Appendix A

Desalination Technologies

There is no best method of desalination. A wide variety of desalination technologies effectively remove salts from salty water (or extract fresh water from salty water), producing a water stream with a low concentration of salt (the product stream) and another with a high concentration of remaining salts (the brine or concentrate). Most of these technologies rely on either distillation or membranes to separate salts from the product water (USAID 1980, Wangnick 1998 and 2002, Wangnick/GWI 2005). Ultimately, the selection of a desalination process depends on site-specific conditions, including the salt content of the water, economics, the quality of water needed by the end user, and local engineering experience and skills. These processes are described in greater detail below.

Membrane and Filtration Processes

Membranes and filters can selectively permit or prohibit the passage of certain ions, and desalination technologies have been designed around these capabilities. Membranes play an important role in the separation of salts in the natural processes of dialysis and osmosis. These natural principles have been adapted in two commercially important desalting processes: electrodialysis (ED) and reverse osmosis (RO). Both of these concepts have been understood for a century, but commercialization lagged until the technology for creating and maintaining membranes improved. Although they have typically been used to desalinate brackish water, versions are increasingly being applied to seawater, and these two approaches now account for more than half of all desalination capacity. A growing number of desalination systems are also adding filtration units prior to the membranes in order to remove contaminants that affect long-term filter operation. Box 1 lists the characteristics of major filtration and membrane systems.

Box 1: Filtration/Membrane Systems

Microfiltration (MF) membranes are used to reduce turbidity and remove suspended solids and bacteria. MF membranes operate via a sieving mechanism under a lower pressure than either UF or NF membranes.

Nanofiltration (NF) membranes are used for water softening, organics and sulfate removal, and some removal of viruses. Removal is by combined sieving and solution diffusion.

Reverse osmosis (RO) membranes are used for both brackish water and seawater desalination and are capable of removing some organic contaminants.

Ultrafiltration (UF) membranes are used for removal of contaminants that affect color, high-weight dissolved organic compounds, bacteria, and some viruses. UF membranes also operate via a sieving mechanism.

Sources: Heberer et al. 2001, Sedlak and Pinkston 2001, NAS 2004

Electrodialysis

Electrodialysis is an electrochemical separation process that uses electrical currents to move salt ions selectively through a membrane, leaving fresh water behind. The process was commercially introduced in the mid 1950s, providing a cost-effective way to desalinate brackish water and spurring considerable interest in the use of membranes. ED can produce more product water and

less brine than distillation processes, can treat water with a higher level of suspended solids than RO, and needs fewer pretreatment chemicals. These systems produce water for industrial and power plant cooling towers, freshwater fish farms, and municipal uses; treat industrial wastes; and concentrate polluted groundwater for further treatment. In one innovative application of ED, a plant in Tenerife, Spain removes salts and sodium from wastewater and uses the product water to irrigate bananas (von Gottberg 1999).

ED works on the principle that salts dissolved in water are naturally ionized and membranes can be constructed to selectively permit the passage of ions as they move toward electrodes with an opposite electric charge. Brackish water is pumped at low pressure between stacks of flat, parallel, ion-permeable membranes that form channels. These channels are arranged with anion-selective membranes alternating with cation-selective membranes such that each channel has as an anion-selective membrane on one side and a cation-selective membrane on the other (Figure A-1). Water flows along the face of these alternating pairs of membranes in separate channels and an electric current flows across these channels, charging the electrodes. The anions in the feed water are attracted and diverted towards the positive electrode. These anions pass through the anion-selective membrane, but cannot pass through the cation-selective membrane and are trapped in the concentrate channel. Cations move in the opposite direction through the cation-selective membrane to the concentrate channel on the other side where they are trapped. This process creates alternating channels, a concentrated channel for the brine and a diluted channel for the product water.

ED membranes are arranged in a series of cell-pairs, which consist of a cell containing brine and a cell containing product water. A basic ED unit or "membrane stack" consists of several hundred cell-pairs bound together with electrodes on the outside. Feed water passes simultaneously in parallel paths through all of the cells to produce continuous flows of fresh water and brine (Strathmann 1992, IDA 1999, Lee and Koros 2002).

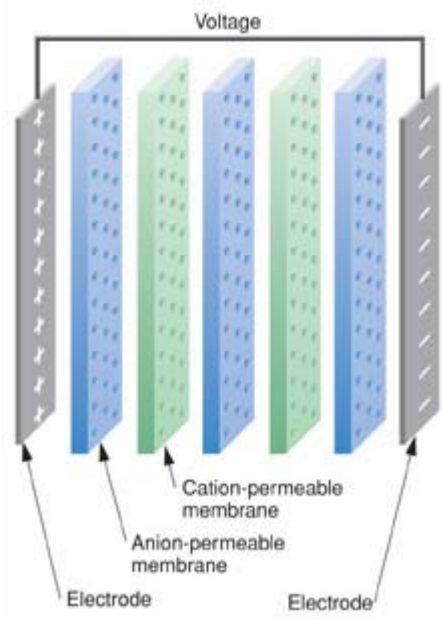

Figure A-1
Schematic of an Electrodialysis Desalination Plant

Source: Lawrence Livermore National Laboratory 2004

Electrodialysis Reversal

In the early 1970s, a modification of ED was introduced – electrodialysis reversal (EDR). An EDR unit operates on the same principle as a standard ED plant except that both the product and the brine channels are identical in construction. Several times an hour, the polarity of the electrodes is reversed, and the brine channel and product water channel flows are switched. Immediately following the reversal of polarity and flow, the ions are attracted in the opposite direction across the membrane stack and product water is used to clean out the stack and lines. After flushing for a few minutes, the unit resumes producing water. The reversal process breaks up and flushes out scale and other deposits in the cells. Flushing also allows the unit to operate with fewer pretreatment chemicals and minimizes membrane fouling.

EDR systems can operate on highly turbid feed water and are less prone to biofouling than RO systems. Experience suggests that EDR can also achieve higher water recovery than RO systems. The major energy requirement is the direct current used to separate the ions in the membrane stack. ED and EDR represent a very minor fraction – less than one percent – of worldwide desalination capacity (Wangnick/GWI 2005).

Reverse Osmosis

Reverse osmosis uses pressure on solutions with concentrations of salt to force fresh water to move through a semi-permeable membrane, leaving the salts behind (Figure A-2). The amount of desalinated water that can be obtained ranges between 30% and 85% of the volume of the input water, depending on the initial water quality, the quality of the product, and the technology and membranes involved.

