WATER DESALINATION

1. Introduction

Of the surface of the earth, 71% (3.60×10^8 km^2) is covered by oceans; their average depth is 6 km and their volume is 8.54×10^8 km^3. Unfortunately, this huge quantity of water is not suitable for very many human uses due to high salinity, ranging between 25,000–45,000 ppm TDS (ppm = parts per million by weight, or mg/L, TDS = total dissolved solids). Water with over 1000 ppm salt is usually considered unfit for human consumption, and water with over 500 ppm is considered undesirable, but in some parts of the world, people and land animals are forced to survive with much higher concentrations of salts, sometimes of over 2500 ppm.

Freshwater with less than 500 ppm (or 0.05%) dissolved solids is generally considered to be potable.

About 60% of the land area of the earth is arid or semiarid and is not generally considered habitable. The oceans hold about 97% of the earth's water. More than 2% of the total water and over 75% of the freshwater of the world is locked up as ice in the polar caps. Of the remaining 1% of total water that is both liquid and fresh, some is groundwater at depths of >300 m and therefore difficutl to obtain, and only the very small difference, possibly 0.06% of the total water of this planet, is available for human use as it cycles from sea to atmosphere to land to sea. Wells produce groundwater, stored from previous rains. However, the fact that in recent years wells have had to be made deeper and deeper to reach water shows that groundwater is being used faster than it is being replenished. Water lying in deep strata for millions of years is being mined like other minerals, never to be replaced. Once pumped, these resources can not be replenished.

2. The Water Problem

Today in the United States, three times as much water is used per capita than in 1900; with inclusion of all industrial and agricultural uses, this quantity is probably ten times as great as at the turn of the century. Individual usage in some southern cities, with swimming pools, lawns, air conditioners, and other local demands, can be as much as twice the national average. Population increase multiplies the total withdrawals, particularly in cities, where they may be as much as 1.06 m^3/d/person. In New York City, for example, which did not practice adequate water metering, per capita consumption grew from 0.69 m^3/d/person in 1970 to 0.76 m^3/d/person in 1981. Improved metering and water conservation have reversed this trend over the recent years, and in 1995 the per capita consumption fell to 0.68 m^3/d/person, about the same as in 1970 (1). Philadelphia is also able to meter or bill only a part of the supplied water, and the total per capita water consumption has grown from 1970 to 1990 by 30%, to about 0.9 m^3/d/person, whereas the metered and billed accounts are expected to have a consumption of only 0.35 m^3/d/person (2).

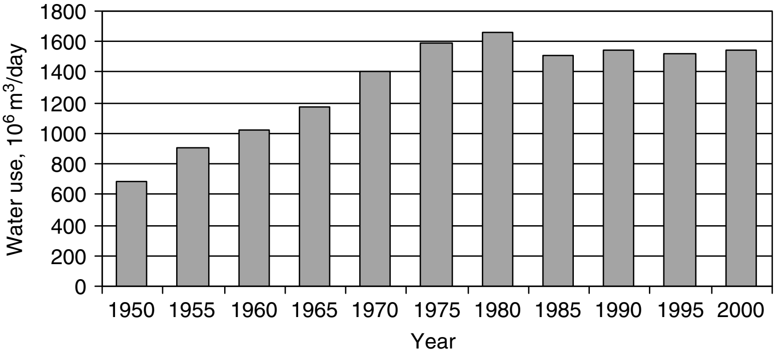


Fig. 1. Trends of estimated water use in United States 1950–2000 (3).

The use of ground and surface freshwater in the United States from 1950–2000 is shown in Figure 1 (3). The overall water usage had been maintain at the same level since 1985 despite of about 17% population grow. The stabilization of water use is primarily because of a reduction in water use for irrigation and thermoelectric power industry. In some part it is a result of conservation and improved efficiency, but in large part it is the result of a decline in agricultural and industrial production. Water for public supply continues to increase, approximately keeping pace with population growth.

An analysis of freshwater use in the United States predicted that by the year 2040, barring major changes in use patterns and natural water cycles, withdrawals and consumption will increase by about 40% (4). Most of this increase must be met by reuse, and thus water recycling should be raised significantly. For example, in manufacturing, recycling currently provides a reuse ratio, defined as the quantity ratio of (water actually used)/(its natural water source), of 1.3, which must be increased by the year 2020 to above 6.3. The balance between water supply and demand is worsening most seriously in the Rio Grande, upper and lower Colorado, Great Basin, High Plains, and California water resources regions of the United States.

