modeled as

$$U = \frac{1}{\left(\frac{1}{h_h} + \frac{t_m}{k_m} + \frac{1}{h_c} \right)} \qquad 10$$

The two heat transfer coefficients h_c and h_h for the hot side and cold side respectively can be calculated using standard heat transfer correlations. They will differ slightly reflecting different temperatures and velocity fields. However, this model assumes they are equal and constant (see Table 1). This same table contains the values assumed for membrane heat transfer coefficient, k_{m}, and membrane material thickness, t_m.

Table 1 lists assumed values for the parameters used in the above equations. Values taken from literature and were originally in SI units, but were converted to US customary units for this paper [11] [12]

Table 1: Constant Parameters Used in Model

Description	Symbol	Value	Units
Total Area	w_iL_i	0.011	*ft*^{2}
Thermal Conductivity	k_m	130	*Btu/h/ft/°F*
Membrane Thickness	t_m	0.00041	*ft*
Membrane Permeability	K_j	3.05E-07	*s/ft*
Heat transfer Coefficient	h_h	1760	*Btu/h/ft²/°F*
	h_c	1760	*Btu/h/ft²/°F*
Control Volumes	n	20	*nd*

2.2.2 Numerical Approach

The Membrane Distillation unit model was implemented as several functions in Microsoft Excel® using Visual Basic for Application (VBA), Microsoft's implementation of ANSI Basic. The model accepts inputs in a manner similar to a normal spreadsheet function. It returns an array as output. The array contains a lot of information about membrane performance.

Counter Flow system models typically use finite element or other large matrix solutions in their solution. These approaches are both fast to run and robust. However, they are time consuming to write and therefore expensive to program. For the purposes of this study, a relatively slower but more cost effective iterative model was acceptable.

The iterative sequence is as follows:

1. Assume cold mass flow = 1.0 lb/s/ft^2. Assume hot mass flow = 1.0 lb/s/ft^2.
2. Using the specified approach temperatures, interpolate a first guess at the hot side and cold side temperature distributions assuming temperature varies linearly through the device. As Figure 9 shows, this assumption is not true, but it provides a reasonable starting point.
3. For each of the *n* segments of the subdivided membrane:
    a. Using the temperatures from step 2, calculate average cell hot and cold temperature using equation 8.
    b. Using the calculated average temperatures, calculate vapor pressure (see Section 2.2.3 immediately following this section) and from there, mass transferred across the membrane using equation 7.
    c. Calculate heat transferred across the membrane using equation 9.
    d. Calculate the mass leaving the control volumes using equations 4 and 6.
    e. Using a shooter technique, balance equations 3 and 5 solving for cell exit temperatures, $T_{h,i}$ and $T_{c,i}$. Note that these two temperatures interrelate so that balancing equation 3 will change the balance on equation 55.[13]
    f. Repeat step 6 until neither of the two temperatures $T_{h,i}$ and $T_{c,i}$ change between iterations.
4. Calculate $\Delta T_{B.}$ This value will probably not match the specified input.

5. Use a shooter routine to find the hot side inflow, $\dot{m}_{h,0}$, that returns correct value of ΔT_B.[13]. This step will require multiple iterations of steps 3 and 4. For this step, assume $\dot{m}_{h,0} = \dot{m}_{c,n}$.
6. Use a shooter routine to find the cold water inflow, $\dot{m}_{c,n}$, that returns correct value of ΔT_A. This process will require multiple iterations of steps 3 through 5.

The algorithm reports out numerous things as an array function, including transferred mass (lb/s/ft^2), transferred heat (Btu/s/ft2), and temperature data. All information is available at the inflow and outflow conditions as well as at each boundary between control volumes.

2.2.3 Water Properties

For computation efficiency, the computer programs resulting from the algorithms described above used custom correlations of water properties. These properties came from the following two computer programs:

- The U.S. National Institute of Science and Technology (NIST) Chemistry WebBook [14]
- NIST Standard Reference Database 23 (REFPROP) [15]

The Webbook is an internet based application that can be convenient for obtaining bulk information but does not support interface with other applications. The REFPROP program on the other hand supports interface with Microsoft Excel® but can be time consuming to set up. Both sources derive from the 1997 implementation of water and steam properties from the International Association for the Properties of Water and Steam. Both sources use identical IAPWS algorithms and datum points so calculations using either one are consistent.[16]

Vapor pressures for input to equation 7 are based on the data and correlations plotted in Figure 10.

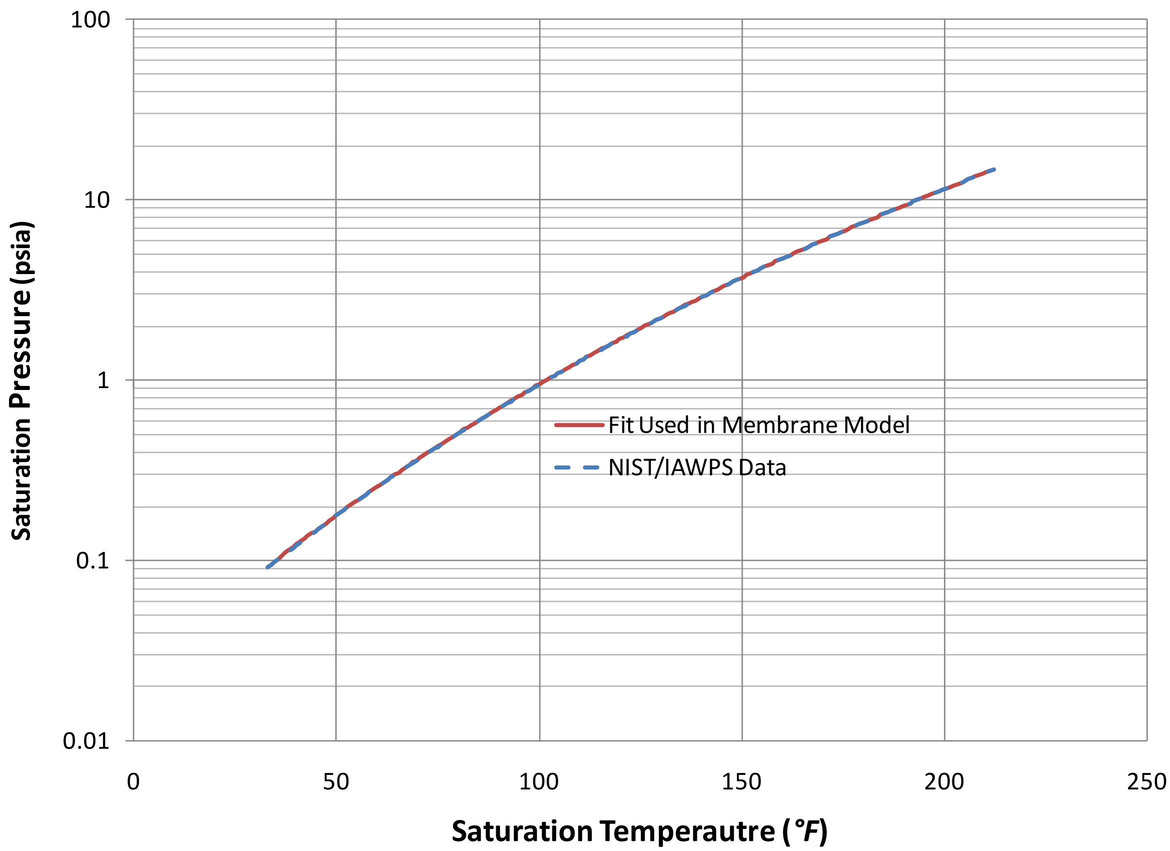

Figure 10: Water Vapor Pressure With Correlation

The vapor pressure correlation plotted in Figure 10 was developed based on 201 vapor pressure points spanning the narrow temperature range of interest here. The data points were downloaded from the NIST Chemistry Webbook site introduced above. The correlation process used a non-linear variance minimization technique. The resulting correlation for vapor pressure as a function of temperature is

$$\ln(P_{v,psia}) = 14.6383 - \frac{7181.25}{T_{°F} + 388.980} \qquad 11$$

Input temperature is in units of °F. Vapor pressure has units of psia. Standard error of the estimate for the correlation is 0.00239 psia.

Figure 11 contains plots of data correlations for water in the temperature range of 32 °F and 212 °F. This range between freezing and boiling water is the range pertinent to this study.

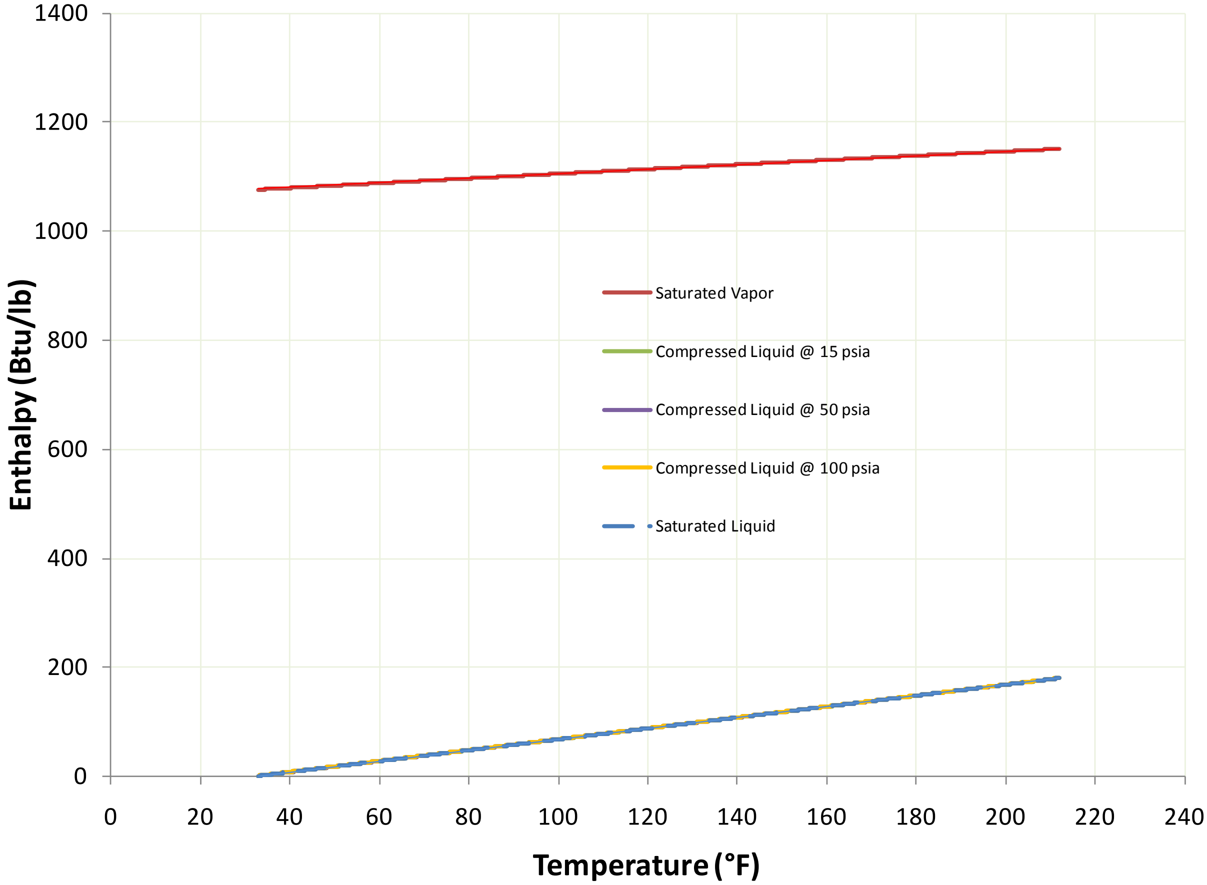

Figure 11: Water Enthalpy

For enthalpy, data for the correlations was extracted using the REFPROP application. Figure 11 contains a single plot for the vapor phase at saturated conditions. It also contains plots of liquid enthalpy for saturated liquid as well as for several compressed liquid conditions. These multiple liquid enthalpy lines effectively overlie each other demonstrating liquid enthalpy is effectively independent of pressure.

The correlation for saturated vapor enthalpy as a function of temperature was developed by assuming a linear relationship and using standard least squares regression algorithms to derive the following equation:

$$h_v = 1063.2 + 0.4193 \cdot T_{\circ F} \qquad 12$$

Input temperature must have the units of °F. Resulting enthalpy has the units Btu/lb. The standard error of the estimate for this correlation is 0.392 Btu/lb, 0.03% of the enthalpy at 120 °F.

The correlation for all liquids' enthalpy was also assumed to be linear and was developed using standard least squares regression algorithms. All liquid data is correlated to a single line which was developed using standard linear regression techniques and statistics.

$$h_l = -31.9984 + 1.0008 \cdot T_{^\circ F} \tag{13}$$

As above, input temperature must have the units of °F. Resulting enthalpy has the units Btu/lb. Standard error of the estimate for correlation when compared to saturated liquid is 0.0473 Btu/lb, 0.054% of the enthalpy of saturated water at 120 °F. Standard error of the estimate when compared to 100 psia compressed liquid is 0.255 Btu/lb, 0.29% of the enthalpy for 100 psia water at 120 °F.

2.2.4 Impact of Sensible Heat Transfer

State of the art for membranes means they must be thin in order to support mass transfer. Unfortunately, a thin membrane with a finite (realistic) heat transfer properties means a membrane also transmits energy from the hot side to the cold side via conduction. This conduction is the primary reason behind the declining hot side/increasing cold side temperatures so evident in Figure 9. Without this conduction, the hot fluid would flow through the unit with little change, as would the cold fluid. The temperature difference between the two sides could be as much as an order of magnitude larger. This higher temperature difference could translate into much higher vapor pressure differences and much more effective water purification systems.

2.3 Results and Insight into Membrane Distillation

In chapters 4 5 and 6 discussion centers on specific processes that recover water using membrane filters. Often in these discussions of large process systems details on the workings of the membrane itself gets lost. This section is meant to remedy that loss. It begins by presenting some overall curves for membrane unit sizing that were extracted from the model validation process. It follows up by presenting two prototypical membrane designs. The first, a model result for high temperature feed is typical of the designs used in Chapter 4, Water Recovery Using Boiler Blow Down and Chapter 5, Water Recovery Using Bleed Streams. The second, a model result for low temperature feed is typical of the designs used in Chapter 6, Cooling Water System with Water Recovery. In all cases the results here are prototypical. Actual designs used in subsequent chapters may vary slightly in approach temperature, cold water feed temperature and other parameters.

The graph in Figure 12 shows specific membrane area in square feet per pound per second of water distilled on the left axis and its inverse on the right axis. The right axis uses SI units to facilitate the comparison of results comparison to published data. In the literature, flux is generally on the order of 6 to 15 kg/h/m² depending on the hot water feed temperature. [12][11] These values match our model results.

This figure is based on model runs with a constant approach temperature of 10 °F and a cold water feed temperature, $T_{c,n}$ of 60 °F.

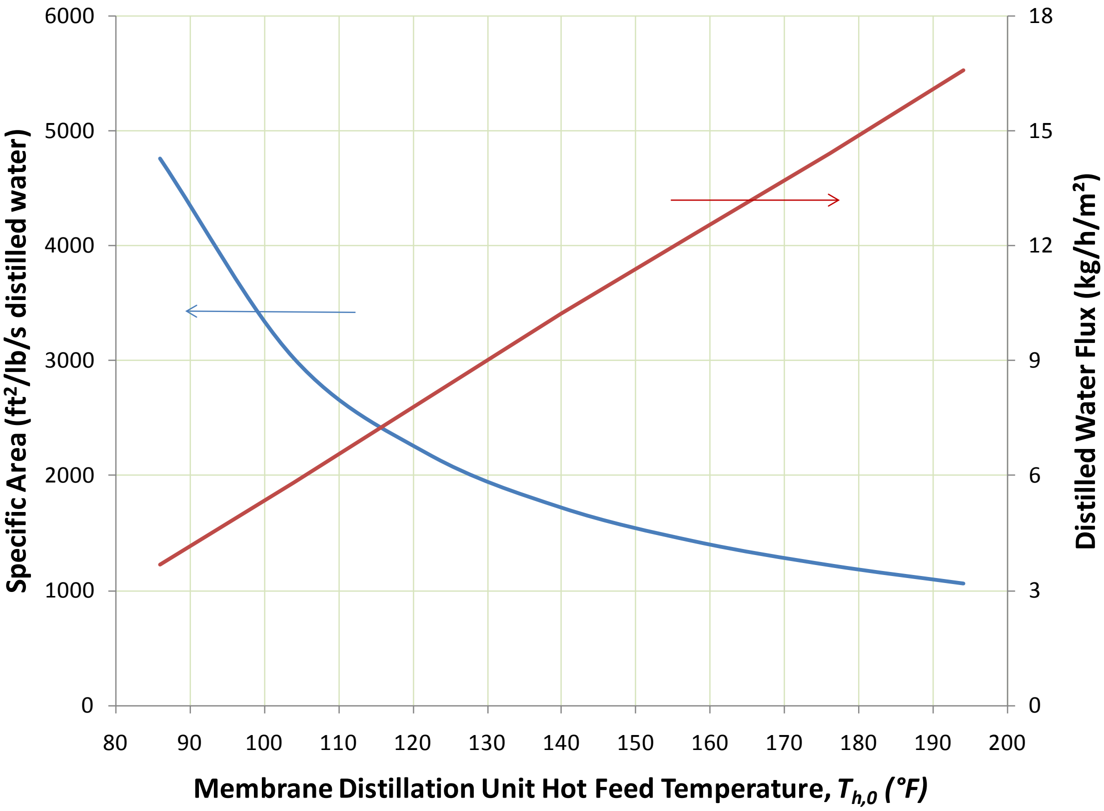

Figure 12: Impact of Water Inlet Temperature on Area Requirement

Specific surface area is a useful parameter because it scales linearly with capacity. A unit designed to provide 1 lb/s at 194 °F and the other parameters of Figure 12 would require 1061 ft^2. The same unit designed for 10 lb/s would require 10,610 ft^2.