An RO system is made up of the following basic components: pretreatment, high-pressure pump, membrane assembly, and post-treatment. Pretreatment of feed water is often necessary to remove contaminants and prevent fouling or microbial growth on the membranes, which reduces passage of feed water. Pretreatment typically consists of filtration and either the addition of chemicals to inhibit precipitation or efficient filtering to remove solids. A high-pressure pump generates the pressure needed to enable the water to pass through the membrane (Fisia Italimpianti 1999, IDA 1999).

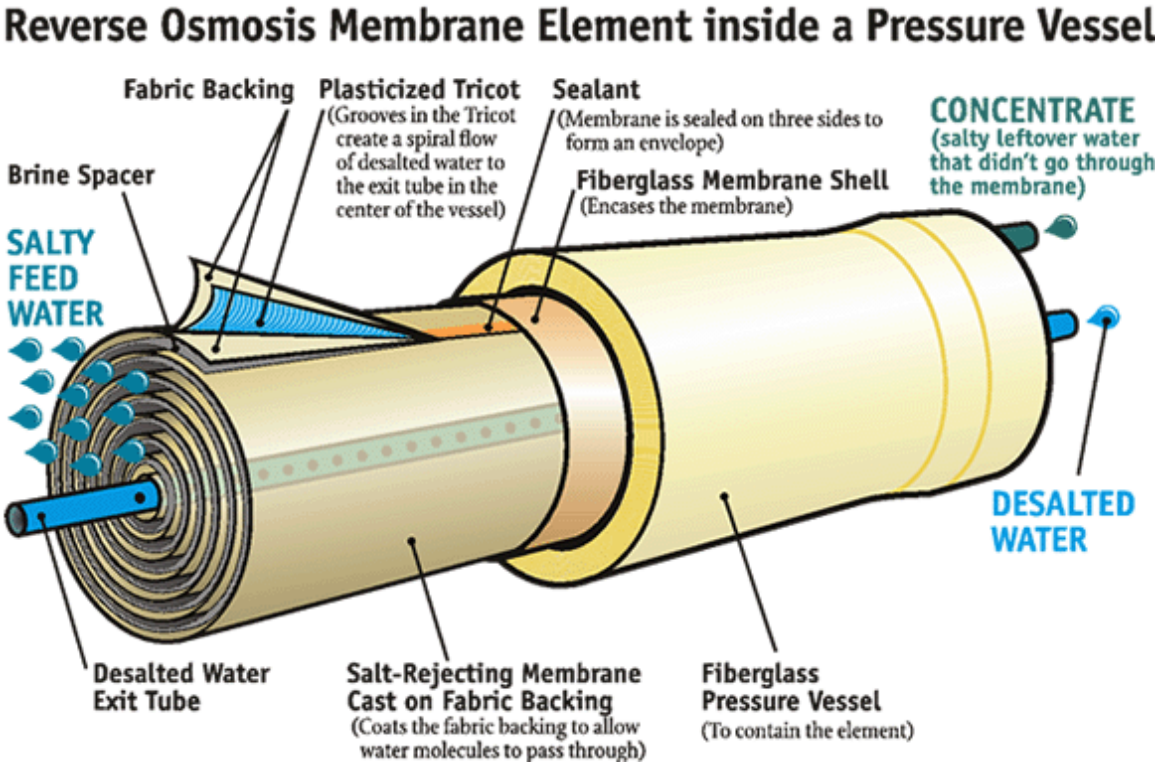

Figure A-2

Schematic of a Reverse-Osmosis Desalination Plant.

Source: U.S. Bureau of Reclamation Undated

The membrane assembly consists of a pressure vessel and a membrane that permits the feed water to be pressurized against the semi-permeable membranes. The membranes are fragile and vary in their ability to pass fresh water and reject salts. RO membranes are made in a variety of configurations. The two most commercially successful membrane configurations are spiral-wound and hollow-fine fiber. Post-treatment prepares final product water for distribution, removes gases such as hydrogen sulfide, and adjusts pH.

The energy requirement for RO depends directly on the concentration of salts in the feed water. Because neither heating nor phase change is necessary for this method, pressurizing the feed water accounts for the major use of energy. As a result, RO facilities are most economical for desalinating brackish water and increase in cost as the salt content of the water increases.

RO has become a relatively mature technology and membrane approaches are experiencing fast growth. Some of the largest new desalination plants under construction and in operation use RO membranes, including Ashkelon in Israel and the new plant at Tuas in Singapore. Ashkelon, the largest RO plant in the world, desalinates seawater for municipal purposes with a capacity of 100 million gallons per day (MGD), or 395,000 cubic meters per day (m^3/d) (Wangnick/GWI 2005).

Among the needed improvements in RO systems are better pretreatment of feedwater to reduce the use of chemicals that often end up in the brine and cause a disposal problem; improved membranes that are more durable and increase the flux of pure water; new approaches to reduce biofouling in membranes; more effective energy recovery and use; and development of less expensive materials (Awerbuch 2004).

Thermal Processes

Approximately 40% of the world's desalted water is produced with processes that use heat to distill fresh water from seawater or brackish water. The distillation process mimics the natural water cycle by producing water vapor that is then condensed into fresh water. In the simplest approach, water is heated to the boiling point to produce the maximum amount of water vapor. Water will boil at 100°C under atmospheric pressure. By decreasing pressure, however, the boiling point can be reduced. At one-quarter of normal pressure, water will boil at 65°C, and at one-tenth of normal pressure it will boil at only 45°C. To take advantage of this principle, systems have been designed to allow “multiple boiling” in a series of vessels that operate at successively lower temperatures and pressures. The concept of distilling water with a vessel operating at a reduced pressure has been applied for well over a century.

Distillation systems are often affected by scaling, which occurs when substances like carbonates and sulfates[1] found in seawater and brackish water precipitate out of solution and cause thermal and mechanical problems. Scale is difficult to remove and reduces the effectiveness of desalination operations by restricting flows, reducing heat transfer, and coating membrane surfaces. Ultimately scaling increases costs. Keeping the temperature and boiling point low slows the formation of scale.

[1]One of the most significant concerns is gypsum, a hydrate of $CaSO_4$ that forms from solution when water approaches about 95°C. Gypsum is the main component of concrete and can coat pipes, tubes, and other surfaces.