In some areas of the United States, twice as much water is pumped from the ground as soaks into it. Although the amount of groundwater within 0.8 km of the surface is estimated at 3.8×10^{16} m^3, in some places the water table has dropped by 1–5 m for each year of the present generation, thus exhausting a historical treasure. Not only is the groundwater being depleted, but often this withdrawal causes sinking of the ground level, as it has near Houston, Texas; Mexico City; and in Florida. Las Vegas, Nevada, is growing rapidly upon a ground level that has sunk about a meter in recent years as a result of greatly increased mining of prehistoric water. This water supplies many acres of swimming pools, many thousands of "tons of refrigeration," for air conditioning, and other water uses. It is water from wells in a desert where only 7.5–10 cm of rain falls each year. Over-pumping since the 1920s, mostly to supply the needs of Southern California, has caused the land in the California San Joaquin Valley to sink by as much as 9 m.

Southern California was suffering from seawater intrusion in certain aquifers because of overpumping. To minimize the problem, the Orange County Water

District in 1975 began a pioneering effort, known as Water Factory 21, to treat 57,000 m^3/d of sewage by secondary effluent treatment processes, purify about one third of the product water by reverse osmosis (RO), blend the remainder of the treatment product with the RO process product and with high quality groundwater, and inject the mixture into the aquifer. Apart from recharging thereby the aquifer, it also reduces, if not stops, seawater intrusion into it (5,6). The Orange County reclamation plant produces drinking-quality water for injection with no adverse effects on groundwater. This plant is currently undergoing a significant expansion and modification, changing the previous conventional treatment configuration to integrated membrane system design (see Section 5.3).

Water is far from evenly distributed in the United States, with major shortages in some very populous areas. Restrictions of water use in many states have become virtually annual occurrences in years of low rainfall. Whereas the overall water supply in the United States is expected to meet demand in the foreseeable future, serious imbalances are expected to continue and worsen owing to geographic, seasonal, and annual variations in the supplies (4).

Two out of five U.S. cities have inadequate water supplies, and at least a quarter of the U.S. population faces serious water shortages. Yet, half the states, having two-thirds of the industry and over half of the population, have direct access to as much as they can draw of the approximately 3.3×10^8 km^3 of seawater. The solids content of this water, mainly salts, varies worldwide from 25,000 ppm (2.5%) in the Baltic Sea to over 45,000 ppm (4.5%) in some of the more confined gulfs of the Indian Ocean. The waters of the wide oceans are almost constant, at 35,000 ppm (3.5%). Also, many inland areas have access to large quantities of water too brackish to drink.

Increasing pollution of water sources worldwide is a major contributor to the climbing shortage of usable water. A recent report of the U.S. Environmental Protection Agency (2) points out that about 40% of the U.S. rivers, lakes, and estuaries surveyed are too polluted for even basic uses such as fishing or swimming. Especially grim is the condition of the Great Lakes, which contain one-fifth of the world's fresh surface water. About 97% of the Great Lakes waters were found to be substandard for designated uses, and particularly worrisome is the fact that the lakes are continuously polluted by toxic chemicals. Some of the good water in the U.S. is considered to be "threatened," and a projection of need for sewage treatment costing over $100 billion was made. In this context it is important to note the vital importance of water-quality legislation. Along with the Clean Air Act, the Clean Water Act of 1972 is of unprecedented value in U.S. environmental protection history. Launched to "restore and maintain the chemical, physical, and biological integrity of the Nation's waters," it has resulted in a significant slowdown of the pollution processes and in many cases reversed the trend. A World Bank report issued in 1995 stated that 80 countries with 40% of the world's population already have water shortages which could cripple agriculture and industry, and that about 95% of the world's urban areas dump raw sewage into rivers and other water bodies, with dirty water killing 10 million or more people every year and causing untold economical damage. The water supply and quality situation is worsening rapidly. It was estimated that the developing countries alone would need $\$600\text{–}800 \times 10^9$ for water projects over the next 10 years, with no identifiable available funding sources.

At some tourist destinations, to conserve available water supplies, some hotels might use double water systems, ie, seawater for flushing, but they still must provide on the order of 400–600 L/d per tourist to assure a comfort to guests unaccustomed to water shortages. Production of these quantities of water by desalination techniques has become an important expense to the hotels. Today, the cost of production of desalinated water at remote locations can be as high as $\$4.00–5.25/\text{m}^3$, and in some locations even higher.