Figure 13 contains a schematic showing the inputs and some results from a model run with a high hot feed temperature, $T_{h,0}$. This hot feed temperature is 194 °F. The cold inlet temperature, $T_{c,n}$, is 60 °F and approach temperatures are equal at 10 °F. The resulting distillate rate is 0.101 lb/lb hot feed. The cooling feed rate needed to provide the specified approaches is 0.908 lb cold feed/ lb hot feed. The area needed for this design is 1061 ft^2/lb hot feed.

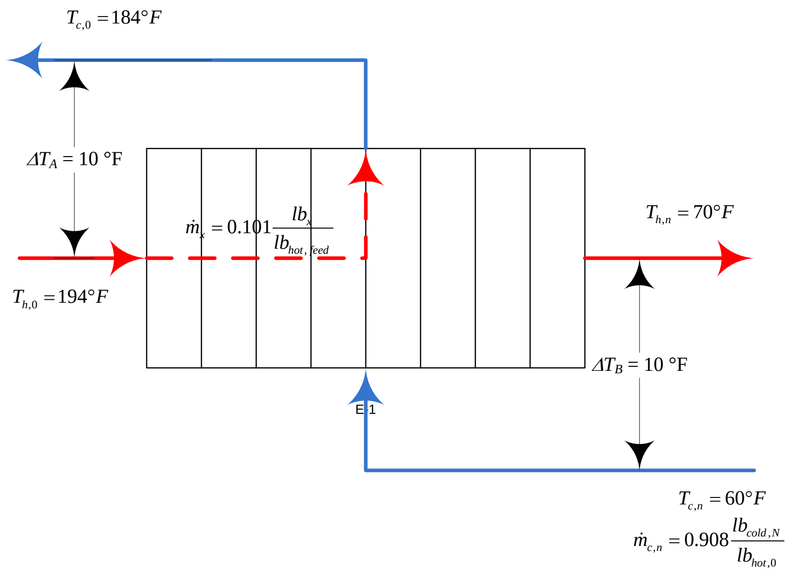

Figure 13: Schematic Summary for High Temperature Feed Example

Figure 14 contains a schematic showing the inputs and some results from a model run with a low hot feed temperature, $T_{h,0}$. This hot feed temperature is 90 °F The cold inlet temperature, $T_{c,n}$, is 60 °F and approach temperatures are equal at 10 °F. The resulting distillate rate is 0.00903 lb/lb hot feed, a factor of 10 lower than for the previous example. The cooling feed rate needed to provide the specified approaches is 0.995 lb cold feed/ lb hot feed. The area needed for this design is 4376, ft²/lb hot feed, four times the area needed for the hot feed example.

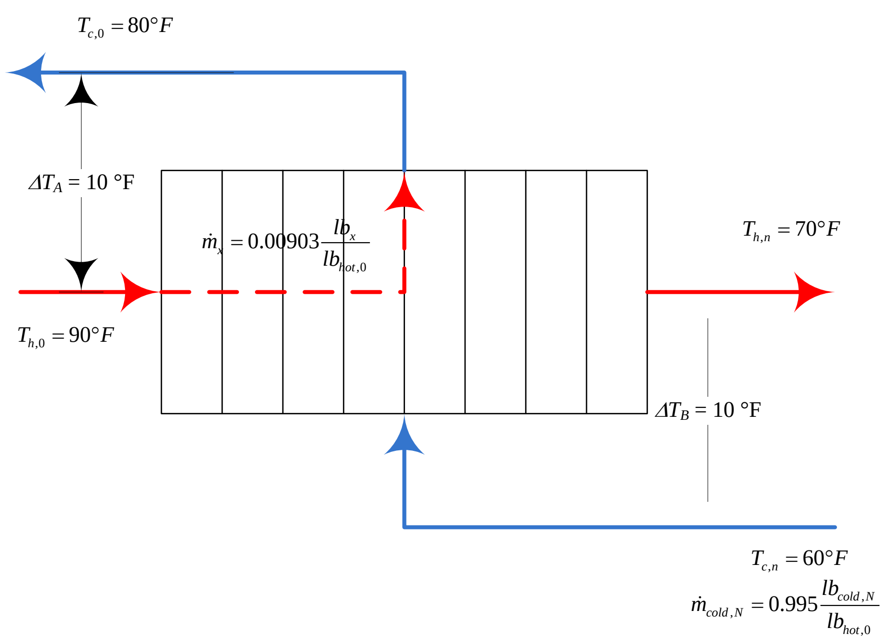

Figure 14: Schematic Summary for Low Temperature Feed Example

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3 POWER PLANT MODEL

This chapter describes the power plant scheme used in this study. Section 3.1 describes the thermodynamic cycle used to convert power. This same section also describes the process schemes used to implement that cycle. Section 3.2 lists the locations within that scheme where waste heat might be available to use in water recovery processes.

3.1 Power Cycle Description

The initial scope of work for this study specified that the power plant to be studied would be a Rankine (steam) cycle system. Even with this specificity, other options exist. The plant can be coal fired, hydrogen fueled or nuclear power. The plant can be critical or subcritical depending on maximum operation temperature. The plant design can incorporate CO_2 removal or exhaust products of combustion directly to the environment. Finally, power plant designers use a wide variety of streams of steam bled off the main turbine process to preheat boiler feed water.

This study uses Case 11 from the DOE/NETL report *Cost and Performance Baseline for Fossil Energy Plants* as the basis for its power plant model. Case 11 provides a design for a supercritical cycle without CO_2 removal with a specific and highly optimized steam bleed heater scheme. Most modern power plant designs use supercritical steam in order to take advantage of the increased efficiency associated with the higher temperatures. The decision to exclude CO_2 removal was predicated mostly by this study's scope of work. Since it involves only the steam system, stack gas processing is irrelevant to this study. Selection of a design with a straightforward stack design simplifies bookkeeping. A real value for Case 11 is the multiple optimized steam bleed streams. This report relies on these streams to assess the impact of diverting some of these streams off for use in water recovery schemes.

Table 2 provides a list of Case 11 parameters significant to this study. It provides two efficiency statistics. The Steam cycle efficiency listed in the above table deals with only the thermodynamics of the steam cycle. The Overall Plant efficiency incorporates a thorough accounting of all energy inputs and outputs. For example, the steam cycle efficiency uses the heat input from coal combustion to the power cycle while the overall plant efficiency includes all of the energy released when coal is burned, including all the energy that escapes from the stack and all the energy invested in sulfur removal. The overall plant efficiency provides a comprehensive measure of total plant performance. However, the steam cycle efficiency provides a much more straightforward means of tracking changes to the process. While steam cycle efficiency omits any issues external to the power conversion cycle, it does address all the issues important to this study. Considering ease of use and relevance to this study, we will use steam cycle efficiency as the basis for comparison for this report.

Table 2: Power Plant Selected Parameters

Parameter	Value	Units
Maximum Process Temperature	1100	°F
Maximum Process Pressure	3514.7	psia
Maximum Steam Flow Rate (hourly basis)	3,665,000	lb/h
Maximum Steam Flow Rate (seconds basis)	1018	lb/s
Steam Cycle Efficiency	46.9%	
Heat Input to Steam Cycle	1,237	MW
Turbine Output	580	MW
Overall Plant Efficiency	39.1%	
Coal Energy Released (HHV)	1,407	MW
Net Electricity to bus bar	550	MW
Net Water Consumption	755	lb/s
Flue Gas Desulfurization	75.7	lb/s
Boiler Feed Water (1% of steam flow)	10.1	lb/s
Cooling Tower	669	lb/s

Figure 15 contains plots of the saturated vapor and saturated liquid lines of the selected steam cycle. Water exists in liquid phase to the left of the saturated liquid curve and in vapor phase to the right of the saturated vapor curve. This leaves the vapor liquid envelope (VLE) between the two lines. In the VLE liquid and gas coexist. The wide region above the critical point is the supercritical region. The closed polygon plotted in red in the figure represents the power cycle. The polygon's area is proportional to the total work extracted from the plant's turbines. Turbine output changes dramatically when maximum pressure changes. Hence, the higher pressures of supercritical designs result in higher power return. Power return changes even more dramatically with changing minimum pressure. The light green shaded area represents the difference between a cycle with turbines discharging at 1 psia (101 7 °F) and turbines discharging at 14.7 psia (212 °F).

Condensation occurs by its nature in the two phase region. In this region, pressure dictates a unique temperature. For example, condensation at a constant temperature of 101.7 °F equates to condensation at a constant pressure of 1 psia.

Figure 16 contains a sketch of the process showing the major process paths through boilers, turbines and heaters as well as secondary steam bleed streams. Figure 16 is a process sketch that illustrates a practical implementation of the thermodynamic cycle shown in Figure 15. This process sketch shows major power production streams as well as secondary bleed heater streams. The following describes the complete cycle:

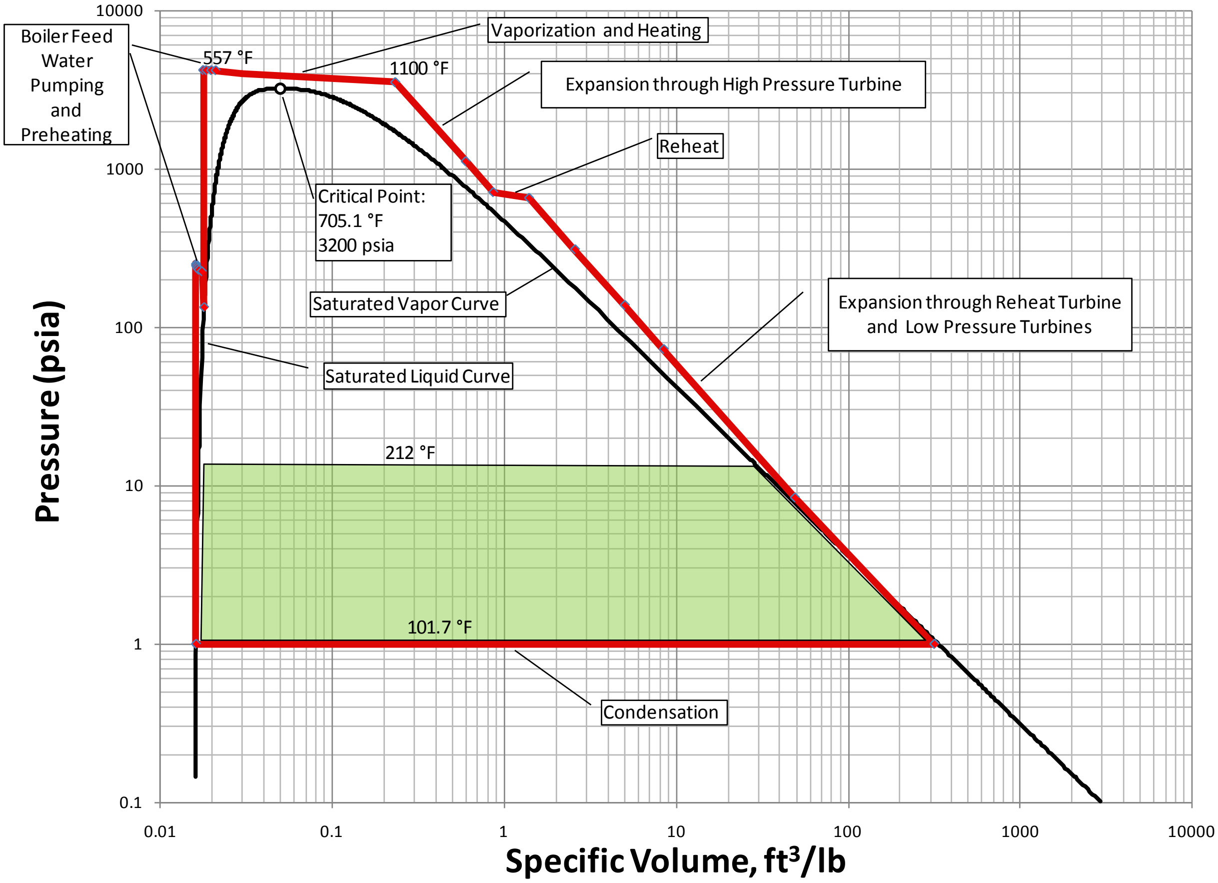


Figure 15: The Rankine Steam Cycle as implemented

1. The process begins with hot high pressure supercritical water exiting the boiler for feed to the high pressure (HP) turbines. The HP turbines generate approximately 183 MW, 31.6% of the plant's power.
2. On exiting the HP turbines, the steam goes back into the boiler to gain additional energy. This reheater adds 223 MW to the steam.
3. Upon exiting the reheater, the steam enters the reheat turbines, which generate 167 MW, 28.7% of the plant's power.
4. The now relatively low pressure steam goes directly from the reheat turbines into the low pressure (LP) turbines to generate the remaining 230 MW, 39.7% of plant output. Very low pressure steam exits the LP turbine at 101.7 °F and 1 psia (14 psi vacuum).
5. LP turbine exhaust proceeds to the condenser which removes sufficient latent heat to condense the steam to liquid phase (approximately 589 MW).

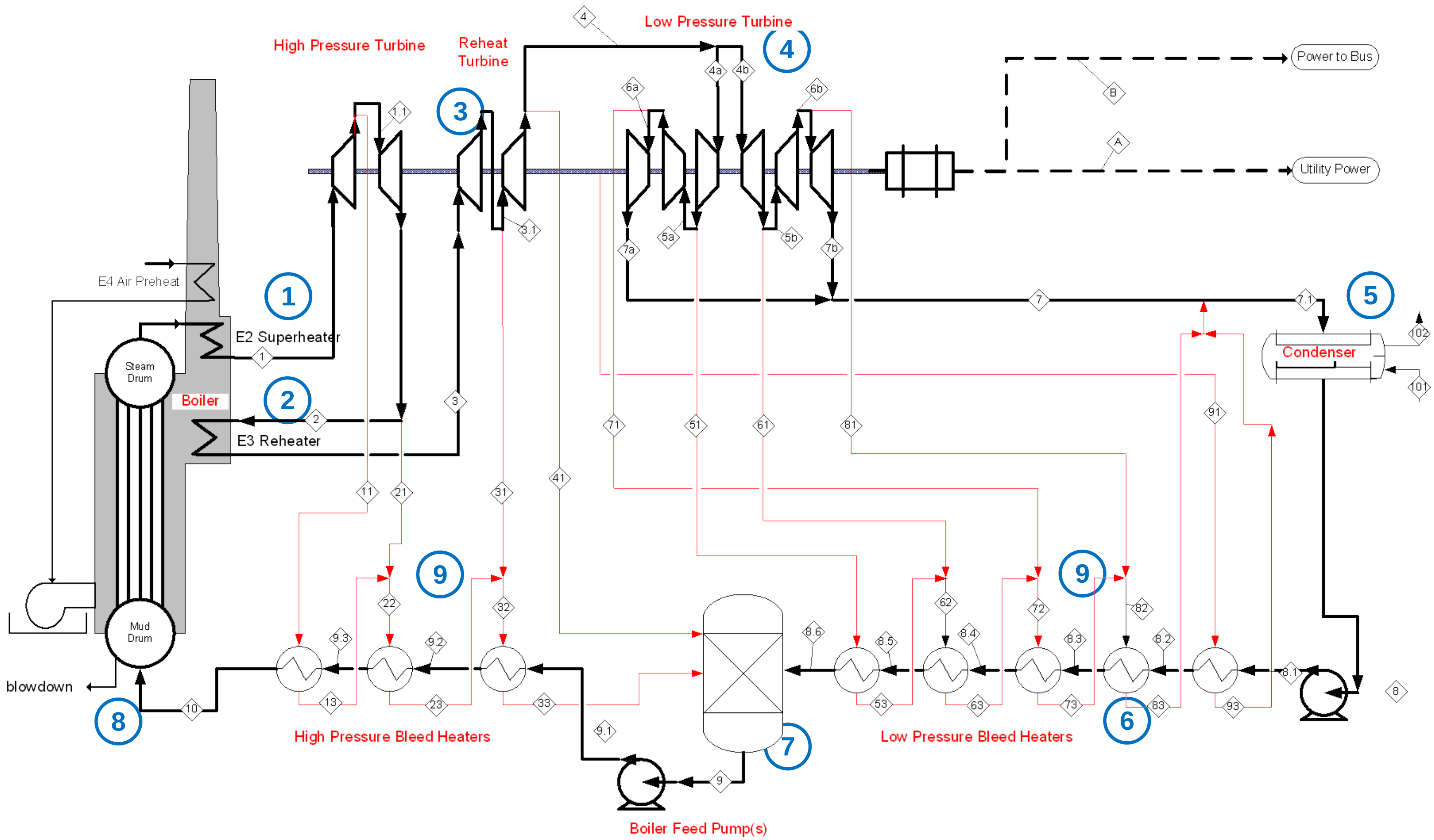


Figure 16: Process Sketch for the Case 11 Power Plant. Numbers refer to the description of the cycles in the text.

6. From the condenser, a series of pumps and heaters progressively increases the temperature and pressure of the hot water, preparing it for reinjection into the boiler. The heating medium is high energy steam bled of the turbine train from strategic locations.
7. Midway through the heating process, a de-aerator removes foreign gases, mostly air, which might have infiltrated into the process stream during the vacuum phase.
8. The boiler accepts the hot high pressure water and heats it further, adding 1,012 MW of energy.
9. The bleed streams preheat boiler feed water by a combination of heat exchangers and direct contact. Bleed rates are typically on the order of 5% to 10% of the major streams' flow rates.

Sizing the bleed streams is an art. A pound of steam bled off the turbine process stream is a pound of steam not generating power. However, that same pound of steam reduces the load on the boiler by preheating boiler feed water. The power plant designers' art is to hit that perfect point where bleed streams reduce boiler load more than they reduce turbine output.

3.2 Water Recovery Schemes Used in This Study

The primary focus of this study is determining potential water recovery via membrane distillation using heat from the three locations identified in Figure 17. The three are:

1. Water recovery using the waste heat contained in blow down from the power plant's steam boiler (boiler blow down).
2. Water recovery using redirected flow from one of the eight steam streams bled off of the turbine train. This is redirected flow is not necessarily a waste stream.
3. Water recovery using waste heat removed during low pressure steam condensation by the cooling water system.