Multi-Stage Flash Distillation

Multi-stage flash distillation (MSF) accounts for the greatest installed thermal distillation capacity. Like all evaporative processes, MSF can produce high-quality fresh water with very low salt concentrations (10 ppm or less), from salt concentrations as high as 60,000 to 70,000 ppm total dissolved solids, nearly twice the salinity of seawater. In MSF, evaporation or "flashing" occurs from the bulk liquid, not on a heat-exchange surface, as is the case with other distillation processes (see Multiple-Effect Distillation, below). This minimizes scale and is a major reason MSF has been popular for several decades (Birkett 1999). Until recent advances in membrane technology, MSF was the primary technology used for desalinating seawater.

In MSF distillation, water is heated in a series of stages. Typical MSF systems consist of many evaporation chambers, each with successively lower pressures and temperatures that cause flash evaporation of hot brine, followed by condensation on cooling tubes. The steam generated by flashing is condensed in heat exchangers that are cooled by the incoming feed water. This warms up the feed water, reducing the total amount of thermal energy needed.

Generally, only a small percentage of feed water is converted to water vapor, depending on the pressure maintained in each stage. MSF plants may contain between 4 and 40 stages, but most typically are in the range of 18 to 25. Multi-stage flash plants are typically built in sizes from 2.6 MGD (10,000 m^3/d) to over 9.2 MGD (35,000 m^3/d), with several units grouped together. As of early 2005, the largest MSF plant in operation was in Shuweihat in the United Arab Emirates. This plant desalinates seawater for municipal purposes with a total capacity of 120 MGD (455,000 m^3/d) (Wangnick/GWI 2005).

Multiple-Effect Distillation

Multiple-effect distillation (MED) is a thermal method that has been used successfully for well over 100 years, substantially predating MSF (Birkett 1999). MED takes place in a series of vessels or "effects" and reduces the ambient pressure in subsequent effects. There are 8 to 16 effects in a typical large plant. This approach reuses the heat of vaporization by placing evaporators and condensers in series. Vapor produced by evaporation can be condensed in a way that uses the heat of vaporization to heat salt water at a lower temperature and pressure in each succeeding chamber, permitting water to undergo multiple boilings without supplying additional heat after the first effect. In MED plants, the salt water enters the first effect and is heated to the boiling point. Salt water may be sprayed onto heated evaporator tubes or may flow over vertical surfaces in a thin film to promote rapid boiling and evaporation.

Only a portion of the salt water applied to the tubes in the first effect evaporates. The rest moves to the second effect, where it is applied to another tube bundle heated by the steam created in the first effect. This steam condenses to fresh water, while giving up heat to evaporate a portion of the remaining salt water in the next effect. The condensate from the tubes is recycled.

Although some of the earliest distillation plants used MED, MSF units – with lower costs and less tendency to scale – have increasingly displaced this process. In the past few years, however, interest in the MED process has been renewed and MED appears to be gaining market share. According to the Wangnick/GWI desalting inventory, MED has a 15% share of the thermal market, but 21% share of proposed projects (Wangnick/GWI 2005). MED plants are typically built in units of 0.3 to 3 MGD (1,000 to 10,000 m^3/d) for smaller towns and industrial uses.

Vapor Compression Distillation

Vapor compression (VC) distillation has typically been used for small- and medium-scale desalting units. These units also take advantage of the principle of reducing the boiling point temperature by reducing ambient pressure, but the heat for evaporating the water comes from the compression of vapor rather than the direct exchange of heat from steam produced in a boiler. The two primary methods used to condense vapor to produce enough heat to evaporate incoming seawater are mechanical compression or a steam jet. The mechanical compressor can be electrically driven, making this process the only one to produce water by distillation solely with electricity (Buros 2000).

VC units use a compressor to create a vacuum, compress the vapor taken from the vessel, and condense it inside a tube bundle that is also in the same vessel, producing a stream of fresh water. As the vapor condenses, it produces fresh water and releases heat to warm the tube bundle. Salt water is then sprayed on the outside of the heated tube bundle where it boils and partially evaporates, producing more fresh water. Steam jet-type VC units, also called thermocompressors, create lower ambient pressure in the main vessel. This mixture is condensed on the tube walls to provide the thermal energy (through the heat of condensation) to evaporate salt water on the other side of the tube walls. VC units are usually built in the 0.066 to 0.50 MGD (250 to 2,000 m^3/d) range and used for tourist resorts, small industries, and remote sites.

Other Desalination Processes

Water can be desalted though many other processes including small-scale ion-exchange resins, freezing, and membrane distillation. None has achieved much commercial success, and together they account for less than one percent of total desalination capacity (Wangnick/GWI 2005). Nevertheless, some of these approaches can be effective, and even preferable, under special circumstances.

Ion-Exchange Methods

Ion-exchange methods use resins to remove undesirable ions in water. For example, cation-exchange resins are used in homes and municipal water-treatment plants to remove calcium and magnesium ions in “hard” water. The greater the concentration of dissolved solids, the more often the expensive resins have to be replaced, making the entire process economically unattractive compared with RO and ED. At lower concentrations and for small-scale systems, however, these methods have proven effective. Thus some form of ion exchange is sometimes used for the final polishing of waters that have had most of their salt content removed by RO or ED processes (Birkett 1999).

Freezing

Freeze separation takes advantage of the insolubility of salts in ice. When ice crystals form, dissolved salts are naturally excluded. If the resulting pure ice crystals can be separated from the brine, desalinated water can be produced. Extensive work was done in the 1950s and 1960s on separation technology using freezing of water. In this approach, seawater is cooled to form crystals. Before the entire mass of water has been frozen, the mixture is usually washed and rinsed to remove the salts adhering to the ice crystals. The ice is then melted to produce fresh water. The most efficient freeze methods use vapor-compression freeze-separation systems.

Freezing has some theoretical advantages over distillation, including a lower minimum energy requirement, minimal potential for corrosion, and little scaling or precipitation. Among the disadvantages, however, is the difficulty of handling and processing ice and water mixtures. A small number of demonstration plants have been built over the past 40 years, but the process has never proven commercially feasible. The few demonstration plants built have largely been abandoned. Better commercial success has been achieved in the application of freezing to the treatment of industrial wastes.