The newly acquired wealth of the countries in the Middle East has given an unexpected boost in desalination for the region. Starting with nearly zero, Saudi Arabia, Kuwait, and the United Arab Emirates have over the past three decades added plants producing about 9×10^6 m^3/d of freshwater, representing nearly one-half the worldwide desalination capacity (7). This includes the world's largest single-site, 946,000-m^3 multistage flash distillation (MSF) complex, in Al-Jubail, Saudi Arabia; the world's largest (24,000 m^3/d each) MSF units in Abu Dhabi; and the world's largest hybrid desalination complex in Fujairah, UAE 454,000 m^3/day consisting of 284,000 m^3/day MSF and 170,000 m^3/day seawater RO plant. The current largest, stand alone, seawater RO desalination plant of permeate flow capacity of 330,000 m^3/day commenced operation by the end of 2005 in Ashkelon, Israel (8). Following the Arab Peninsula in order of desalination capacity are the United States, North Africa, the rest of the Middle East, the European Mediterranean area, Southeast Asia, the rest of Europe, and the Caribbean islands. Water desalination is increasingly used for the treatment of effluent waters for reuse, and of river and city waters to improve purity for various industrial applications such as boiler feed and ultrapure water for the electronics industry.

In many of the developing nations, lack of water hampers the profitable exploitation of material resources (9). Technological progress in the last decade has made readily available packaged small (10–400 m^3/d) seawater RO and vapor-compression systems. In fact, virtually all offshore drilling rigs today have their own seawater conversion plants. Desalination facilities, and in turn desalinated water, become readily available, contingent only on the availability of funds.

Even in highly industrialized countries disposal of treated municipal effluents is a serious problem. Cities on the lower reaches of a large river, with many cities above, use water that has been through sewers upstream many times. On the lower Mississippi, a water inventory indicates such reuse averages 14 times, with biochemical oxidation of the wastes during the flow between cities. The Rhine River, in passing through several countries, all of which drink from and dump into its waters, is subject to international problems of pollution. Of considerable import here is the fact that many so-called hard contaminants present in sewage increase almost proportionally with reuse of the water. Such materials do not ferment or oxidize under ordinary sewage treatments, and some are known carcinogens, which presents serious health problems to potable waters drawn from these bodies.

The U.S. government spent ca $\$150 \times 10^6$ in fiscal 1969 on water resources research relating to artificial rainmaking, soil conservation, waste treatment, desalting, public health, and planning research. This level of annual expenditure, large compared to the rest of the world, has in the meantime diminished (10), even though it was even then considered to be rather small for supporting a need estimated at $>\$100 \times 10^9$ for water facilities worldwide within 10–15 yr (11). Several government agencies conduct water research, primarily the U.S. Geological Survey, the Environmental Protection Agency, the Bureau of Reclamation, and the Department of Agriculture. Practically all of the reports and results of research on water in the United States are available from the National Technical Information Service, U.S. Department of Commerce, Springfield, VA 22161.

2.1. Potential for Saline Water Use in Municipal Distribution. Only a very small amount of potable water is actually taken by people or animals internally, therefore it seems quite uneconomical to desalinate all municipally piped water, although all distributed water must be clear and free of harmful bacteria. Most of the water piped to cities and industry is used for little more than to carry off small amounts of waste materials or waste heat. In principle in many locations, seawater can be used for most of this service. If chlorination is required, it can be accomplished by direct electrolysis of the dissolved salt (12). Arrayed against the obvious advantage of economy, there are several disadvantages: use of seawater requires different detergents; sewage treatment plants must be modified; the usual metal pipes, pumps, condensers, coolers, meters, and other equipment corrode more readily; chlorination could cause environmental pollution; and dual water systems must be built and maintained.

Pipes, valves, fittings, and almost all other components of small equipment are now available in plastic or ceramics, which do not corrode in salt water and are less expensive than the metals now used. Synthetic detergents are now available for use with seawater, although a final rinse with freshwater may be desired. Saltwater sewage can be treated successfully. Dual water systems using freshwater and seawater are already in use on ships and in many island resort hotels. Many of these also have seawater systems for firefighting. This trend could potentially grow in some new developments at costal locations. However, modifying the existing piping systems in costal cities into a dual water system, to enable conveying seawater, is to expensive and complicated to be practical in any foreseen future.