The boiler blow down is attractive for two reasons. The stream is relatively high energy with a temperature of 512 °F at a pressure of 4185 psia. In addition, the stream is relatively clean, having a total dissolved solids (TDS) level of 600 to 650 ppm. In this report, the units of ppm are lb of dissolved solids per million lb of water. A negative for this option is the relatively low flow rate. Only around 1% of boiler steam production on a mass basis, or about 10 lb/s (73 gpm). Since we can only hope to recover a fraction of this mass flow, the value of water recovery using blow down is limited to the reduction of boiler feed water makeup costs.

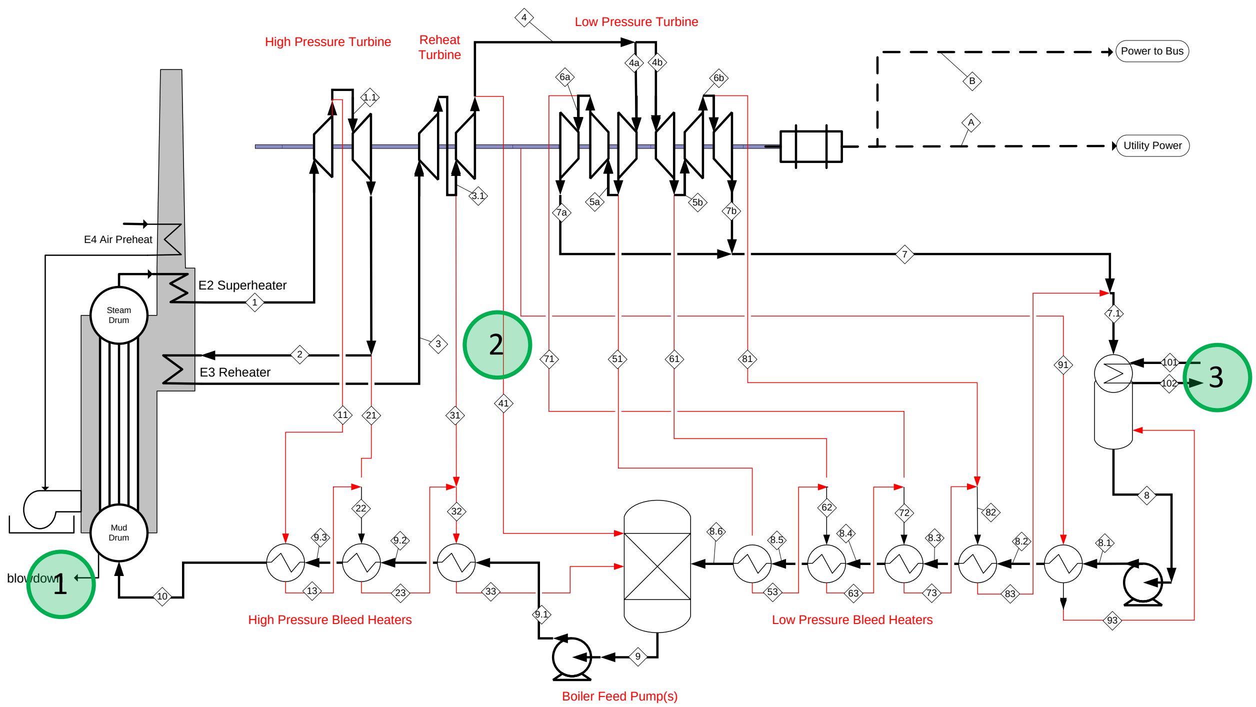


Figure 17: Proposed Schemes to Recover Water Using Waste Heat

There are two problems involved with using redirected flow from steam streams for water recovery. The first, and most important is that bled steam is not a waste stream. Any energy removed from the steam for water recovery directly impacts the power produced by the plant. The second is that bleed streams' rate, location and injection point are carefully optimized by plant designers. All of these parameters affect not only the thermodynamic cycle, but also the design of virtually all the equipment in the plant.

On the other hand, bleed streams have a lot of energy, and flow rates can be high. The volume of recovered water can be significant. In all, there are eight bleed streams, varying in pressure from 3.5 psia (vacuum) to 1115 psia. Rather than attempt an encyclopedic analysis of the impact of redirection of each bled stream, including repetition of the optimization process, this study looks at a limited number of these streams and concentrates on the highest pressure one, stream 11.

The cooling water system removes energy given up by the steam when it condenses. This is a huge amount of energy, approximately 47% of all heat input by the boilers. Unfortunately steam condenses to water at the very low value of 101.7 °F (1 psia), making approaches between the condenser and ambient cooled heat exchangers difficult. As the following chapter describes, difficulty appears tailor made for a membrane distillation scheme.

4 WATER RECOVERY USING BOILER BLOW DOWN

This section evaluates the feasibility of using boiler blow down (Location 1 in Figure 17) as the heat and feed source for membrane distillation. Whenever a system vaporizes a fraction of its throughput, impurities are left behind. The concentration of the impurities in the remaining flow increases. Process designers accommodate this by providing blow down for systems. Blow down removes high impurity fluid by dumping it to waste and replacing it with relatively more pure makeup. This blow down is a waste stream. Blow down can exist for different parts of the system, including boiler blow down and cooling tower blow down.

The power plants main boiler blow down is a good example. In this case, boiler blow down is a high energy waste stream – one that can be used for water recovery. The question is how to use it. Figure 18 contains two sketches showing on the left a typical power plant's normal process for handling boiler blow down and on the right a boiler blow down process designed for water recovery using the inherent properties of that stream.

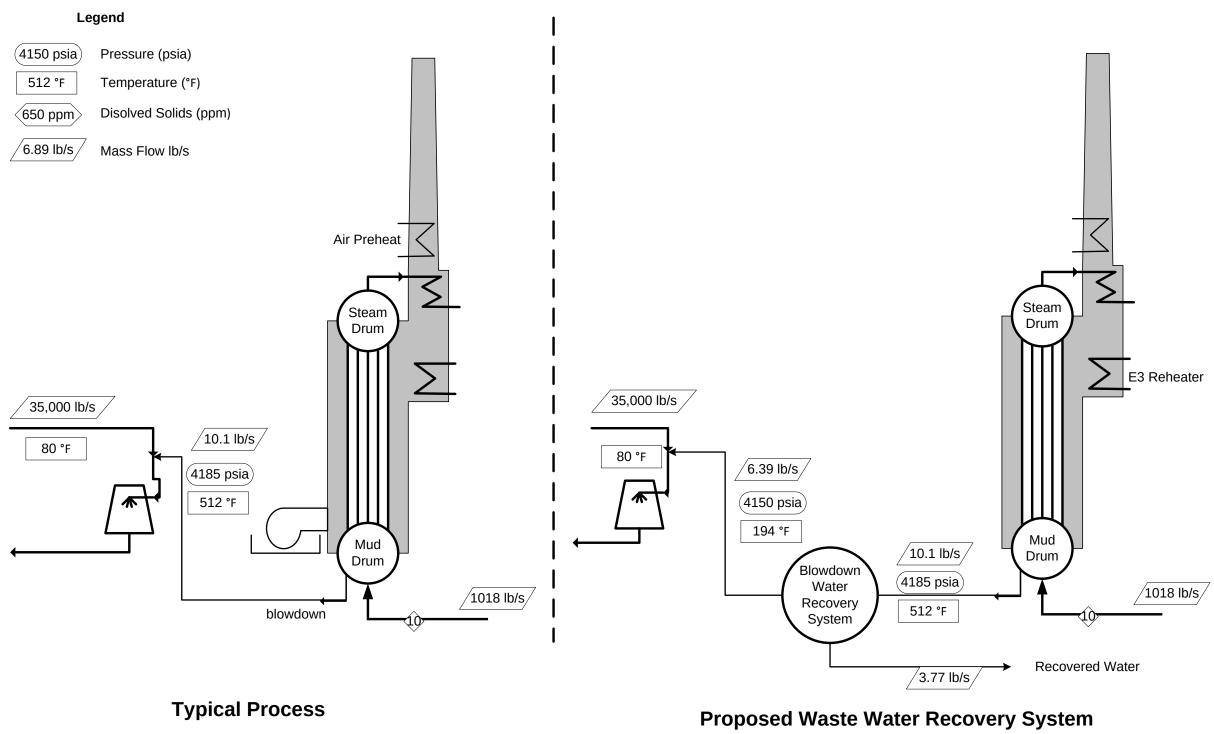


Figure 18: Boiler Blow Down Recovery Process – Position in Main Process

A typical boiler blow down process is shown on the left of the figure. The stream exits the boiler at 512 °F and 4185 psia. This high energy stream is dumped into the cooling water system, more to put it somewhere than to mitigate that system's own makeup requirements. The high mass rate of the cooling water system (35,000 lb/s) tends to swamp the tiny boiler blow down rate

(10.1 lb/s). As a consequence, the high energy of the boiler blow down has as little impact on the cooling water system's energy balance as its mass rate had on the cooling water systems mass balance. Nonetheless, using the energy for water recovery does go in the right direction.

This small mass flow rate (10.1 lb/s or 1% of steam production) means that despite the high specific energy content of boiler blow down the amount of water recovered won't be that high. For example, the process proposed here can only recapture 37% of the original input. With a boiler blow down rate of 10 lb/s, the water recovery rate will be only 3.7 lb/s. On the one hand, this amount is peanuts when compared to the cooling water circulation rate of 35,000 lb/s. Nonetheless, this 3.7 lb/s does mean a significant reduction in the size and cost of a boiler makeup system.

Section 4.1 of this chapter expands on boiler blow down and its calculation. It also takes the opportunity to introduce issues regarding boiler blow down important to Chapter 6. Sections 4.2 and 4.3 describe the proposed process and its equipment respectively. Finally, Section 4.3 discusses the impact of reducing membrane filter feed temperature on the process.

4.1 An Introduction to Blow Down

For water systems, impurities are dissolved salts like NaCl, CaCl, $CaCO_3$, and $MgCO_3$. The actual list is much longer. These impurities are measured as total dissolved solids (TDS) and in this study will be reported in ppm (lb_{salt}/million lb_{water}).

In a power plant, two systems involve vaporized water. These are the steam boiler and the cooling water tower. Both are of interest to this study. Both are represented in Figure 19. In addition, the subject of the study, the membrane filtration systems involve vaporized water. Hence membrane distillation design includes blow down also and is thus represented in the figure.

The steam boiler flow rate for this study is 1018 lb/s (See Table 2). A reasonable allowable TDS level for a boiler is 600 ppm.[17] Steam boilers take makeup typically from a demineralization plant which can produce water with TDS levels around 6 ppm. The major parameter driving blow down is the cycle of concentration, CoC, defined as

$$CoC = \frac{ppm_{blowdown}}{ppm_{makeup}} \tag{14}$$

Since boilers will blow down from their flowing stream, blow down will have a TDS of 600 ppm. Hence, CoC for the boiler is 600/6 = 100. From the figure, appropriate blow down rate is 10 lb/s.

Cooling water is more interesting. Cooling water manufacturers recommend TDS levels of 5,000 ppm or less [18]. Typical makeup water will be in the range of 50 ppm, resulting in a CoC of 100 or less. From the curve, blow down will be about 350 lb/s, slightly more than half the total cooling tower water makeup requirement. Based on the Table 2 total makeup of 669 lb/s for a cooling tower, 350 lb/s replace blow down and 319 lb replace evaporation

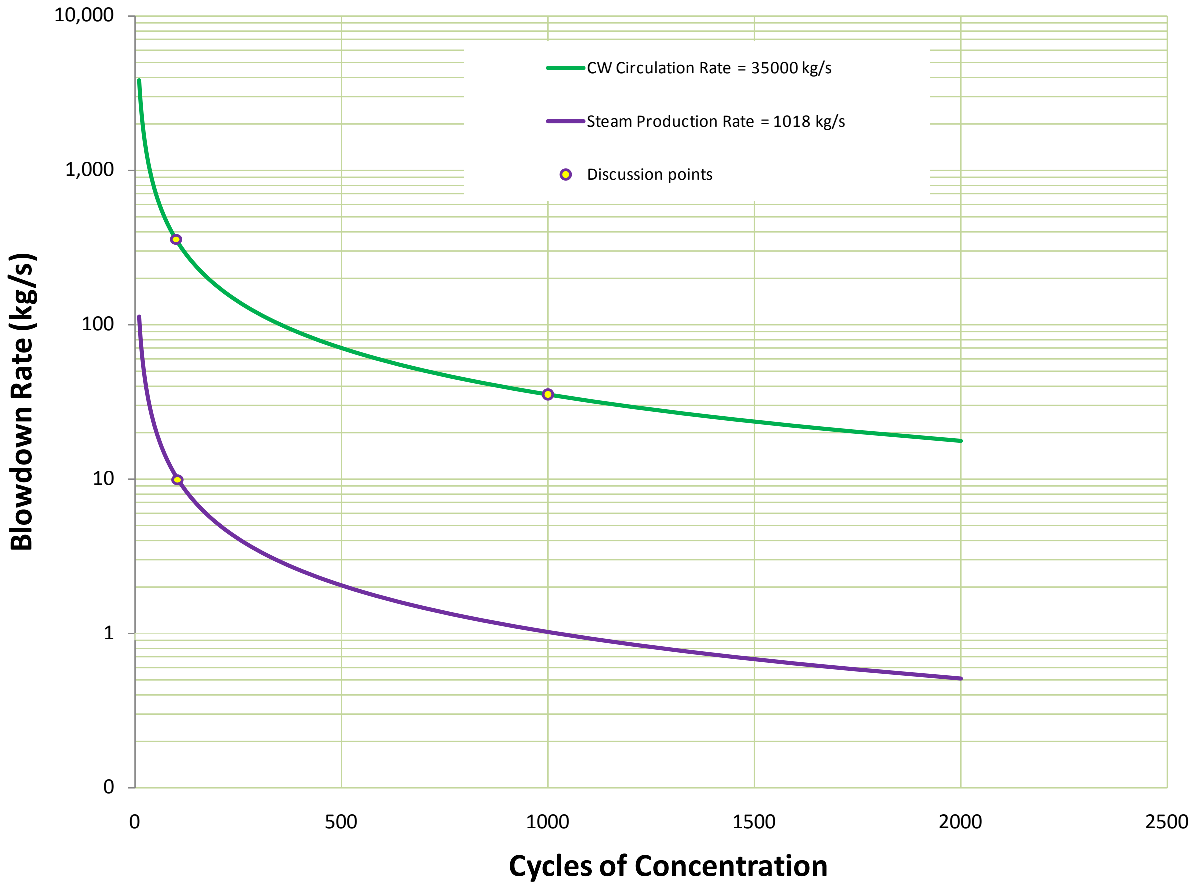


Figure 19: Boiler Blowdown Dependence on Cycles of Concentration

Consider, however, what happens if membrane filtration systems are used to make up water. A membrane distillation unit is effectively a low temperature distillation system. Distilled water is very pure, having TDS on the order of 5 ppm. The CoC goes up to 1000. From Figure 19, blow down requirements drop to 35 lb/s, a drop of 90% from 350 lb/s. Total water makeup requirements for this configuration is 350 lb/s

4.2 Process Description

Figure 20 contains a process sketch of the blow down recovery system. The proposed process addresses some basic problems:

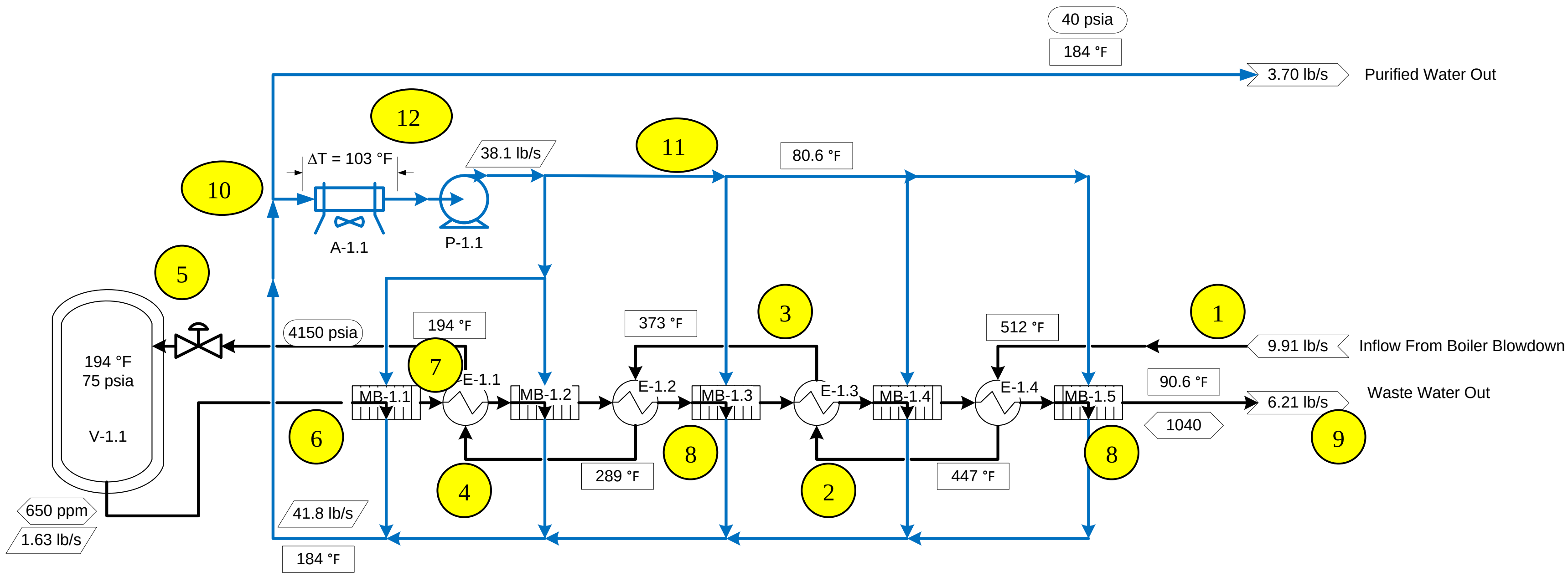
1. The boiler blow down (process feed) is too hot to enter directly into a membrane filtration system. A membrane distillation unit works only if the hot stream is below its

boiling point (See Section 2.1). The stream needs pre-cooling from 512 ºF to 194 ºF, a temperature reasonably below boiling.
2. The effluent from a single membrane distillation unit remains of sufficient quality that additional water could be recovered. However, the membrane distillation unit effluent is too cold for further extraction. It needs reheating.