Membrane Distillation

Membrane distillation (MD) combines the use of both thermal distillation and membranes and was introduced commercially on a small scale in the 1980s. The process relies primarily upon thermal evaporation and the use of membranes to pass vapor, which is then condensed to produce fresh water.

Thus far, MD has been used in only a few areas. Compared to the more commercially successful processes, MD requires more space and more pumping energy per unit of fresh water produced. The main advantages of MD lie in its simplicity and the need for only small temperature differentials to operate. MD probably has its best application in desalting saline water where inexpensive low-grade thermal energy is available, such as from industries or solar collectors.

Desired Technological Improvements

The technology for desalinating water continues to improve, driven by advances in technology, the need to reduce costs, and commercial competition. Recent reviews recommend that research focus on several areas, which include water quality sensor development, improved filtration, improved heat-transfer materials, and improved intake methods (NAS 2004). See below for a more detailed discussion.

Water Quality Sensor Development

In order to permit more effective application of filters and chemicals and reduce membrane fouling, development of sensors able to quickly and inexpensively analyze water quality and identify pathogens are needed. These improvements apply to all desalination systems, as they would allow more effective post treatment application.

Improved Filtration

Nanofiltration and ultrafiltration membranes are designed to reduce the concentration of certain ions or contaminants early in the desalination process. The use of such filters in other desalination processes can increase overall productivity of both membrane and distillation systems by removing sulfate, calcium, and other compounds in feedwater. Improving these approaches and reducing their cost would help performance of all desalination systems.

Improved Heat-Transfer Materials

Most desalination methods use various heat-transfer surfaces to facilitate the process and reduce costs. Current heat-transfer surfaces are often made of expensive corrosion-resistant materials, such as titanium and high-grade stainless steel. New nonmetallic or polymeric heat-transfer materials could reduce capital costs but additional research is required to produce reliable and

effective ones. Improvement in the design of heat-transfer surfaces could also improve operating efficiencies and reduce costs.

Improved Intake Methods

With few exceptions, current intake methods for seawater desalination plants, especially large plants, impinge and entrain substantial amounts of marine life. Intake methods should be improved to reduce the marine impacts of desalination plants.

Membrane Integrity Improvements

Membranes fail for a number of reasons, including oxidation by chlorine and metals and mechanical damage from sediment. These failures are expensive and permit pathogens or contaminants to compromise the quality of the final product. Improvements in membrane durability and integrity would reduce costs and increase system performance.

Membrane Selectivity

“Selective” membranes capable of removing specific contaminants from a water stream would increase the flexibility of a system and potentially reduce costs. Contaminants not targeted for removal, such as algal toxins, may remain in the purified water and pose a human health, economic, and/or environmental risk.

Reduced Membrane Fouling

Innovations in approaches to clean and restore fouled membranes are still needed. The fouling of membrane systems by organic and inorganic materials, including algae and bacteria, reduces membrane life and increases overall costs. Efforts to control fouling, i.e., pretreatment of source water, application of membrane-cleaning chemicals, or operational changes, are costly. Some efforts to develop fouling-resistant elements are underway. The wide variety of feedwater qualities makes it unlikely that complete resistance to fouling will be achieved.

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Appendix B

Desalination Capacity in California

Location	Total Capacity m^3/d	Process	Customer	User	Contract Year	Operation Year	Included in CCC 2004?	Notes
Carlsbad	110	RO	Municipality	MUNI	2002	2003	No	Demonstration plant
CA	3,270	HYBRID		POWER	1992	1992	No	Location not listed in Wangnick/GWI 2005
CA	1,090	HYBRID		POWER	1992	1992	No	Location not listed in Wangnick/GWI 2005
CA	1,440	RO		MUNI	1994	1994	No	Location not listed in Wangnick/GWI 2005
CA	660	VC	PG&E	POWER	1994	1995	No	Location not listed in Wangnick/GWI 2005
CA	1,635	RO		MUNI	1996	1997	No	Location not listed in Wangnick/GWI 2005
Avila Beach	2,180	RO	PG&E	POWER	1985	1985	No	Operational status unknown; calls not returned
Diablo Canyon	2,180	RO	PG&E	POWER	1986	1987	No	Redundant entry (see above)? Calls not returned
Diablo Canyon	2,180	RO	PG&E	POWER	1990	1990	Yes	Calls not returned
Huntington Beach	817	RO	CA EDISON	POWER	1992	1992	No	

Location	**Total Capacity m^3/d**	**Process**	**Customer**	**User**	**Contract Year**	**Operation Year**	**Included in CCC 2004?**	**Notes**
Lompoc	23,360	RO	City	MUNI	1992	1993	No	Never built
Long Beach	37,850	RO	City	MUNI	2001	2003	No	Not in operation yet
Los Angeles	18,925	RO	City	MUNI	1998	1999	No	Never built
Los Banos	1,302	RO	ACOE	IRR	1982	1983	No	Wrong source water; plant located in Central CA
Morro Bay	465	OTHER	PG&E	POWER	1961	1962	No	No longer in operation
Morro Bay	465	OTHER	PG&E	POWER	1962	1963	No	No longer in operation
Morro Bay	568	VC	PG&E	POWER	1984	1985	Yes	No longer in operation
Morro Bay	379	VC	PG&E	POWER	1985	1987	Yes	No longer in operation
Morro Bay	654		Duke Energy	POWER		mid 90s	No	Goschke 2006
Morro Bay	2,271	RO	City	MUNI	1991	1991	Yes	Emergency use only
Moss Landing	711	OTHER	PG&E	POWER	1964	1965	No	Operational status unknown; calls not returned
Moss Landing	711	OTHER	PG&E	POWER	1965	1966	No	Operation status unknown; calls not returned
Moss Landing	1,893	VC	PG&E	POWER	1986	1987	Yes	
Moss Landing	852	VC		POWER	2001	2002	No	Operational status unknown; calls not returned