It seems natural to use these two problems to mutual benefit. The feed reheats membrane effluent for additional extraction while simultaneously cooling itself. In fact, this process needs repeating multiple times before the stream reaches a temperature low enough to support membrane filtration.

The following numbered paragraphs describe a water recovery from boiler blow down process that uses membrane distillation. The system is show graphically in Figure 20. The circled numbers there correspond to the numbered paragraphs below.

1. Mud drum effluent (feed) enters the new process and immediately enters the heat exchanger E-1.4 where it preheats the water entering MB-1.5.
2. This heat exchanger cools the boiler blow down effluent from 512 °F to 447 °F. This stream continues directly to E-1.3 where it gets cooled to 373 °F.
3. Flow continues through E-1.2 (289 °F exit temperature).
4. Flow finally continues through E-1.1, where the exit temperature is 194 °F, low enough to enter a membrane distillation unit. At this point, the stream is cool enough, but remains at high pressure.
5. The flow undergoes an isentropic flash to 70 psia into surge drum V-1.1. The flash reduces stream temperature from 194 °F to 191 °F. At 70 psia and 191 °F, the water is ready to enter membrane distillation units.
6. Returning flow enters the first membrane distillation unit MB-1.1 which extracts 0.901 lb/s (8.9% of inflow) as pure water. The process cools the stream to 81 °F. The MD distillate product enters the closed cooling water loop.
7. MB-1.1 effluent enters the cold water side of E-1.1 where it is reheated to 194 °F. Removal of pure water increases TDS form 650 ppm at the inlet to 714 ppm at the exit.
8. This sequence of membrane distillation unit followed by reheater repeats three times with one additional membrane distillation unit without reheat (5 MD versus 4 reheaters).
9. Final effluent exits the system bound for the cooling water system with a TDS level of 1040 ppm, still substantially below the blow down limit of 5000 ppm for the cooling water system. Effluent has a temperature of 91 °F and a pressure of 20 psia. Total extracted flow is 3.77 lb/s, 37.3% of feed.
10. The 3.77 lb/s of pure water extracted by membrane distillation units goes into the closed cooling water system from which it is removed prior to cooling and recycle. The removed water is now available without further processing.
11. This dedicated cooling water system circulates enough cold water to support all five membrane distillation units.
12. It uses a closed dry air exchanger to avoid contamination. Without the advantage of evaporative cooling, minimum cooling water is, on the average, 80 °F.



Legend

4150 psia	Pressure (psia)
512 °F	Temperature (°F)
650 ppm	Disolved Solids (ppm)
6.89 lb/s	Mass Flow lb/s

Heat Exchanger Summary

Unit	Btu/s	Watts	Area ft²
E-1.1	933.4	983,900	400
E-1.2	850.2	896,200	150
E-1.3	774.3	816,200	92
E-1.4	705.2	743,400	67
A-1.1	3941	4,154,000	13000

Pump Summary

Unit	P-1.1	
ΔP	5.00	psi
Flow	38.103	lb/s
Power	1.002	hp

Membrane Filter Summary

Unit	Inflow	Purification	D. Solids	Area
	lb/s	lb/s	ppm (mass)	ft²
MB-1.1	10.100	0.901	714	803
MB-1.2	9.199	0.821	784	731
MB-1.3	8.378	0.748	860	666
MB-1.4	7.630	0.681	945	606
MB-1.5	6.949	0.620	1040	552

Figure 20: Process Sketch -- Membrane Distillation units with Boiler Blow Down

4.3 Process Equipment

Figure 20 contains equipment lists with sizing parameters. In general, equipment for this proposed system is small. For example, the cooling water pumps are only 1.00 hp. The membrane distillation units range from 550 to 800 ft^2. The shell and tube heat exchangers range from 70 ft^2 to 400 ft^2. All of these sizes are on the low end of commercial equipment offerings.

The exception is the aerial cooler, A-1.1. This application usually would take a cooling tower. However, this system must avoid the potential for contamination associated with an open cooling tower. The alternative is a closed aerial cooler. Without the evaporative cooling available with a cooling tower, approach temperatures are low. These low approaches drive the area requirement up to 13000 square feet. This is slightly larger than a single commercial fin fan cooler, so two would be required [21]

4.4 Lower Temperature Operation

As discussed in Chapter 2, membrane distillation unit must operate below the boiling point of water. For the purposes of this discussion, assume this means 212 ºF. While the possibility is low, the probability exists that operation close to 212 ºF might be detrimental to the life of the membrane. The answer is probably an economic one. Increased temperature might accelerate aging, shortening useful life. Elevated temperature might reduce the strength of the membrane to the point that creep begins. Elevated temperature might reduce the hydrophobic nature of the material. It might also accelerate chemical decomposition or other reactions. All of these possibilities are low probability and theoretical, requiring extensive research into and about appropriate membrane materials – activities well beyond the scope of this study. What would happen to the above process if the operating temperature was lower than the 194 ºF used in the original description of this process?

The answer is straightforward. The heating duty of each heat exchanger would drop. Since the available energy in the blow down remains constant, the number of heat exchangers must go up. The number of membrane distillation units would increase with heat exchanger count. There would always be one more distillation unit than there are heat exchangers. However, with reduced maximum temperature the rate of water purification per membrane distillation unit would drop.

Figure 21 summarizes the integrated impact of the above discussion on the process. Surprisingly, rate of pure water extraction stays relatively constant at between 0.29 and 0.38 $lb_x/lb_{h,\cdot c}$. The lower rate corresponds to lower membrane inflow temperature (122 ºF). However, the number of heat exchangers required goes up dramatically from the four needed for an inflow temperature of 194 ºF (see above) to fifteen (15) for the 122 ºF inlet case. Membrane distillation unit count would go from 5 for the base case to 16 for the 122 ºF case.

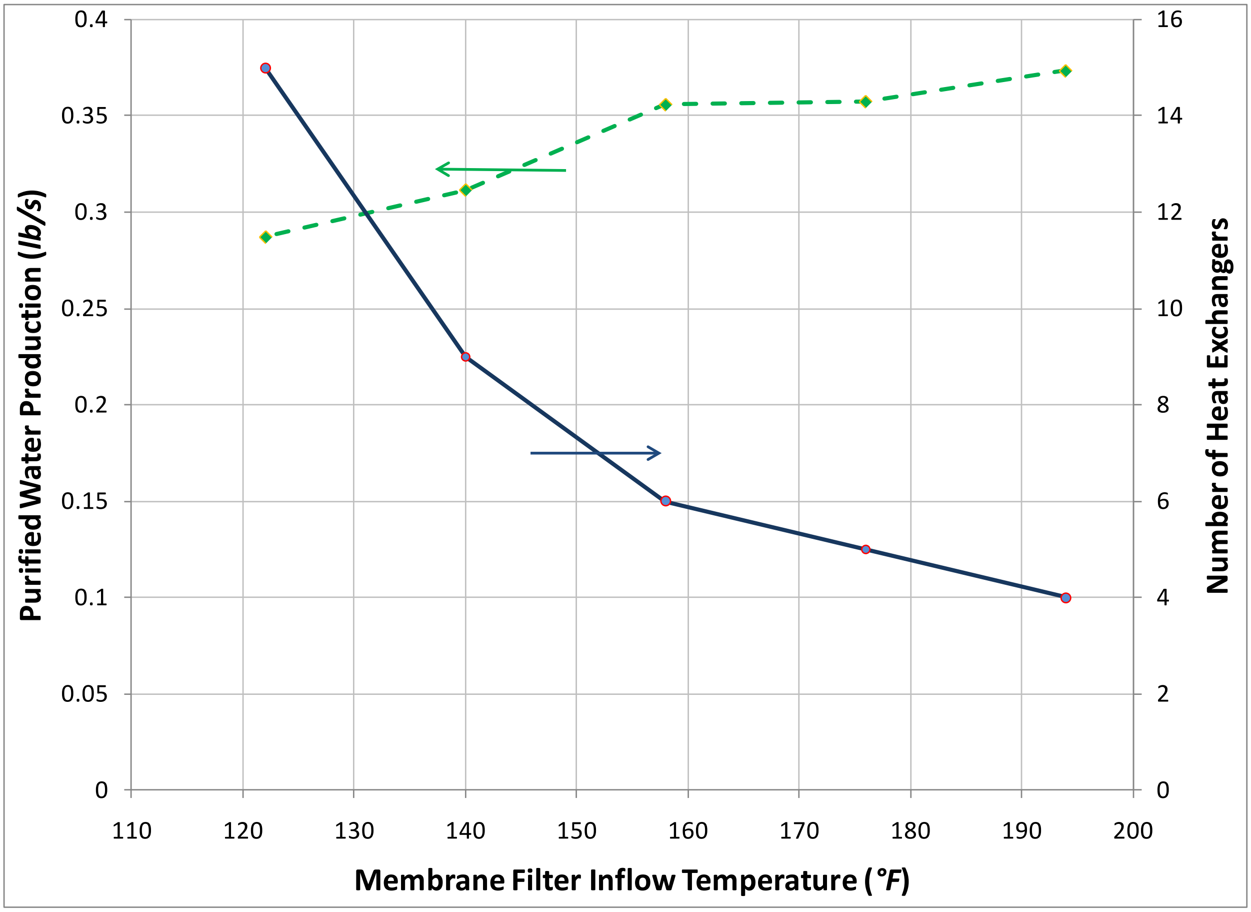


Figure 21: Blow Down Water Recovery -- Effect of Reduced Inlet Temperature on Process

For reference, effluent water quality would improve slightly. Total dissolved solids in the inflow remains constant. As the amount of removed water goes down, the quantity of effluent available to dilute the dissolved solids increases. The effluent TDS is 1040 for the base case, and is 910 for the 122 ºF case.

Fortunately, total capital cost would not increase linearly with equipment count as the equipment size would drop with duty or flow rate. Figure 22 provides a rough estimate of the cost increases using the estimators rule of thumb 7/10 rule (see Section 7.2). At the lowest temperature case, capital cost would go up by a factor of 2.52. When one considers that purified water production drops to 2.90 lb/s for the 122 ºF case, an increase by 250% for the capital cost of the system will be significant economically.

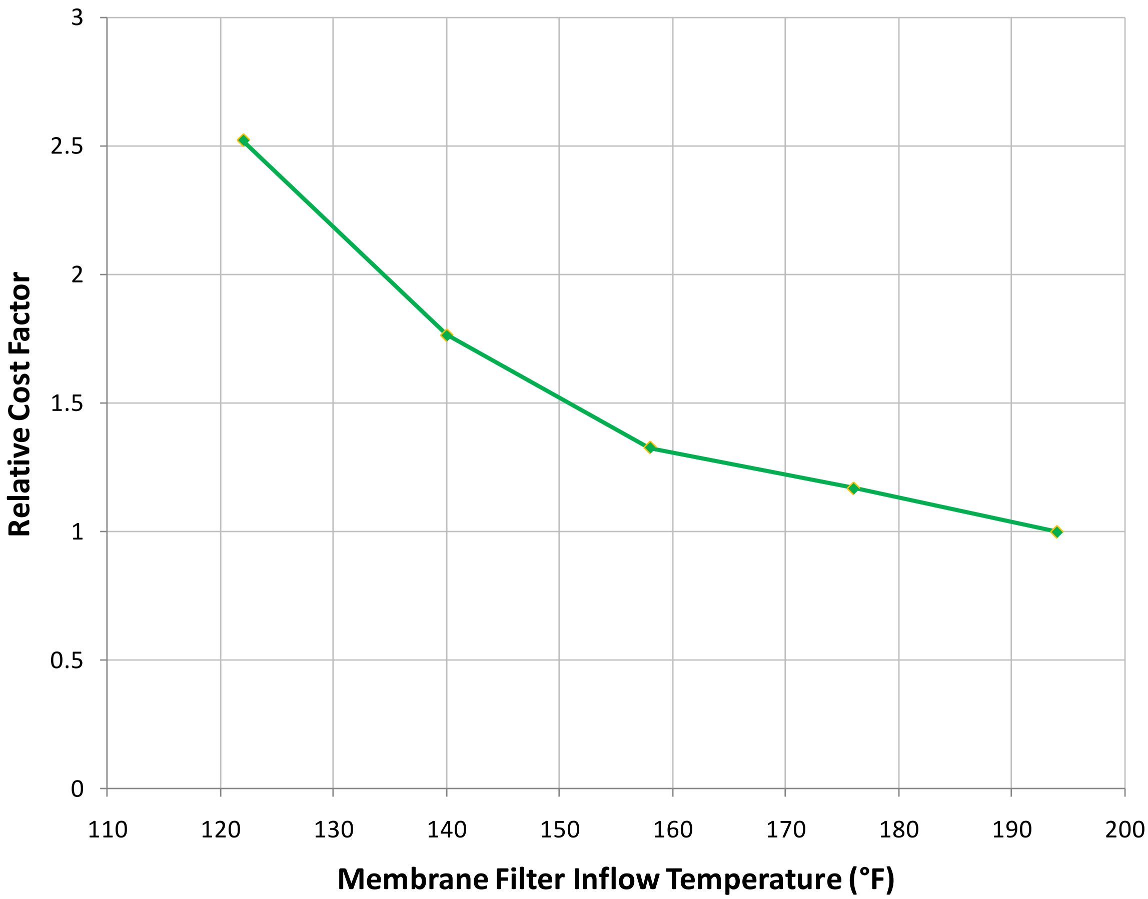


Figure 22: Blow Down Water Recovery -- Effect of Reduced Inlet Temperature on Cost

5 WATER RECOVERY USING STEAM DIVERTED FROM BLEED STREAMS

This section evaluates the feasibility of using steam bleed (Location 2 in Figure 17) as the heat source for membrane distillation. In this scheme, steam bled off the turbine train from one of 8 points and currently dedicated to pre-heating boiler feed water is redirected from the bleed heater to a water recovery process. While using energy from a steam bleed is not waste heat as the energy is otherwise be used (see Section 3.2), it was worth examining in order to understand the potential water recovery from membrane distillation and at the same time better understand the negative impact to the power plant of losing the steam bleed energy source.

Table 3 contains a list of the eight bleed streams available for redirection. Refer to Figure 16 for the locations of the stream numbers

Table 3: List of Available Bleed Streams

Stream Number	Mass Rate (lb/s)	Temperature (°F)	Pressure (psia)
11	74.7	773	1115.5
21	91.2	663	710.8
31	46.9	898	310.1
51	55.6	566	72.7
61	27.3	333	19.2
71	24.4	211	8.4
81	25.7	148	3.5

As said in Section 3.2, the number, mass rate, temperature and pressure of these streams was subject to variation under the intense optimization of design engineers trying to maximize plant output. Redirecting any of these streams will upset that optimized structure. Rebuilding an optimized system after redirection is well outside the scope of this study. Instead, the approach taken was to select one stream and assess the impact and advantages associated with redirection.

Figure 23 provides a bit more detail. The stream chosen was stream 11, the highest pressure and highest temperature bleed-off steam. Some or all of the 74.7 lb/s that makes up this stream will redirected from the bleed heater to a waste water recovery process. Unlike the previous chapter where the high energy stream was also the source of water for distillation, the diverted steam here will only heat brackish water. This brackish water will supply the distillate. Having heated the brackish water the steam/water will be so cold and so low pressure that it will go directly back to the condenser rather than attempting to extract any further process advantage from it. The brackish water stream will be once-through. Distilled water from the membrane distillation process will then be available as a utility stream.

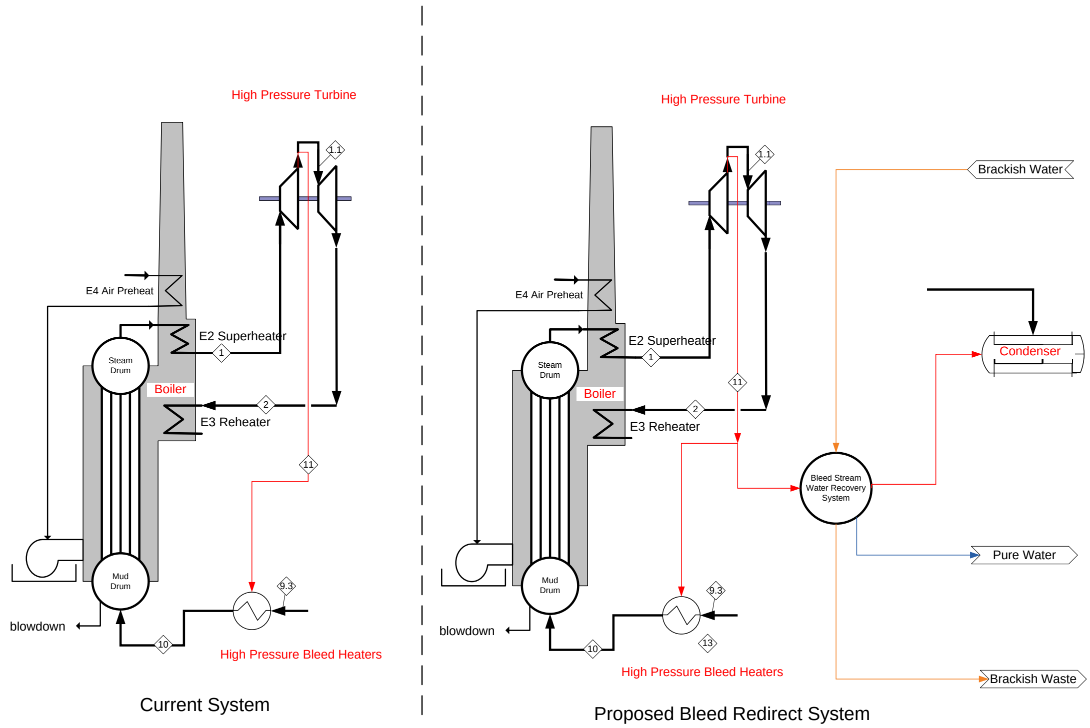


Figure 23: Bleed Stream Water Recovery Process – Position Relative to Main Process

This chapter begins with a discussion of the impacts steam diversion will have on power production (Section 5.1). A process description for the waste water recovery system follows as Section 5.2 with Section 5.3, a description of process equipment, coming last.

5.1 Diversion of Bleed Heat – Impact on Plant Operations

As discussed in Section 3.1, selection of bleed stream location and amount is a designer's art. Rather than pretend to competent levels of that art, this report looks parametrically at the impact on total power production when streams are diverted.

Figure 24 shows the results of this study. The figure plots power versus the rate in lb/s of steam diverted away from the bleed heaters. The figure contains plots for three of the eight bleed steam streams, identified here via the stream numbers used in Figure 16.

- Stream 11 – Bleed from between wheels of the HP Turbine, highest pressure bleed stream.

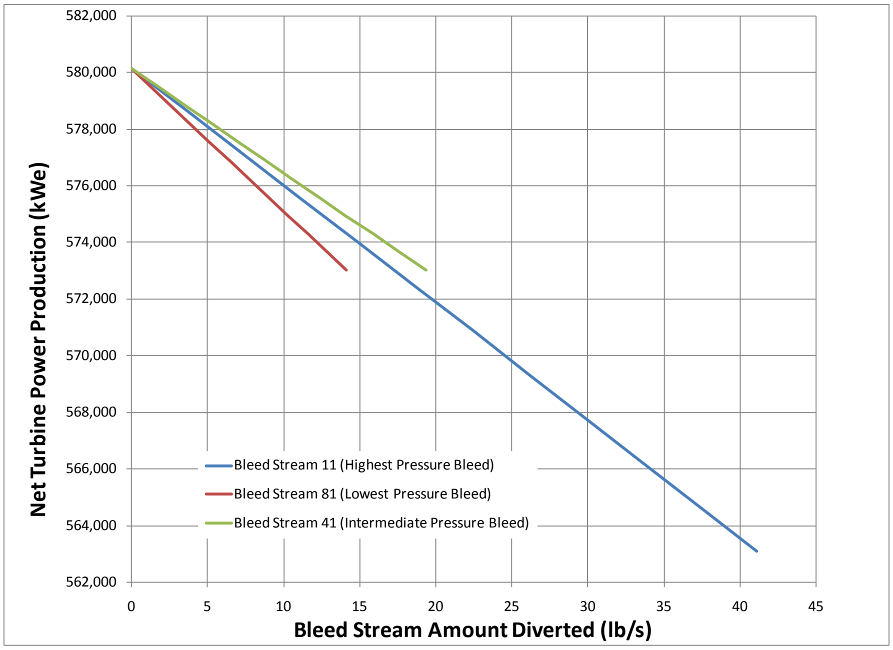


Figure 24: Impact of Bleed sidetracking on Plant output

- Stream 41 – Bleed from the stream discharging from the reheat turbines, moderate pressure stream.
- Stream 81 – Bleed from between the last two wheels of the LP compressor, lowest pressure bleed stream.

In all three cases this study assumes a maximum diverted bleed of approximately 45% of bleed stream rate.

These power loss quantities are significant. For example, the following section will show that to extract approx 10 lb/s of distilled water from a brackish feed (enough to resupply boiler blowdown) one would need 10.28 lb/s of bleed off of highest pressure bleed stream (11). From Figure 24 turbine power production would drop from 580 MW to 576 MW this is about 4000 kW. Using the economics of Section 7.1, this equates to 33.3 million kWh/y. At $0.10/kWh, this diversion would mean a drop in cash flow of $3.3 million dollars per year.

5.2 Process Description

The process described here uses diversion from stream 11. In general, the same process would work for any of the eight streams. Sizes, flow rates and other process parameters would vary depending on the energy available from the chosen stream.

As with blow down in the previous chapter, steam diverted from Stream 11 is too hot for optimal exchange with a single membrane distillation stream. Therefore, like blow down, the process uses an interweaving heat exchanger/membrane sequence. Unlike blow down, the heating stream does not then become the membrane distillation feed stream. Instead, this process takes nontraditional impaired water (herein referred to as brackish water) from an unidentified external source. Table 4 contains the specifications assumed for the brackish water feed.

Table 4: Brackish Water Supply Specification

Description	Value	Units
Temperature	70	°F
Pressure	60	psia
Total dissolved Solids (TDS)	5000	ppm (mass)

For reference, total dissolved solids in the cooling tower tested and coal bed methane produced water examined as a separate part of this project was on the order of 1,000 and 15,000 ppm, respectively.[19] Based on these specifications, cooling water tower blow down would be a reasonable source for brackish water.

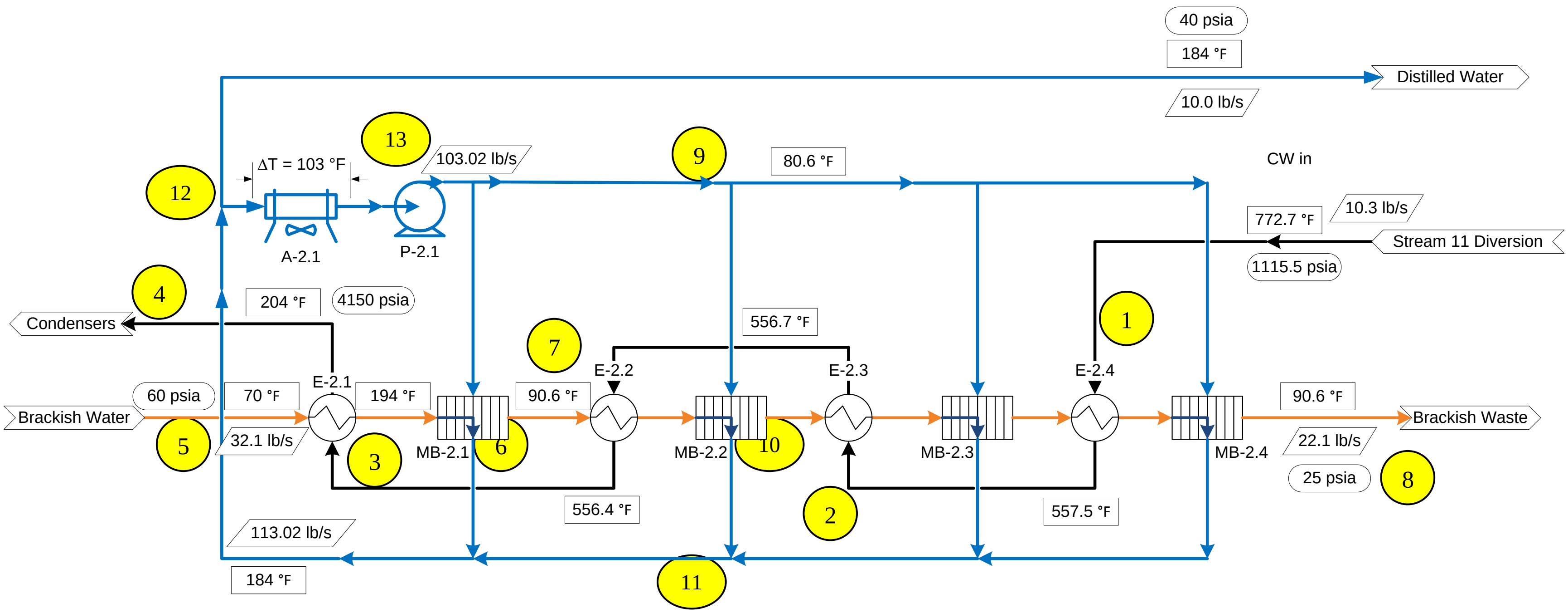
Figure 25 provides a process sketch for a water recovery process using diverted steam. This particular example uses four heater/membrane pairs. As will be discussed later, the design is relatively insensitive to the number of pairs used. This example was sized to retrieve 10 lb/s, enough nominally to make up power boiler blow down. The design can be scaled linearly for other output flows. The intent here was to provide sufficient quantification to assess the impact on the overall plant.

This process uses a closed aerial cooler in lieu of a cooling tower. This decision means the loss of the advantage of evaporative cooling. However it avoids the impurities introduction that are unavoidable with the open evaporative cooling towers.

The following numbered discussion describes the process in some detail. These paragraph numbers correspond to numbered circles in Figure 25.

Diverted Steam

1. Steam diverted from steam bled off the HP turbine's inter-wheel space (Stream 11) enters the process at 772.7 °F and 1115.5 psia. The stream immediately enters E-2.4, where it serves as the heating medium to reheat the brackish water stream. Steam is super heated at this point.
2. Steam exits E-2.4 at 557.5 °F. At this point it has begun to condense some liquid, and has a quality of 0.895. It immediately enters the next heater, E-2.3.
3. The steam continues this process through heat exchangers E-2.3, E-2.2 and E-2.1, losing temperature with each exchanger. By the time it leaves E-2.1, it has completely condensed to mildly hot water (204 °F). Pressure, however, remains high at 4150 psia.
4. This hot water uses its remaining pressure to return to the condensers.



Legend

4150 psia	Pressure (psia)
512 °F	Temperature (°F)
650 ppm	Disolved Solids (ppm)
6.89 lb/s	Mass Flow lb/s

Membrane Filter Summary

Unit	Inflow	Purification	D. Solids	Area
	lb/s	lb/s	ppm	ft2
MB-2.1	32.053	2.861	5490	2546.929
MB-2.2	29.193	2.605	6030	2319.622
MB-2.3	26.587	2.373	6620	2112.601
MB-2.4	24.214	2.161	7270	1924.057

Pump Summary

Unit	P-1.1	
ΔP	5	psi
Flow	103.015	lb/s
Power	2.883	hp

Heat Exchanger Summary

Unit	Duty	Duty	Area
	Btu/s	Watts	ft2
E-2.1	3976	4,191,000	700
E-2.2	3020	3,183,000	230
E-2.3	2751	2,900,000	210
E-2.4	2505.563	2,641,000	160
A-2.1	10653.92	11,230,000	37800

Figure 25: Process Sketch -- Membrane Distillation units with Bleed Stream Diversion

Brackish Water

5. Brackish water at ambient temperature enters the process, going immediately into heater E-2.1 where it is heated to 194 °F in preparation for entry into the first membrane unit, MB-2.1.
6. This membrane unit removes 2.86 lb/s of distillate. The brackish water exits at 90.6 °F.
7. The brackish feed repeats this heater/membrane sequence three more times. Water temperatures cycle between 90.6 °F entering the heater to 194 °F entering the membrane unit.
8. After the last membrane, brackish water exits the system for disposal. Total Dissolved solids levels increase from 5000 ppm entering the first heater to 7270 ppm exiting the system.

Closed Cooling Water System

9. A closed cooling water cycle provides cold side flow for the membrane distillation units. Water enters all four membrane units in parallel. Temperature of this cold stream is 80.6 °F.
10. Within the membrane units, a stream of distilled water merges with the cooling water stream across the membrane from the brackish water.
11. Temperature leaving all four membranes is 184 °F.
12. Cooling water is collected and 10 lb/s is split off as product. Since the probable destination is boiler feed water makeup, no attempt is made to cool or otherwise condition the stream.
13. The remaining water enters the aerial cooler for cooling to 80.6 °F.
14. A centrifugal pump increases pressure to 30 psia for delivery to the membranes again.

Earlier discussion in this section alluded to the relatively insensitivity of this process to the number of heater/membrane pairs deployed. Figure 26 adds detail to that discussion. With 15 pairs, a process would need 9.95 lb/s diverted steam. With 2 heater/membrane pairs, the diverted steam demand goes up to 10.7 lb/s. This is a very small variation when one considers the range of equipment needs.

This insensitivity is less pronounced when one considers brackish water demand. Refer to the right vertical axis in Figure 26. Increasing heater/membrane pairs from 4 to 15 drops brackish water demand from 32.1 lb/s to 13.3 lb/s. The significance of this decrease depends on the cost of brackish water.

Cooling water system needs remain constant for any heater/membrane pair count. Cooling water mass rate is a function of distillate delivery rate. Since for this exercise is delivery rate is constant cooling water rate is also constant.

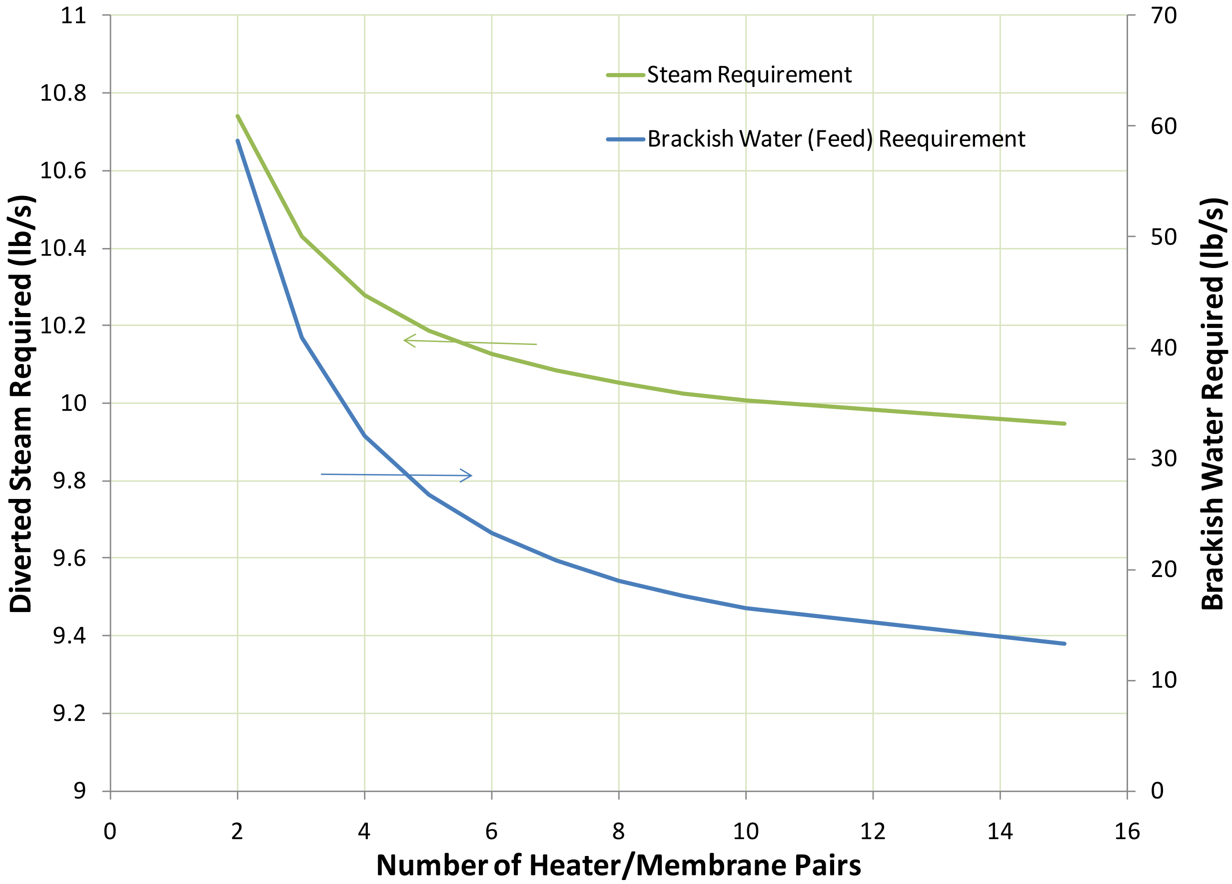


Figure 26: Impact of Heating Cycles on Steam Brackish Water Demand

5.3 Process Equipment

As with the scheme for recovering waste water from boiler blow down equipment demands are minor. The equipment lists in Figure 25 bear this out. For example, the four heat exchangers have areas ranging from 160 to 700 square feet. These are on the low end of commercially available units and should be relatively low cost.

The cooling water pump, P-2.1, is relatively small at 5.8 hp.

The cooling water aerial cooler A-2.1 is the exception. The low approach temperatures mandated by the decision to close the system has resulted in a required heat exchange area of 37,800 ft^2. With a maximum commercially available area per aerial cooler in the range of 10,000 ft^2, we will need 4 units.

Total membrane area for the four unit scenario is 8900 ft^2. No cost information exists for these units as no commercial data base exists yet.

Cooling water system requirements scale linearly with desired distillate rate, but are independent of number of heater/membrane pairs for a fixed rate. Cooling water demand scales with membrane area, which remains constant for a constant distillate demand. With a constant cooling water rate for a fixed distillate demand, pump size and heat exchange area remain constant.

Total heat exchange area will scale linearly with distillate rate, and will rise slightly with number of heater/membrane pairs at a given distillate rate. The heat exchanger transfer area is a strong function of total duty, which is constant and is a slight inverse function of log mean temperature difference (LMTD), which will change slightly. As Figure 26 shows, the demand for diverted steam changes little with the number of heater/membrane pairs. The enthalpy change of the steam is held constant because inlet and outlet temperature and pressure do not change. With constant enthalpies and small variation in rate, variations in total duty are small. However, heat exchange area also varies inversely with LMTD. Since LMTD drops slightly with an increasing number of pairs, we can expect total heat exchange area to go up but only slightly.

Since heat exchangers are so small, their cost will be relatively insensitive to heat exchange area. Consequently costs will be driven by the number of units. The same is probably true for membrane distillation units based on their structural similarity to heat exchangers.