Location	Total Capacity m^3/d	Process	Customer	User	Contract Year	Operation Year	Included in CCC 2004?	Notes
Moss Landing	1,450	VC		POWER	2002	2003	No	Operational status unknown; calls not returned
Orange County	11,340	ME	OCWD Simulator	MUNI	1971	1973	No	
San Francisco	113	RO	Dow Chemical	DEMO	1978	1979	No	Demonstration plant
San Onofre	650	OTHER	Southern Edison	POWER	1966	1967	No	No longer in operation
Santa Cruz	1,817	VC	PG&E	POWER	1985	1987	No	
Santa Cruz	189	VC		DEMO	1997	1998	No	Demonstration plant
Santa Barbara	26,100	RO	City	MUNI	1991	1992	Yes	Capacity reduced by more than 50%
Santa Catalina	500	RO	SCE	MUNI	1990	1990	Yes	Operational status unknown; calls not returned
Santa Catalina	757	RO	SCE	MUNI	1991	1991	No	Operational status unknown; calls not returned
Santa Catalina	600	RO	SCE	INDU	1989	1990	No	Operation status unknown; calls not returned
Total Capacity	**149,485**							

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Appendix C

The Tampa Bay Desalination Plant

The Tampa Bay desalination plant has sparked tremendous public interest. Communities in the United States and abroad were intrigued by both the contract structure between Poseidon Resources and Tampa Bay Water, a public-private partnership, and the incredibly low cost of the product water. The Prime Minister of Singapore, for example, sent a delegation to Florida to examine the plant and learn about the project contract (Johnson 2001). Yet, nearly 10 years after the regional water authority initiated the project, the plant has produced very little water. The desalination plant has been plagued by problems and is not expected to be operational until 2008. The experiences in Tampa Bay should serve as a cautionary tale, warning advocates against excessive optimism on price; indeed, cost cutting is in part responsible for the project's difficulties.

History and Background

In the 1990s, Tampa Bay faced a range of regional water problems. Population forecasts concerned area leaders about their ability to meet future water demands. Local groundwater overdraft was adversely affecting natural wetlands and lakes in the area and leading to salinity intrusion (Wright 1999). When the regional water utility's (West Coast Regional Water Supply Authority) renewable water use permits expired in 1992, a long and difficult review period commenced. The permit finally came before an administrative judge in 1996. As part of the agreement, the West Coast Regional Water Supply Authority accepted legal obligations to dramatically reduce groundwater pumping from 158 MGD (598,000 m^3/d) to 121 MGD (458,000 m^3/d) by December 31, 2002 or face a serious economic penalty. The agreement imposed additional reductions to 90 MGD (340,000 m^3/d) by 2008. Conservation rules were also imposed on the regional water utility. In return, the agency responsible for issuing the permits, Southwest Florida Water Management District (SWFWMD), was required to provide $183 million in tax money to help develop new water supply sources (NOAA 2003).

The West Coast Regional Water Supply Authority (which later became Tampa Bay Water) examined a number of solutions to reduce groundwater pumping, including groundwater pumping in other areas, reuse, surface storage, and a seawater desalination plant. Additional groundwater pumping and reuse were rejected. Desalination[2] emerged as a winner, in part due to its reliability during droughts. In October 1996, West Coast Regional Water Supply Authority issued a Request for Proposal for a commercial developer to design, build, operate, and own a desalination plant. The plant would provide 25 MGD (95,000 m^3/d) of the mandated withdrawal reduction, or about 15% of the utility's water supply (Pittman 2003a).

Four vendors submitted initial proposals in December 1997, and binding offers in the competitive bidding process were received in October 1998. In early 1999, Tampa Bay Water (formerly the West Coast Regional Water Supply Authority) selected S&W Water, LLC, a consortium of Poseidon Water Resources and Stone & Webster. Their proposal called for construction of the plant to commence in January 2001 on the site of the Big Bend Power Plant on Tampa Bay.

[2] Additional surface storage was completed in 2005.

Operation was to begin in the second half of 2002 (Heller 1999, Hoffman 1999). A total of 44 MGD (167,000 m^3/d) of feed water would be required to produce around 25 MGD (95,000 m^3/d) of potable water and 19 MGD (72,000 m^3/d) of brine. The potable water would then be sent 22 kilometers by pipeline to the municipal water supply plant for distribution to customers. On completion, the privately owned and operated facility would supplement drinking water supplies for 1.8 million retail water customers in Hillsborough, Pasco, and Pinellas counties and the cities of New Port Richey, St. Petersburg, and Tampa (Wright 1999, U.S. Water News 1999).

Some local residents and community groups opposed construction of the plant because of concerns about impacts on local ecosystems due to water withdrawals and brine disposal (Karp 1999, Pittman 2003b). A group called Save Our Bay, Air and Canals (SOBAC), though not opposed to desalination plants in general, argued that the proposed location was inappropriate. SOBAC contended that flushing of Tampa Bay is too slow to dilute the effluent, which would have elevated salinity and temperature as well as chemicals introduced during the treatment process. In addition, they argued that the dissolved oxygen content of the discharge from the Big Bend Power Station already violates its permits, and that the brine from the desalination plant might exacerbate the oxygen problem (SOBAC 2005). While their opposition was not sufficient to delay or halt construction of the plant, it does indicate that the siting of the plant was a contentious issue.

Desalination advocates were excited by the project and by the apparent price breakthrough. S&W Water, LLC made a binding commitment to deliver desalinated water in the first year of operation at an unprecedented wholesale cost of \$1.71/kgal (\$0.45/m^3), with a 30-year average cost of \$2.08/kgal (\$0.55/m^3) (Heller 1999). Even the highest of the four bidders offered a price between \$2.12 to \$2.54/kgal (\$0.56 and \$0.67/m^3), well below the cost of water from other recent desalination plants. Southwest Florida Water Management District (SWFWMD) agreed to provide 90% of the projected \$110 million in capital costs for construction of the plant and the pipeline needed to transport the water to the water supply system (Heller 1999).

By comparison, the same year, the Singapore Public Utility Board announced plans to build a 36 MGD (136,000 m^3/d) desalination plant, to produce water at an estimated cost of between \$7.50 and \$8.74/kgal (\$1.98 and \$2.31/m^3) (U.S. Water News 1999). Around the same time, new plants in Cyprus and Trinidad were projected to produce water for under \$7.57/kgal (\$2.00/m^3).

Unique conditions, difficult to reproduce elsewhere, contributed to low produced water costs. Energy costs in the region are low (around \$0.04 per kWh) compared to other coastal urban areas. Co-location also lowered the cost because the power plant provided infrastructure, supporting operations, and maintenance functions. Salinity of the source water from Tampa Bay is substantially lower than typical seawater: only about 26,000 parts per million (ppm) instead of 33,000 to 40,000 ppm typical for most seawater. In addition, financing was to be spread out over 30 years and the interest rate was only 5.2 percent (Wright 1999).