6 COOLING WATER SYSTEM WITH WATER RECOVERY

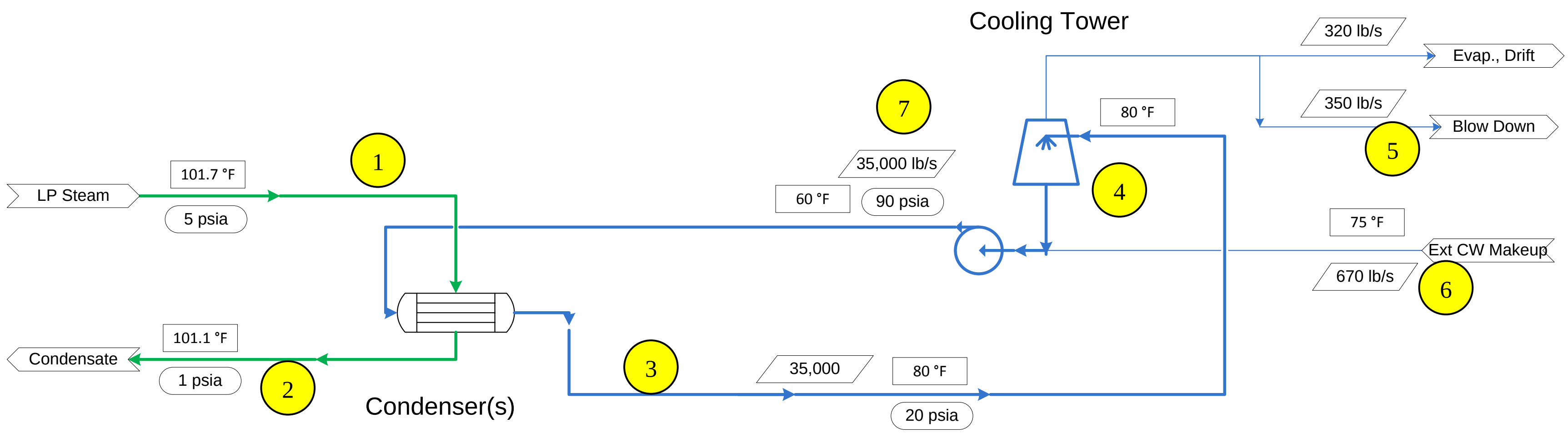
This chapter evaluates Location 3 in Figure 17, the cooling water system, as the heat source for membrane distillation. Of the 1.235 GW of energy input to steam power cycle, 0.589 GW (48%) gets rejected into the plant's cooling water system. It is by far the largest waste heat source in the plant. As a consequence, it is the most likely source of waste heat for water recovery. Unfortunately, as discussed in Section 3.1, power plants reject waste heat at low temperatures. In the case of the plant design used for this study, the temperature is 101.7 °F. This low temperature continues to frustrate analysts trying to do something constructive with this much energy. Membrane distillation may be the answer. However, the process proposed in this chapter does come with an increase in capital cost. Future optimzation studies will probably show that some of this capital cost can be traded off against compromises in power plant efficiency.

Section 6.1 contains a process description, including the process sketch in Figure 27 of a cooling water system typical of steam power plants and sized for the one used in this study. With this or any cooling system, steam condensers play a major role in power plant optimization. For this reason, Section 6.2 spends some effort discussing condenser design and the impact of approach temperatures on condenser size. With this background in place, Section 6.3 provides the process description for the recommended new cooling water system with water recovery using membrane distillation. The process sketch shown in Figure 30 can radically reduce cooling water makeup, taking it down to 130 lb/s from 670 lb/s. Section 6.4 describes the equipment, concentrating on new equipment and on changes to existing equipment. Section 6.5 provides a discussion of the impact of altered operating conditions on plant output. The intent is to provide enough information to begin the process of optimization/compromize between new equipment capital cost and tradeoff with plant operating income.

6.1 Typical Cooling Water System

The process description of Case 11 [5] includes few details regarding it's the cooling water system. Review of equipment summaries and process tabulations indicate the design assumes a closed loop water system with a cooling tower. The process is simple and would look like the one shown in Figure 27.

1. LP turbine effluent enters the condenser as a mixture of vapor and a small amount of aerosol sized water droplets. The condenser removes latent heat, converting all vapor to liquid.
2. Liquid exits the condenser with minimum pressure and temperature change.
3. Cooling water circulates between the condenser and a cooling tower.
4. The cooling tower cools water by direct contact with air, evaporation of some water removes the latent heat cooling the water further.



Legend

- 4150 psia — Pressure (psia)
- 512 °F — Temperature (°F)
- 650 ppm — Dissolved Solids (ppm mass)
- 6.89 lb/s — Flow Rate (lb/s)

Heat Exchanger Summary			
Unit	**Duty**		**Area**
	Btu/s	**Watts**	**ft²**
Condenser	669,000	7.05E+08	400,000
CW Tower	700,000	7.38E+08	n/a

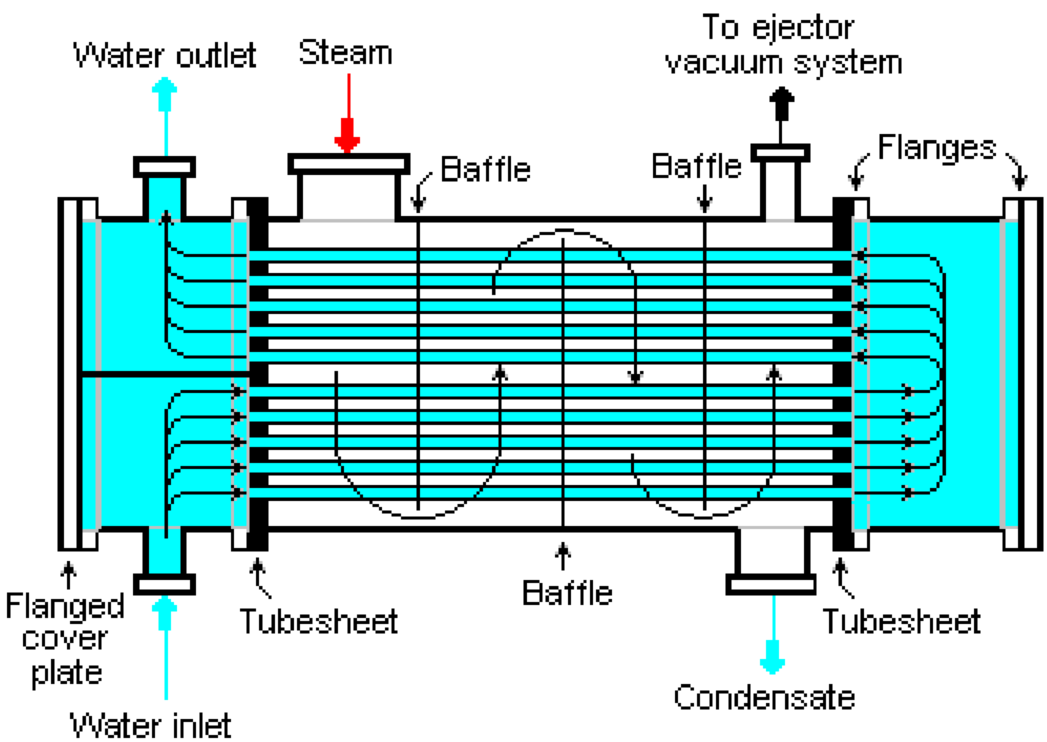
Pump Summary		
Unit	CWP	
System	CW	
ΔP	70	psi
Flow	35040	lb/s
Power	10290	hp

Figure 27: Typical CW System Sized for Case 11

5. Evaporation leaves behind salts in the cooling water. Buildup is controlled via cooling tower blow down.
6. Between evaporation and cooling tower blow down, a cooling water system requires significant makeup water.
7. Chilled water returns to the condenser via low head but extremely high rate pumps.

6.2 Condenser Size

The condenser plays an important part in cooling system design. Figure 28 shows schematically a typical condenser. Water flows through tubes while steam flows across the tubes' external surface, condensing in the process. Baffles encourage cross flow.



source: http://upload.wikimedia.org/wikipedia/commons/8/8b/Surface_Condenser.png

Figure 28: Typical Steam Condenser

An important parameter to condenser operation is the difference between cooling water temperature and steam temperature. The algorithms used in the final design of a heat exchanger are complex and best left to experts. However, at a conceptual level the design equation for the exchange shown in Figure 28 simplifies to

$$\dot{Q} = UA\Delta T_{LMDT} f_T \qquad 15.$$

In words, the heat exchanger duty (heat exchanged between steam and cooling water, $\dot{Q}$, is a function of a relatively constant heat transfer coefficient, U, the heat transfer area, A, a measure of the temperature difference, ΔT_{LMTD}, between the steam and the cooling water and f_T, a modifier to the temperature difference term that accounts for exchanger configuration and other things. In the case of the condenser, $f_T \approx 1$. For a condenser system operating at constant duty, heat exchange area depends on difference between the steam and the cooling water. The temperature difference between the steam and the inflowing cooling water is called the approach. It will vary through the condenser. The average difference, represented by ΔT_{LMTD} in equation 15, is also called the log mean temperature difference (LMTD).

Figure 29 illustrates the impact of changing water parameters on heat exchange area. In the next sections, this study will add components to the simple cooling water system that will raise both the hot temperature and cold temperature of the cooling water loop. Assume they both go up 10 F (T_{hot} to 90 °F and T_{cold} to 70 °F). From the curve, the needed heat exchange area for the condenser will increase by a factor of 1.55.

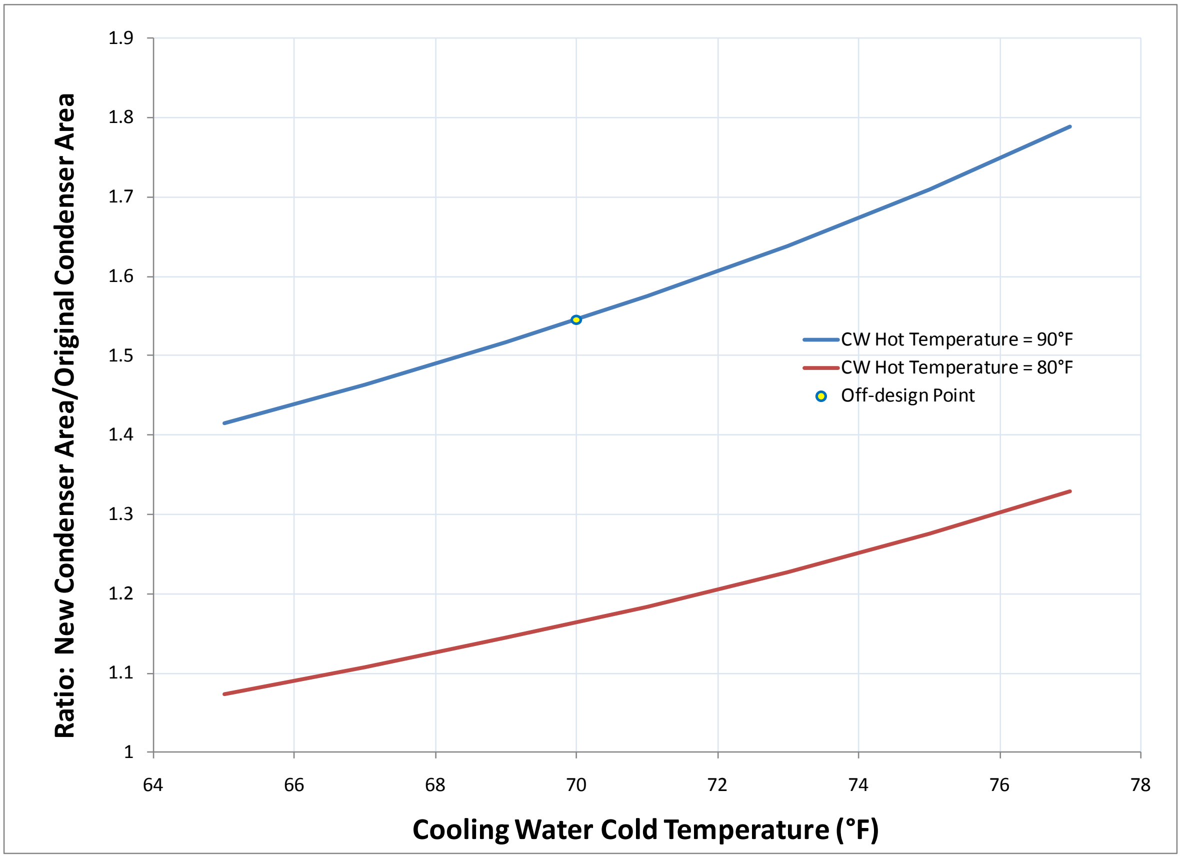


Figure 29: Impact of Approach Temperatures on Condenser Heat Exchange Area

Using the simple costing rule of thumb discussed in Section 7.2, condenser costs will rise by 36%.

6.3 Water Recovery Using Cooling Water System -- Process Description

Refer to Figure 30 for a process sketch of the scheme proposed for this water recovery system. The process is designed to remove all the waste heat rejected from the steam during condensation while not changing the power cycles operating parameters.

Steam Condensing Loop

1. LP turbine effluent enters the condenser as a mixture of vapor and a small amount of water as aerosolized droplets. The condenser removes latent heat, converting all vapor to liquid.
2. Liquid exits the condenser with minimum pressure and temperature change.

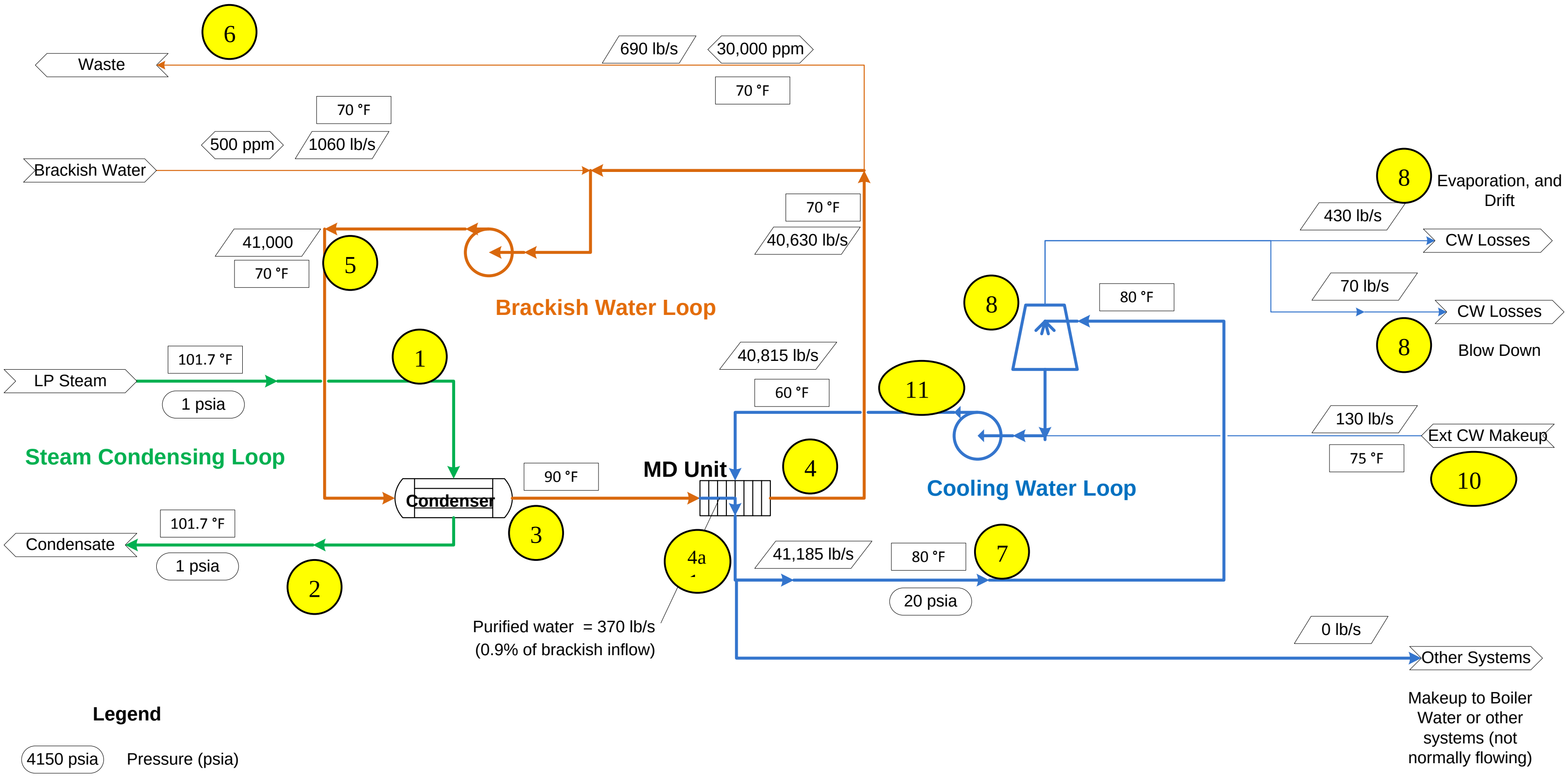
Brackish Water Loop

3. The condensing steam heats brackish water to about 90 °F.
4. This hot brackish water enters membrane distillation units where it cools to 70 °F. In the process, the brackish water contributes 370 lb/s of distilled water to the cooling water loop (4a). This pure water is in effect a pure makeup to the cooling water system.
5. Cooled water circulates through brackish water pumps back to the condenser inlet.
6. The distillation process in the membrane filters results in increased TDS in the brackish water stream. This is controlled through aggressive brackish system blow down.

Cooling Water Loop

7. Cooling water leaves the cold side of the membrane unit. The exiting water has gained temperature and distilled water. The temperature will be removed by the cooling tower.
8. The cooling tower cools water by direct contact with air. Evaporation of some water removes the latent heat of that water from the stream, cooling water further.
9. Evaporation leaves behind salts in the cooling water. Buildup is controlled via blow down. However, the use of distilled water as makeup significantly reduces the amount of blow down needed.
10. Nonetheless, the system still requires some makeup.
11. Chilled water returns to the condenser via low head but extremely high rate pumps.

Schematically, the steam condensing system remains essentially unchanged. However, the condenser heat exchange area will go up. This area increase results from the reduction in approach temperatures necessary to support the added brackish water loop, as discussed in Section 6.2.



Legend

4150 psia	Pressure (psia)
512 °F	Temperature (°F)
650 ppm	Dissolved Solids (ppm mass)
6.89 lb/s	Flow Rate (lb/s)

Membrane Filter Summary

Unit	Inflow lb/s	Purification lb/s	D. Solids ppm (mass)
MB-2.1	40600	370	10000

Heat Exchanger Summary

Unit	Btu/s	Watts
E-2.1	817589.5	8.618E+08
E-2.2	8.29E+05	8.735E+08

Pump Summary

Unit	P-2.1	P.2-2	
System	Brackish	CW	
ΔP	30	25	psi
Flow	41000	40815	lb/s
Power	6460	5353	hp

Figure 30: Process Sketch – Proposed CW System with Water Recovery

The brackish water loop removes latent energy from the low pressure steam, rejecting it to the cooling water system via membrane distillation units. At the same time, the membrane distillation units add purified water to the cooling water system.

This approach takes advantage of a membrane distillation unit characteristic typically considered a flaw. As discussed in Section 2.2.4, in an ideal membrane system the latent heat of the purified water would be the only heat transferred. However, today's membranes are so thin that sensible heat transfer becomes unavoidable. In this process, removal of excess heat is a bonus in that the membrane distillation unit acts as both a heat exchanger and a water purifier.

Waste heat removal requires a flow rate of. 41,000 lb/s (295,000 gpm). This large pump requirement effectively equals the requirements of the cooling system itself.

The membrane distillation unit means that dissolved solids will accumulate in the brackish water system. As a result, the system needs brackish water blow down. Assuming a very high allowable dissolved solids level (30,000 ppm TDS), the blow down will be 695 lb/s. This high blow down quantity combines with the amount of membrane cross flow adds up to a total brackish water makeup requirement of 1065 lb/s (7660 gpm).

A cooling water system has changed only slightly. It now cools the membrane distillation unit instead of the condenser directly. Reduced approach temperatures mean that equipment is slightly larger.

Cooling tower duty has increased, bringing with it slightly larger losses from vaporization and drift. Cooling water makeup comes almost entirely from the membrane distillation units, with the improved quality of internal makeup water (membrane transfer), required blow down drops to only 70 lb/s. See Section 4.1. In all the system needs 130 lb/s of external makeup.

6.4 Water Recovery Using Cooling Water System -- Process Equipment

This system comes with some significant equipment needs. The brackish water system consists of almost all new equipment. The steam condensing system and the cooling water system both experience significant size increases.

For example, condenser size will increase by 34.4%. The conceptual design equation for condensers is

$$\dot{Q} = UA\Delta T_{LMTD} f_T \qquad 16$$

Duty, $\dot{Q}$, is constant. Assume heat transfer coefficient, U, and temperature correction factors, f_T, are constant. One can derive quickly the following ratio:

$$\frac{A_{24}}{A_{23}} = \frac{\Delta T_{LMTD,27}}{\Delta T_{LMTD,29}} \quad (17)$$

The subscripts refer to the systems in Figure 27 and Figure 30. Based on Figure 27, condenser LMTD would be 30.3 °F. With the new process shown in Figure 30, LMTD would be 19.7 °F. The area ratio will be 1.54. This equates to a 54% increase in required surface area. Using the estimator's 7/10 rule (Section 7.2), condenser cost would increase by 52.5%.

Table 5 contains a summary of impacted equipment showing both original size (Figure 27) and new size (per Figure 30). Cost increases are estimated in the table using the 7/10 rule. Note that in a typical case all three pieces of equipment would need to be resized, so that cost impact would be additive.

Table 5: Cooling System Water Recovery Process Resized Equipment

Description	Sizing Parameters			Cost Impact
	Original Size	New Size	Units	
Condenser HEX area	400,000	610,000	ft^2	34.4% Incr
Cooling Water Pump	252,000	295,000	gpm	11.4% Incr
Cooling Tower Duty	2,890	2,520	MBtu/h	12.5% Incr

Table 6 is a list of the equipment needed for the new process (Figure 30) that is additional to the original (Figure 27). The cost of the brackish water pump was estimated based on similar pumps costed in the report DOE/NETL-2007/1281 [5] and adjusted by the the 7/10 rule. Bare module costs are the cost of the pump, its piping, instrumentation and infrastructure as installed without overhead of profit. This cost is sometimes called the total direct cost.

Table 6: Cooling System Water Recovery Process Additional Equpment

Description	Sizing Parameter		Estimated Installed Cost (Bare Module)
	Value	Units	\$2006 million
Brackish Water Pump	295,000	GPM	\$2.094
Membrane Distillation unit	2,391,000	ft^2	TBD

No estimate is available for the membrane distillation units. They are a new component without a commercial basis for their costs. While a reasonable estimate can probably be derived based on plate and frame exchangers, doing so would be a significant effort and would be outside the scope of this project.

6.5 Impact of Condenser Pressure/Temperature on Plant Power Output

As presented in this study, approach temperatures between the brackish water loop and the cooling tower loop were adjusted to maintain the same operating temperature at the condenser as if there were no MD system. Doing this requires large heat transfer areas in the condenser. The alternative is sacrificing some power output by increasing condenser temperature and pressure. This would reduce the high capital cost of the increased heat transfer area. This section provides an estimate of the relationship between increased pressure/temperature and total power plant output.

As already discussed, the plant net power output is proportional to the area of the power polygon plotted in Figure 15. One can infer by inspection of that figure that the area of that polygon changes radically with changes in condensation pressure/temperature. As condensation temperature rises, condensation pressure uniquely follows and the area of the polygon goes down.

Figure 31 illustrates this concept further. The figure was developed by varying LP Turbine discharge pressure while holding all other independent power plant parameters constant. As the figure shows, raising the LP Turbine discharge pressure (the bottom line of the polygon in Figure 15) from 1.0 psia (13 psi vacuum) to 14.7 psia (atmospheric pressure) reduces the Steam Plant output from 580 MW to 485 MW (Figure 29). With this scenario, 16.9% of the original power is lost. Under these circumstances, steam cycle efficiency drops from 47.2% to 39.3%.

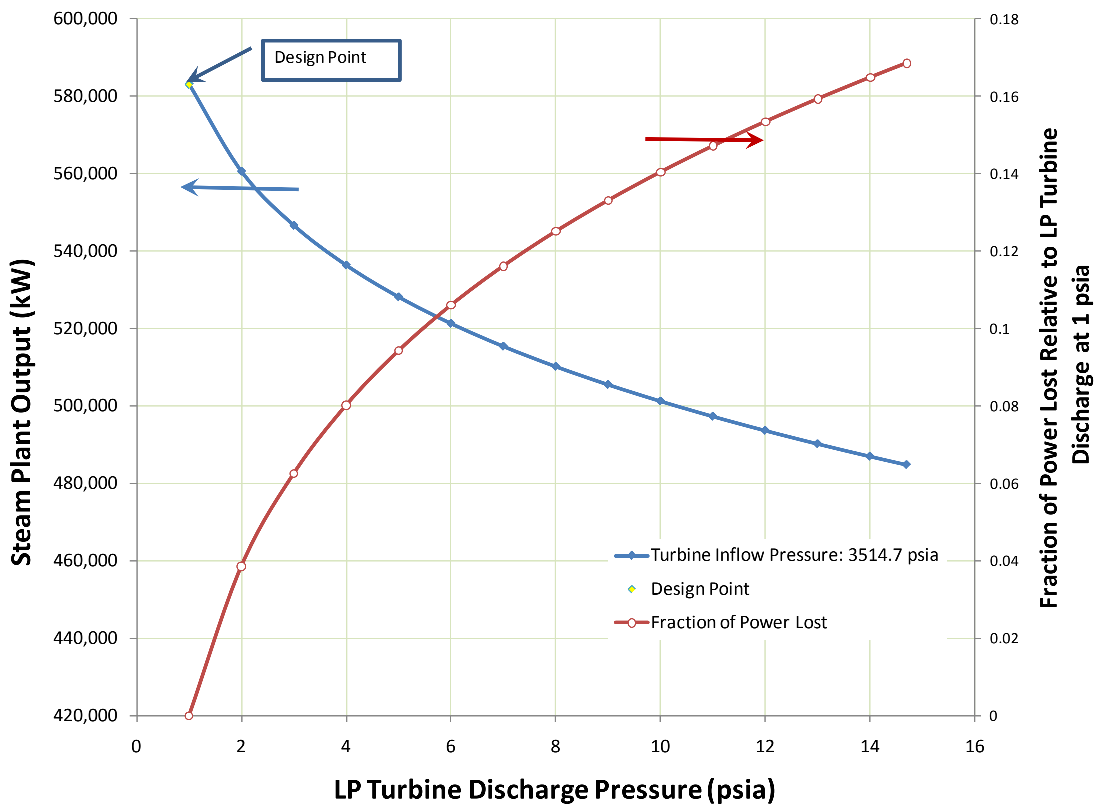


Figure 31: Impact of LP Turbine Discharge Pressure on Plant Power Output

Losing this much power production capacity significantly impacts power plant profits. As discussed in more detail in Section 7.1, even a reduction in efficiency of only 1.6 percentage points can result in an impact to profits of $7.3 million for a plant this size.

7 ECONOMICS AND COSTING

Detailed cost estimating, life cycle costing and capital cost/operating cost optimizations are beyond the scope of this study. However, these issues lurk in the background of many of the technical discussions in this document. This chapter attempts to provide enough information regarding costs and cost trade-offs to make the document useful to readers who would take the economics further.

The first section takes the reader through some simple power plant economics. The intent here is to provide enough information to allow users to gauge for themselves the relative merits of alternatives. The second section documents a useful rule of thumb used throughout the paper to allow rapid adjustments to known costs for changes in capacity. The last section provides some information regarding capital cost estimation.

7.1 Power Plant Economics

Table 7 provides a simple analysis of the value of produced electricity. The power plant used in this study, case 11 of the reference [5], has a nameplate capacity of 0.550 GWe. If this plant existed and were on line selling electricity for 95% of the hours in an average year, it would sell 4.58 billion kWh of electricity during that year. On the average for the United States, the average retail price in 2010 for electricity is $0.0981/kWh.[20]

Table 7: Simple Power Plant Economics

Description	Value	Units
Plant Name Plate Capacity	5.50E+08	We
	5.50E+05	kWe
Utilization	95%	
Hours per year on line	8327.7	h/y
Annual Sales	4,580,000,000	kWh/y
Retail Sales Price	$0.0981	$/kWh
Gross Sales	$449,000,000	$/y
Sales Decrement	1.60%	
Impact of Decrement	$7,321,780	

This single power plant would produce $449 million per year in gross sales revenue. This is revenue for the whole electric power industry. It goes to not only the power plant but also to high voltage transmission, distribution and retail sales.

What is important here is that the industry is at equilibrium with this revenue stream. Staffing levels, maintenance plans, dividends, profit projections and everything else the comprising a utility's financial structure has been set. While all the components of this list have some elasticity, all but profit have a fixed component. For this reason, profit will be hit hardest if unexpected costs arise.

For an unexpected cost, assume our power plant output drops by 1.6%. Gross sales drop from $449 million to $442 million. Most of the loss in revenue will come from profit.

7.2 Costing Rule of Thumb

Costs don't rise linearly with capacity. Double the size of a pump and the cost of pump will not quite double. Several reasons exist for this:[9]

- Capacity depends on volume while cost tracks more with surface area,
- The original cost includes most of the infrastructure cost, and
- The original cost includes most of the development cost.

Irrespective of why, long experience has shown that for small changes costs tend to vary by the ratio of capacity to 0.7 power.

$$\$C_{new} = \$C_{old}\left(\frac{Cap_{new}}{Cap_{old}}\right)^{\frac{7}{10}} \qquad 18$$

The new and old costs, $\$C_{new}$ and $\$C_{old}$ respectively, can be the purchase price, the bare module installed price or the total price so long as both prices are on the same basis. New and old capacities, Cap_{new} and Cap_{old} respectively, depend on the equipment being costed. For example, the capacity of interest for a heat exchanger is heat transfer area. The capacity of interest for low pressure centrifugal pumps is flow rate. For high pressure pumps or compressors, the appropriate capacity would be horsepower.

For example, the addition of the brackish water loop resulted in an increase in heat transfer area from 400,000 ft^2 to 610,000 ft^2 (Section 6.4). The DOE/NETL reference study listed the purchase price for the original condensers if purchased in 2006 as $6.405 million.[5] Hence:

$$\$C_{new} = 6.403\left(\frac{610}{400}\right)^{\frac{7}{10}} = \$8.606\, million \qquad 19$$

This is an increase in condenser cost of $2.4 million, assuming the purchase was in 2006.

7.3 Typical Cost Curves

Capital cost estimates for equipment start with a purchased price (f.o.b. the factory). This cost then accumulates other costs (usually through factors) for inflation, transportation, infrastructure, foundation and housing, power connection, connecting and ancillary piping and local instrumentation for a total bare module or direct cost. This direct cost then accumulates such other indirect costs as engineering, off-site office activities, construction management and financing.

This section presents for comparison only information on this the first item, purchase price. The curves are reproduced from an NETL report.[21]

Figure 32 contains a curve of cost versus heat transfer area for shell and tube exchangers. For larger units costs follow the trends of the costing rule of thumb presented in the previous section. However, smaller units less than about 1,000 ft², exchanger cost becomes almost independent of area. Both the water recovery from blow down scheme (Chapter 4) and the proposed water recovery using redirect bleed steam (Chapter 5) specify heat exchangers in the 100 to 1000 ft² range. In this range, cost will follow more the number of exchangers than the exchange area.

Notice also that the curve stops at about 70,000 ft². In Chapter 6, changes in approach temperatures result in a need to increase condenser area by 210,000 ft2 from 400,000 ft2 to 610000 ft2. From this curve, the area increase would equate to 3 additional condenser units.

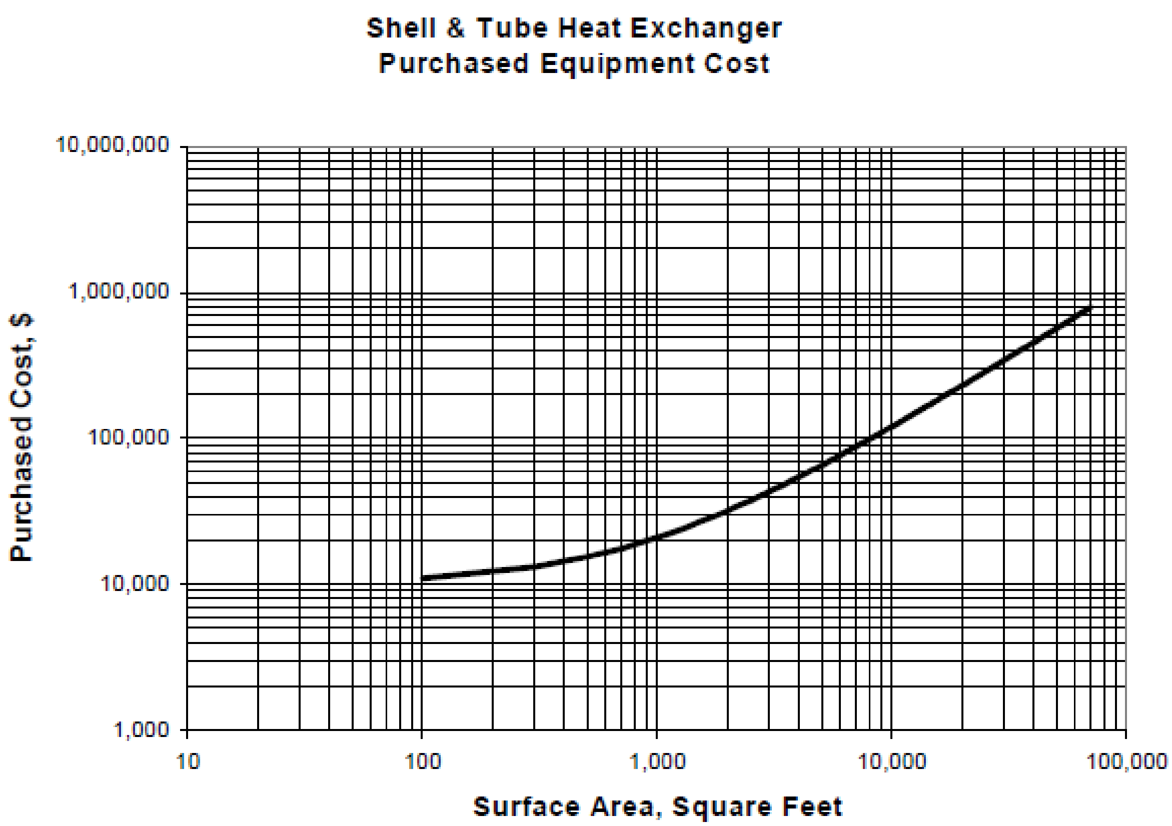


Figure 32: Purchase Cost for Shell and Tube Heat Exchangers in 1998 dollars

Each of these units would cost in 1998 $800,000. All three would have cost $2.4 million. Compare this to the estimated increase in the same system of $2.1 million calculated in Section 7.2. Some of the difference between the two estimates for the same system's increase can be attributed to inflation between 1998 and 2006. An equally likely source of the difference is the inherent uncertainty of the estimating methods.

A similar cost curve for aerial (fin-fan) coolers appears in Figure 33. It exhibits the same characteristic as did the heat exchanger curve. Small units are almost independent of size while larger units follow the rule of thumb discussed above. Note that the maximum size for an off the shelf aerial cooler is 10,000 ft². In Chapter 5, the new process will require 37,800 ft² of aerial cooling bare tube area. This equates to 4 cooler units at a cost in 1998 of $210,000 each at the factory.