Status of Tampa Bay (Mid 2006)

The Tampa Bay project has been plagued with problems. In 2000, Stone & Webster declared bankruptcy and Poseidon became full owner of S&W Water, LLC. Poseidon Water Resources then hired Covanta Energy to construct the plant, and the partnership became Tampa Bay Desal.

Because Covanta Energy had such a poor bond rating and was unable to secure financing for construction bonds, Tampa Bay Water (the regional water authority), which had an AA bond rating, decided to purchase Tampa Bay Desal. As owner of the plant, Tampa Bay Water contracted with Covanta Construction (a subsidiary of Covanta Energy) to finish building the plant and Covanta Tampa Bay, Inc. (a subsidiary of Covanta Energy) to operate the facility. Before Covanta Tampa Bay could commence the separate 30-year operation and maintenance contract, however, Covanta Construction was required to meet certain test standards. The operation and maintenance contract was worth $300-360 million and was Covanta Tampa Bay's only asset (Wright 2003).

The Tampa Bay plant began the first tests of potable water production in March 2003, nearly a year behind schedule. Two months later, serious and unanticipated membrane fouling became apparent, decreasing the life of the membranes and raising costs considerably. Covanta officials blamed the problems on Asian green mussels growing on the power plant intake pipes and large intake water temperature variations from the power plant (Wright 2003). Tampa Bay Water officials blamed the problems on Covanta's pretreatment system, which did not effectively remove sediment and organisms from the intake water that caused fouling of the membranes (Heller 2004).

Disposing of the cleaning solution also became problematic: the membrane fouling required additional chemical use, and disposal of the additional cleaning solution would have violated the sewer discharge permit issued by Hillsborough County. Covanta Construction was unable to meet the test standards after repeated attempts and declared bankruptcy in 2003, followed rapidly by the bankruptcy of Covanta Tampa Bay, Inc. In 2004, the parties agreed to a settlement by which Tampa Bay Water retained full control of the facilities and operating contract for less than the cost of the construction contract.

On November 15, 2004, Tampa Bay Water agreed to a $29 million, two-year contract with American Water/Pridesa (both owned by Thames Water Aqua Holdings, a wholly owned subsidiary of RWE) to fix the plant. The money will not be paid until the plant is running. As part of the agreement, all the first-pass membranes are to be redesigned and replaced at a cost of over $6 million (Tampa Bay Water 2005). In February 2005, American Water/Pridesa began new test operations to check for undetected equipment problems. Fouling of the membranes with sediment and organic material remains a problem. In addition, many of the water pumps developed rust and corrosion problems because of cost cutting that led to the use of inappropriate materials (Pittman 2005).

The plant is expected to re-open in late 2006 for another assessment period after the $29 million in repairs are finished and is currently scheduled to be fully operational in January 2008, six years behind schedule. In a press release issued in early 2004, the new cost was described as $2.54/kgal ($0.67/m^3), up from an initial expected cost of between $1.71 and $2.08/kgal ($0.45 to $0.55/m^3) (Business Wire 2004). If this cost were accurate, it would still be a relatively inexpensive desalination plant.

To further complicate matters (and increase costs), Tampa Bay Water announced that they intend to operate the plant at less than full capacity because of concerns about exorbitant increases in customer rates. Because the cost of the desalinated water was higher than other options in the

area, blending with cheaper supply options would lower customer rates. In response, SWFWMD threatened to withhold financing for the plant because they believed that they were misled. The dispute went into mediation. In January 2006, the water authorities agreed that the plant could be operated at less than full capacity as long as groundwater pumping was reduced.

Environmentalists and activists strongly opposed the deal because they were concerned about the damage caused by additional pumping needed to meet demand if the plant operated at less than full capacity, particularly during droughts (Skerritt 2006). Operating the desalination plant at less than full capacity will also increase the cost of the water produced.

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Appendix D
Some Portfolio Theory Math for Water Supply

Constant-Reliability-Benefit Unit Costs

The reliability and cost of different water-supply options can vary, making comparisons between different options difficult. To create a level playing field, the Pacific Institute developed a method for adjusting estimated unit costs of water-supply options (including conservation and end-use efficiency) so as to keep the reliability for all options the same. The method borrows and adapts tools from financial portfolio theory.[3] It leads to constant-reliability-benefit unit costs that provide a more fair comparison between supply options with different uncertainty characteristics.

Finding constant-reliability-benefit unit costs involves a two-step process. First, a planner must specify a constant-reliability-benefit standard. For example, the water planner might say that water supply (or conservation measures) must equal drought year demand 97.5% of the time. Mathematically, this means that the annual average of the supply portfolio, A(P), minus two times[4] the standard deviation (SD) of the supply portfolio, SD(P), must be equal to future (planned for) drought-year demand, D_F:

$$A(P) - 2SD(P) = D_F \qquad (1)$$

Other reliability standards can be chosen according to a table present in any statistics textbook that shows the percentage of time a random variable will be more than a chosen multiple of the standard deviation from the average. For example, specifying a "1" in Equation 1 rather than a "2" yields a reliability standard of about 84 percent. Stated differently, a normally distributed random variable will be less than the average minus one standard deviation about 16% of the time, or one in six years.

The average supply of a portfolio is the sum of the average supplies of each of its parts. In our example, one compares combinations of the existing supply, A(E), with a new supply, A(N):

$$A(P) = A(E) + A(N) \qquad (2)$$

where $A(X) = \frac{1}{n}\sum_{i=1}^{n} Q_{xi}$

n = Number of years of annual flow data

Q_{xi} = Annual flow in year i from Source x

The standard deviation of a portfolio of sources depends on the standard deviation and average of each source, the correlation between the sources, and the percentage of water from each source. The standard deviation of a portfolio is the square root of the variance of the portfolio.

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[3]This work was supported in part by the U.S. Bureau of Reclamation. See Wolff and Kasower (2006).
[4] Or if expressed with an additional significant figure, as is common in statistics textbooks, 1.96.