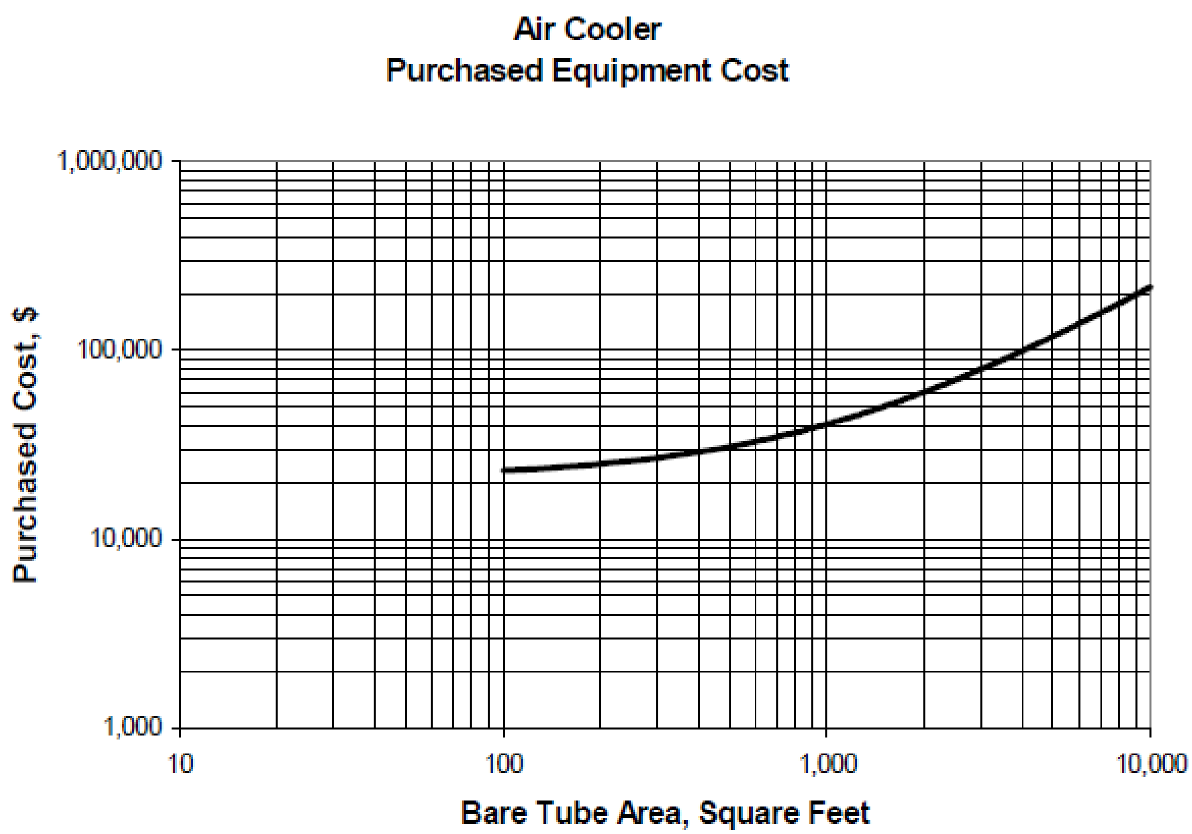


Figure 33: Purchase Price for Fin-Fan Coolers in 1998 dollars

The last curve included here is Figure 34 which shows the cost trend for centrifugal pumps. View this example as a cautionary tale. This curve stops at 10,000 gpm. The power plant design reference provides a cost for a single pump capable of 126,000 gpm for $590,000 apiece in 2006 dollars.[5]

Adjusting this cost to 1998 dollars, the pumps would cost 480,000. In the author's personal experience, these pumps would be special order, would represent the state of the art at time of purchase and would be engineered from the ground up.

Using the estimator's rule of thumb, the pump would have cost $400,000 apiece. In order to match the estimated 1998 purchase price, the exponent in equation 18 would have to increase from 0.7 to 0.855. The increased exponent reflects the added engineering cost. Buying multiple pumps would be the equivalent of an exponent of 1.0. Hence, for a pump costing a half a million dollars, is is still cost effective to pay for the engineering. This may not be so for lesser equipment.

This example leads to some lessons:

1. Equipment is available on special order and it can exceed the maximum capacity on the figures by at least an order of magnitude.
2. Rules of thumb do not track well with special orders.
3. Special orders only by out for very large purchases.

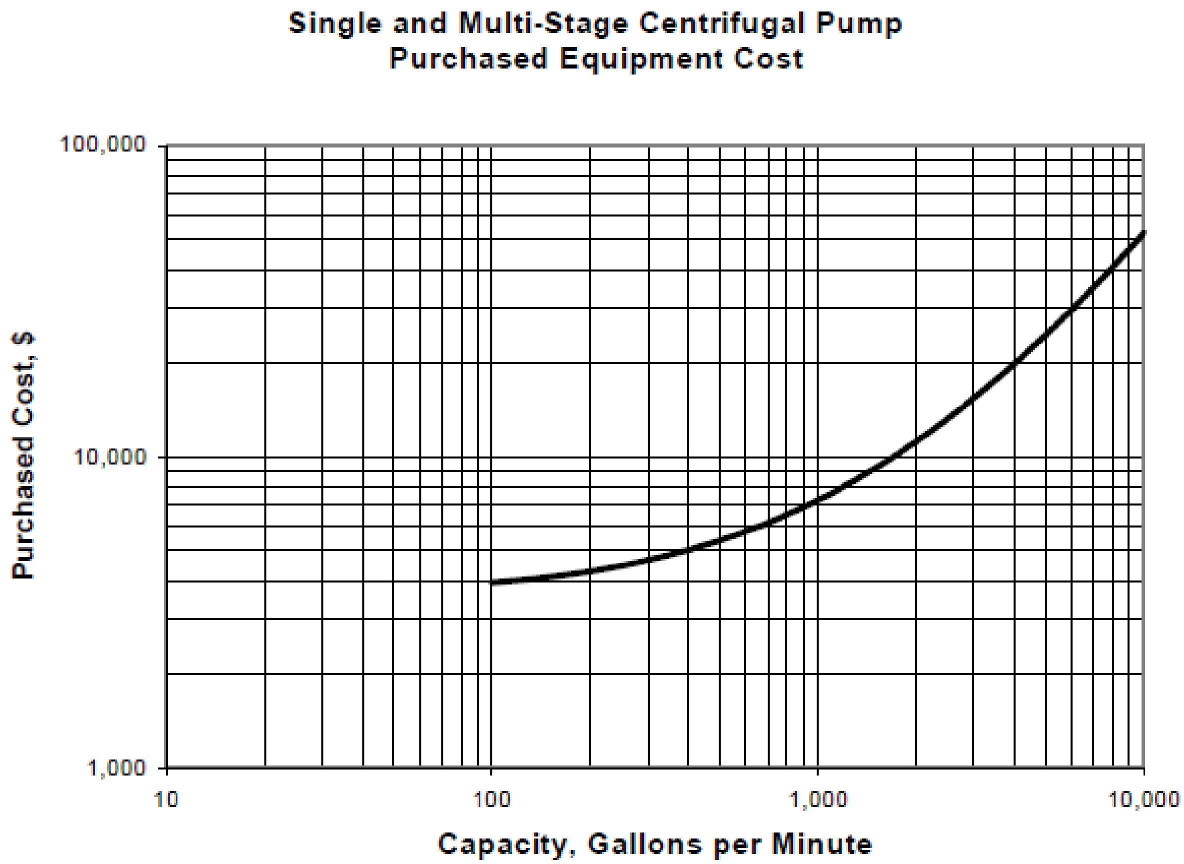


Figure 34: Purchase Price for Centrifugal Pumps in 1998 dollars

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APPENDIX A – STREAM SUMMARIES

Table A-1: Power Plant Process Stream Summary

	Description	Fluid	Mass Flow	Temperat	Pressure	Quality	Enthalpy	Entropy	Volume
			lb/s	°F	psia	lb.vap/lb.	Btu/lb	Btu/lb-°F	ft³/lb
1	HP Turbine In	supercritical steam	1017.99806	1100	3514.7	1	1497.525	1.520508	0.231775
1.1	HP Turbine Interwheel	steam	1004.16028	772.7499	1115.5	1	1366.801	1.5386	0.590583
2	HP Turb Disc/Reheat In	steam	843.749167	662.7981	710.8	1	1322.005	1.545349	0.852203
3	Reheat Out/Reheat Turbine In	steam	843.733611	1100	654	1	1572.59	1.743299	1.389596
3.1	Reheat Turb Interwheel	steam	796.026389	897.5706	310.1	1	1473.242	1.755759	2.560036
4	LP Turbine In	steam	795.206389	702.6444	137.7	1	1379.713	1.76948	4.954715
4a	LP Turb Interwheel	steam	366.966389	702.6444	137.7	1	1379.713	1.76948	4.954715
4b	LP Turb Interwheel	steam	342.038056	702.6444	137.7	1	1379.713	1.76948	4.954715
5a	LP Turb Interwheel	steam	311.385833	565.9719	72.7	1	1315.411	1.780141	8.299131
5b	LP Turb Interwheel	steam	314.716667	333.3108	19.2	1	1208.479	1.807257	24.37439
6a	LP Turb Interwheel	steam	289.0475	211.396	8.4	1.012795	1153.279	1.822057	47.17499
6b	LP Turb Interwheel	steam	289.0475	147.5146	3.5	0.977298	1102.538	1.836608	100.3846
7a	LP Turbine a Discharge	lp steam	289.0475	101.692	1	0.933815	1037.57	1.978924	333.4965
7b	LP Turbine b Discharge	lp steam	289.0475	101.692	1	0.933511	1037.255	1.978924	311.3239
7	LP Steam Combined Flow	lp steam	578.095	101.692	1	0.933511	0	0	#DIV/0!
7.1	Condenser In	lp steam	578.095	101.692	1	1.027373	1037.255	1.978924	311.3239
8	LP Boiler Feed Pump Suction	water		0	1	1.054374	1019.8	0	#DIV/0!
8.1	LP Boiler Feed Pump Discharge	water		0	0	0	0	0	#DIV/0!
8.2	Inter Bleed Heater Flow	water		0	0	1.054374	0	0	#DIV/0!
8.3	Inter Bleed Heater Flow	water		0	0	1	0	0	#DIV/0!
8.4	Inter Bleed Heater Flow	water		0	0	0.977298	0	0	#DIV/0!
8.5	Inter Bleed Heater Flow	water		0	0	0.977298	0	0	#DIV/0!
8.6	Dearator In	water		0	0	0.977298	0	0	#DIV/0!
9	HP Boiler Feed Pump Suction	water		0	0	0.944702	0	0	#DIV/0!
9.1	HP Boiler Feed Pump Discharge	water							
9.2	Inter Bleed Heater Flow	water							
9.3	Inter Bleed Heater Flow	water							
10	Boiler Feed	water							
11	HP Steam Bleed 1	steam							
13	Hot Water	water							
21	HP Steam Bleed 2	steam							
22	HP Steam Bleed 2	steam							
23	Hot Water Inter Bleed Heater	water							
31	IP Steam Bleed 1	steam							
32	IP Steam Bleed 1	steam							
33	Hot Water Inter Bleed Heater	water							
41	IP Steam Bleed 2	steam							
51	LP Steam Bleed 5a	steam							
53	Hot Water Inter Bleed Heater	water							
61	Lp Steam Bleed 5b	steam							
62	Lp Steam Bleed 5b	steam							
63	Hot Water Inter Bleed Heater	water							
71	LP Steam Bleed 6a	steam							
72	LP Steam Bleed 6a	steam							
73	Hot Water Inter Bleed Heater	water							
81	LP Steam Bleed 6b	steam							
82	LP Steam Bleed 6b	steam							
83	Hot Water Inter Bleed Heater	water							
91	Seal Leakage	steam							
93	Hot Water Inter Bleed Heater	water							
101	Cooling Water Cold	water							
102	Cooling Water Hot	water							

Table A-2: Water Recovery Using Boiler Blow Down – Equipment Sizing Summary

	Mass Flow					Temperature				Pressure				Enthalpy		
Description	Hot Inflow	Hot Outflow	Xfer	Cold Inflow	Cold Outflow	Hot Inflow	Hot Outflow	Cold Inflow	Cold Outflow	Hot Inflow	Hot Outflow	Cold Inflow	Cold Outflow	Hot Inflow	Hot Outflow	Cold Inflow
	lb/s	lb/s	lb/s	lb/s	lb/s	°F	°F	°F	°F	Psia	Psia	Psia	Psia	Btu/lb	Btu/lb	Btu/lb
	10.1					191.399				70.00				159.73		
MB-1.1	10.100	9.199	0.901	9.286	10.187	191.40	90.60	80.60	184.00	65.00	60.00	20.00	15.00	159.72	58.85	48.75
E-1.1	10.100	10.100	0.000	9.199	9.199	288.87	194.45	90.60	194.00	4155.00	4150.00	60.00	55.00	266.41	172.18	58.85
MB-1.2	9.199	8.378	0.821	8.457	9.278	194.00	90.60	80.60	184.00	55.00	50.00	20.00	15.00	162.31	58.82	48.75
E-1.2	10.100	10.100	0.000	8.378	8.378	373.09	288.87	90.60	194.00	4165.00	4160.00	50.00	45.00	352.24	266.42	58.82
MB-1.3	8.378	7.630	0.748	7.702	8.450	194.00	90.60	80.60	184.00	45.00	40.00	20.00	15.00	162.29	58.79	48.75
E-1.3	10.100	10.100	0.000	7.630	0.748	447.34	373.09	90.60	194.00	4175.00	4170.00	40.00	35.00	430.41	352.25	58.79
MB-1.4	7.630	6.949	0.681	7.015	7.696	194.00	90.60	80.60	184.00	35.00	30.00	20.00	15.00	162.26	58.77	48.75
E-1.4	10.100	10.100	0.000	6.949	0.681	512.00	447.34	90.60	194.00	4185.00	4180.00	30.00	25.00	501.61	430.42	58.77
MB-5	6.949	6.329	0.620	6.389	7.009	194.00	90.60	80.60	184.00	25.00	20.00	20.00	15.00	162.24	58.74	48.75

Table A-3: Water Recovery Using Steam Diverted from Bleed Streams – Equipment Sizing Summary

	Mass Flow					Temperature				Pressure				Enthalpy			
Description	Hot Inflow	Hot Outflow	Xfer	Cold Inflow	Cold Outflow	Hot Inflow	Hot Outflow	Cold Inflow	Cold Outflow	Hot Inflow	Hot Outflow	Cold Inflow	Cold Outflow	Hot Inflow	Hot Outflow	Cold Inflow	Cold Outflow
	lb/s	lb/s	lb/s	lb/s	lb/s	°F	°F	°F	°F	Psia	Psia	Psia	Psia	Btu/lb	Btu/lb	Btu/lb	Btu/lb
Feed	32.053					70				60							
E-2.1	10.01	10.01	0.00	32.05	32.05	542.15	171.59	70	194	1100.5	1095.5	60	55	539.6	142.3	38.3	162.3
MB-2.1	32.05	29.19	2.86	29.47	32.33	194	90.6	80.6	184	55	50	30	20	162.3	58.8	48.8	152.2
E-2.2	10.01	10.01	0.00	29.19	29.19	556.97	542.15	90.6	194	1105.5	1100.5	50	45	841.5	539.6	58.8	162.3
MB-2.2	29.19	26.59	2.61	26.84	29.44	194	90.6	80.6	184	45	40	30	20	162.3	58.8	48.8	152.2
E-2.3	10.01	10.01	0.00	26.59	26.59	557.53	556.97	90.6	194	1110.5	1105.5	40	35	1116.4	841.5	58.8	162.3
MB-2.3	26.59	24.21	2.37	24.44	26.82	194	90.6	80.6	184	35	30	30	20	162.3	58.8	48.8	152.2
E-2.4	10.01	10.01	0.00	24.21	2.37	772.7	557.53	90.6	194	1115.5	1110.5	30	25	1366.8	1116.4	58.8	162.2
MB-2.4	24.21	22.05	2.16	22.26	24.42	194	90.6	80.6	184	25	20	30	20	162.2	58.7	48.8	152.2

Table A-4: Water Recovery Using Steam Diverted from Bleed Streams – Stream Summary

System	Description	Rate lb/h	Rate lb/s	Rate gpm	Temp F	P psia	Quality	Enthalpy Btu/lb	Density lb/ft3	Entropy Btu/lb-F	TDS
Steam	Condenser In	2772244	770.0677778		101.1	5	1	1131.479794	0.002949		0
	Condensate	2772244	770.0677778	5538.564402	101.1	1	0	69.7686834	61.97731		0
Brackish	Pump Suction	147601437.5	41000.3993	294887.4873	70	15	0	38.14149928	62.30137	0.074638999	29729.25
	Brackish Into Cond	147601437.5	41000.3993	294887.4873	70.02803501	45	0	38.25295238	62.30705	0.074681082	29729.25
	Brackish Out Cond	147601437.5	41000.3993	294887.4873	90	40	0	58.19396586	62.11797	0.111662396	29729.25
	Brackish Out MBF	146269331.6	40630.3699	292226.122	70.00000002	15	0	38.1414993	62.30137	0.074638999	30000
	Transfer	1332105.862	370.0294061	2661.365344	80.00000001	15	0	48.13639477	62.2162	0.093333141	0
	Brackish Blow Down	2502105.503	695.0293065	4998.864628	70	15	0	38.14149928	62.30137	0.074638999	30000
	Brackish Makeup	3834211.365	1065.058713	7660.229971	70	15	0	38.14149928	62.30137	0.074638999	500
CW	Cooling Water.Cold	146935394.5	40815.38735	293556.8244	60	40	0	28.2090096	62.37167	0.055565931	1000
	Cooling Water.Hot	148267500.3	41185.41676	296218.1898	80	20	0	48.15005756	62.21717	0.093330883	1009.066
	CW Evap/Drift	1549568.583	430.4357176	3095.826123	60	15	0	28.13816067	62.36666	0.055572427	1009.066
	CW Blow Down	247525.0423	68.75695619								
	CW Ext Makeup	464987.7637	129.1632677	928.9819638	75	15	0	43.13972625	62.26113	0.084031233	500
	CW Pump Suct	147182919.5	40884.14431	294051.3456	59.93735706	15	0	28.07545964	62.367	0.055451764	1000
	CW Pump Disch	147182919.5	40884.14431	294051.3456	59.95926249	40	0	28.16824003	62.37189	0.055487475	1000

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Jared Ciferno
National Energy Technology Laboratory
Pittsburgh, PA
Jared.Ciferno@netl.doe.gov

Richard Hammack
National Energy Technology Laboratory
Pittsburgh, PA
hammock@netl.doe.gov

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