The appropriate formula (modified by the author from Tucker et al. 1994) when two sources are involved is:

$$SD(P) = \sqrt{W(E)^2 S(E)^2 + W(N)^2 S(N)^2 + 2W(E)W(N)Rho(E,N)S(E)S(N)} \qquad (3)$$

where $W(E) + W(N) = 1$

$$W(X) = \frac{A(X)}{A(P)}$$

$$S(X) = \frac{SD(X)}{A(X)}$$

Rho(E,N) is the correlation coefficient between E and N

Formulas for the standard deviation and correlation coefficient (Rho) are provided in any statistics textbook, and one can calculate these summary statistics using a spreadsheet program. Combining Equations 1, 2, and 3 yields:

$$\sqrt{\left(\frac{A(E)}{A(P)}\right)^2 S(E)^2 + \left(\frac{A(N)}{A(P)}\right)^2 S(N)^2 + 2\left(\frac{A(E)}{A(P)}\right)\left(\frac{A(N)}{A(P)}\right)Rho(E,N)S(E)S(N)} = \frac{A(P) - D_F}{2A(P)} \qquad (4)$$

where $A(P) = A(E) + A(N)$

If one knows the average existing supply, the standard deviations of the existing and new sources of supply, and the correlation coefficient between supplies, Equation 4 will contain only one unknown, A(N). This is the average new supply required to ensure that the chosen reliability standard (97.5% in this case)[5] will be achieved. A(N) can be found by assuming a value for A(N), seeing how close or far apart the left and right hand sides of the equation are, and iteratively adjusting the assumed value until the value of A(N) that solves the equation is found. Table D-1 presents the solutions found in the body of this report (new surface water supply, desalination, and outdoor water conservation). Finally, the constant-reliability-benefit unit price for each option differs from the average unit price for each option by the ratio of $A(N)/D_N$. When A(N) equals growth in drought demand (D_N)[6], as with desalination and similar options, the average unit price for that water supply option is also the constant-reliability-benefit unit price. When A(N) is greater than or less than D_N, as with the surface water and outdoor conservation examples in Table D-1, the constant-reliability-benefit unit price for each option is higher or lower than the average unit price for that option, respectively.

[5] Replacing the "2" in the denominator on the right hand side with the appropriate value, as discussed above, yields the appropriate equation for other reliability standards.

[6] Recall that D_N = equals D_F-D_E.

Table D-1: Unit Cost Reliability Premiums Under Various Assumptions

Water Supply Options	Coefficient of Variance (SD/A)	Correlation of Supply Options (Rho(E,N))	A(N)
Surface Water	20%	1.0	3,333 AFY
Desalination	0%	0.0	2,000 AFY
Outdoor Water Conservation	10%	-1.0	1,667 AFY

Notes: AFY = acre-feet per year

D_F = future drought-year demand

Assumes coefficient of variance of the existing source of 10%; A(E)=10,000 AFY; D_F=10,000 AFY; reliability level of about 97.5 percent.

Mathematics of Blending When Water Quality Is Uncertain

As in the reliability mathematics, a two-step process is used to determine the appropriate blending of water supply sources needed to obtain a specified water-quality objective. First, a planner must specify a water-quality standard and probability of achieving that standard. For example, the planner might specify that water quality must be 500 parts per million (ppm) total dissolved solids (TDS) at least 99.5% of the time. Mathematically, this means that the average quality of the supply portfolio, A(QP), minus three times the standard deviation of the portfolio's quality, SD(QP), must equal the water quality target (500 ppm):

$$A(QP) - 3SD(QP) = 500 \qquad (5)$$

Other probabilities of achieving the target standard can be chosen using a table present in any statistics textbook that shows the percentage of time a random variable will be more than a chosen multiple of the standard deviation from the average. For example, a reliability standard of about 84% requires specifying a "1" in Equation 5 rather than a "3." Specifying a "0" rather than "3" would mean water quality will be worse than 500 ppm 50% of the time. In this case, blended quality is simply the arithmetic average of the quality of the water sources.

The average quality of a portfolio is the weighted sum of the average qualities of the blended water sources. In our example, only two sources are blended at a time:

$$A(QP) = W(1)A(Q1) + W(X)A(QX) \qquad (6)$$

where $W(1)$ = Percent of the portfolio from Source 1

$W(X)$ = Percent of the portfolio from Source X

X = Source 2 or 3

$W(1) + W(X) = 1$

$$A(Qy) = \frac{1}{n}\sum_{i=1}^{n} q_{yi}$$

$A(Qy) =$ Average quality of Source y

y = Source 1, 2, or 3

n = number of years of annual average quality data

q_{yi} = annual average quality in year i from Source y

The standard deviation of the quality of a portfolio of sources, SD(QP), depends on the standard deviation and average quality of each source, the correlation between the source qualities, and the percentage of water from each source. The standard deviation of a portfolio is the square root of the variance of the portfolio. The appropriate formula (modified by the author from Tucker et al. 1994) when two sources are involved is:

$$SD(QP) = \sqrt{W(1)^2 S(1)^2 + W(X)^2 S(X)^2 + 2W(1)W(X)Rho(1,X)S(1)S(X)} \qquad (7)$$

where $S(y) = \dfrac{SD(Qy)}{A(Qy)}$

SD(Qy) = Standard deviation of the quality of Source y

Rho(1,X) = correlation coefficient between

the quality of Source 1 and the quality of Source X

Formulas for the standard deviation (SD) and correlation coefficient (Rho) are provided in any statistics textbook and one can calculate these summary statistics using a spreadsheet program. Combining Equations 5, 6, and 7 yields:

$$\sqrt{(1-W(X))^2 S(1)^2 + W(X)^2 S(X)^2 + 2(1-W(X))W(X)Rho(1,X)S(1)S(X)} = \frac{A(QP)-500}{3A(QP)} \qquad (8)$$

where $A(QP) = (1-W(X))A(Q1) + W(X)A(QX)$

and $S(y) = \dfrac{SD(Qy)}{A(Qy)}$

As with the reliability example, there is only one unknown in Equation 8 if one knows the summary statistics related to water quality for the water-supply options (average quality, standard deviation of quality, and correlation coefficient between quality measures). The unknown is W(X), the fraction of the blend with Source 1 that must come from Source X in order to maintain 500 ppm or better 99.5% of the time. As before, one must solve for W(X) by

iteration. One then finds the fraction of the blend from source 1 by subtracting W(X) from 1. The cost of each blend that satisfies the quality specification is the weighted average cost using these fractions.

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Appendix E

Data and Statistics for the Water Quality Example

Table E-1: Flow Weighted Total Dissolved Solids (TDS) in Colorado River Water Below Parker Dam

Calendar Year	TDS (milligrams per liter)	Calendar Year (cont.)	TDS (cont.) (milligrams per liter)
1970	760	1986	535
1971	758	1987	538
1972	734	1988	540
1973	709	1989	559
1974	702	1990	600
1975	702	1991	624
1976	690	1992	651
1977	687	1993	631
1978	688	1994	673
1979	701	1995	671
1980	712	1996	648
1981	716	1997	612
1982	713	1998	559
1983	678	1999	591
1984	611	2000	580
1985	561	**Mean**	**648**
		Standard Deviation	**68**

Source:

Redlinger, J. 2005. Personal communication. Deputy Area Manager, Boulder Canyon Operations Office, U.S. Bureau of Reclamation.

Appendix F

Historical Energy Prices and Rainfall in California

Table F-1 shows the historical retail electricity, natural gas, and consumer price index values in Bureau of Labor Statistics data series (BLS 2006). These represent the San Francisco-Oakland-San Jose and Los Angeles-metropolitan areas. The upward trend in electricity and natural gas prices was stripped from the electricity and natural gas indices by subtracting the geographically appropriate consumer price index, and adding 100 (the index base year value).[7] Those parts of the adjusted annual time series for retail electricity and natural gas price indices that overlapped with the statewide average rainfall time series from the Department of Water Resources (Roos 2006) are presented along with the rainfall data in Table F-2.

Table F-1: Energy Price Indices Used in the Analysis

Year	LA electricity	LA natural gas	LA CPI	SF electricity	SF natural gas	SF CPI
1972	38.3	22.4	41.4	40.3	15.6	40.4
1973	41.8	23.8	43.7	42.2	17.4	42.8
1974	50.9	26.6	48.2	47.1	19.9	47.0
1975	53.8	30.9	53.3	49.5	25.7	51.8
1976	55.6	31.9	56.9	49.3	27.8	54.6
1977	58.7	33.1	60.8	54.2	31.1	58.8
1978	61.9	35.4	65.3	52.6	33.6	64.3
1979	66.8	46.6	72.3	47.4	44.0	69.8
1980	87.0	62.3	83.7	77.6	75.4	80.4
1981	92.4	66.8	91.9	89.8	76.1	90.8
1982	100.4	82.9	97.3	110.0	89.2	97.6
1983	97.8	106.9	99.1	90.0	99.2	98.4
1984	101.8	110.2	103.6	100.0	111.6	104.0
1985	105.5	111.4	108.4	116.3	105.5	108.4

[7] This is an approximation. One could also calculate a line of best fit for the consumer price index and subtract values from that line by year from the appropriate electricity or natural gas price index. Both approximations are rough, and are presented as illustrations of the type of analysis one can perform. That is, the analysis in this report is by no means final. It illustrates issues that should be addressed thoroughly.

Year	LA electricity	LA natural gas	LA CPI	SF electricity	SF natural gas	SF CPI
1986	110.1	110.3	111.9	115.6	97.0	111.6
1987	109.2	124.7	116.7	115.6	97.3	115.4
1988	119.8	116.2	122.1	128.1	104.6	120.5
1989	132.0	113.7	128.3	142.4	109.9	126.4
1990	139.9	116.4	135.9	152.8	113.8	132.1
1991	151.9	129.1	141.4	170.9	119.2	137.9
1992	156.9	116.5	146.5	173.2	119.8	142.5
1993	158.7	128.7	150.3	180.8	112.8	146.3
1994	161.5	129.4	152.3	180.3	116.7	148.7
1995	168.7	130.8	154.6	180.4	126.3	151.6
1996	163.7	128.4	157.5	175.4	100.2	155.1
1997	167.9	135.5	160.0	174.9	122.6	160.4
1998	155.0	128.8	162.0	156.2	121.2	165.5
1999	154.1	124.9	166.1	158.5	127.8	172.5
2000	154.1	159.3	171.6	154.4	158.6	180.2
2001	200.8	159.8	177.3	206.2	194.1	189.9
2002	230.9	140.2	182.2	222.7	132.4	193.0

Table F-2: Statewide Precipitation and Trend-Adjusted Energy Indices

Year	Precipitation (inches)	Adj. LA electricity	Adj. LA natural gas	Adj. SF electricity	Adj. SF natural gas
1971	22.92				
1972	14.89	96.9	81.0	99.9	75.2
1973	30.17	98.1	80.1	99.4	74.6
1974	29.83	102.7	78.4	100.1	72.9
1975	22.59	100.5	77.6	97.7	73.9
1976	14.08	98.7	75.0	94.7	73.2
1977	12.28	97.9	72.3	95.4	72.3
1978	35.41	96.6	70.1	88.3	69.3
1979	21.97	94.5	74.3	77.6	74.2
1980	30.65	103.3	78.6	97.2	95.0
1981	17.07	100.5	74.9	99.0	85.3
1982	35.42	103.1	85.6	112.4	91.6
1983	43.21	98.7	107.8	91.6	100.8
1984	22.53	98.2	106.6	96.0	107.6
1985	19.27	97.1	103.0	107.9	97.1
1986	30.91	98.2	98.4	104.0	85.4
1987	13.99	92.5	108.0	100.2	81.9
1988	17.60	97.7	94.1	107.6	84.1
1989	20.11	103.7	85.4	116.0	83.5
1990	15.29	104.0	80.5	120.7	81.7
1991	17.85	110.5	87.7	133.0	81.3
1992	20.66	110.4	70.0	130.7	77.3
1993	33.14	108.4	78.4	134.5	66.5

Year	Precipitation (inches)	Adj. LA electricity	Adj. LA natural gas	Adj. SF electricity	Adj. SF natural gas
1994	16.11	109.2	77.1	131.6	68.0
1995	38.58	114.1	76.2	128.8	74.7
1996	26.34	106.2	70.9	120.3	45.1
1997	26.64	107.9	75.5	114.5	62.2
1998	41.91	93.0	66.8	90.7	55.7
1999	21.45	88.0	58.8	86.0	55.3
2000	21.83	82.5	87.7	78.2	78.4
2001	17.83	123.5	82.5	116.3	104.2
2002	19.59	148.7	58.0	129.7	39.4
		123.6	86.7	122.6	85.7
Lagged Correlation		-0.27	0.08	-0.32	-0.04

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