Some inland municipalities now distribute water with salt content exceeding 1000 ppm, water so brackish as to be unpleasant to the taste, even though it is distributed as potable water. Each home may produce or purchase the very small requirement of freshwater for drinking and cooking. Small membrane and ion-exchange desalinators are available; these produce the relatively few liters of potable water required (see MEMBRANE TECHNOLOGY; ION EXCHANGE). They can be purchased from and periodically serviced by service companies in many communities. The cost per liter of potable water produced in the home is several times as great as the cost in a central desalination plant, but the amount of desalinated water needed is only a small fraction of the total supply.

Home desalinators are possible only for industrialized countries with a central service organization. Use of purchased and rental water softeners is common in the United States. Also, use of under-sink RO units is quite common there. This is in spite that potable water quality in U.S. is very safe and strictly regulated in respect of contaminants concentration level and presence of pathogens.

Alternatively, small amounts of potable water may be delivered by truck to distribution centers or to tanks on house roofs. This system exists in Kuwait, which has many filling stations from which tank-truck operators buy water at \$1.00/m^3 for distribution at about \$3.00/m^3. Although much water is directly piped to residences in Kuwait today, 12% of the people still get their water by truck. In Khartoum, Sudan, families that buy from vendors, who deliver sacks of water by donkey, pay an average of \$16 per month.

2.2. Water in Industry. Freshwater used in industrial applications can often be replaced by saline or brackish water, usually after sedimentation, filtration, and chlorination (electrical or chemical), or other treatments (13). Another possibility, implemented already in California, is to supply industry with highly treated municipal effluent and return potable water allocation to general use. Extensive treatment of process water is not necessary for the largest user of water, the electric power industry, which in the United States passed through its heat exchangers in 2000 about 40% of the total supply of surface water, a quantity similar to that used for agriculture, and it was 48% of the combined fresh and saline water withdrawals (3). Single stations of 1000 MW may heat as much as 12 Mm^3/d by as much as 10–15°C. The water usage of power industry in U.S. has not changed significantly during the last decade (3).

The power-plant cooling water is either returned directly to its source, such as rivers, lakes, or oceans, or is recycled by circulation and consequent cooling in cooling towers. Cooling towers circulate the warmed water downward against a rising stream of air which removes heat by evaporating a part of the water to cool the balance. In the United States, cooling towers are not so common as in Europe because of the large bodies of water available.

Although 600 m^3 of water is used to make a metric ton of fertilizer, 150–240 m^3 to make a ton of steel, 480 m^3 to make a ton of gasoline, and 1000 m^3 to make a ton of acetate fiber, little if any of it is required chemically in any of these processes. Recycling can reduce industrial requirements by a factor of 10–50. Much of this water, particularly that for cooling, and often that for washing, can be saline. Some petroleum refiners have used salt water to remove heat (water's principal role in gasoline production), and some have actually produced table salt by evaporation in cooling towers.

The pulp and paper industry has tried for many years to use salt water for some of the 250–400 m^3 of water required to make a metric ton of paper (see PULP). Here, however, salt is disadvantageous to the chemical processes, either in pulping the lignocellulose or in the recovery of values from the black liquor after pulping, and can corrode expensive papermaking machinery. The possibility of recovering and reusing at least part of this water after membrane processing is under study.

The textile industry, also a large consumer of water, must always be located in areas with abundant water supply (see FIBERS). Before long, advances in purification of textile industry effluents are expected to free the industry of this limitation and enable textile plants to locate in virtually any area. The great fluid reservoir of heat, the alternative to water, fresh or salt, is the atmosphere. Increasingly, industry is using air coolers to dissipate the large quantities of heat from power plants, petroleum distillation, and other process use. This would reduce the use of cooling water correspondingly.

Air coolers almost invariably add considerably to plant cost, but they are competitive in operating cost based on direct once-through use of water that requires no treatment. If the alternative to air coolers is the use of water that requires substantial treatment or pumping costs, the air coolers will cost less to operate.

2.3. Water for Agriculture. Two liters of water in some form is the daily requirement of the average human, depending on many personal and external conditions. However, at least several hundred liters per day are required for the growing of vegetables, fruits, and grain that make up the absolute minimal daily food ration for a vegetarian. About 0.48 m^3 is required to produce an egg, and 31 m^3 to produce a kilogram of beef, based on the cereals required by the animals. The quality of water (in particular, its salinity) is of considerable importance to agriculture. There are crops that can grow in water of relatively high salinity (eg, sugar beets tolerate salinity of up to 5,000 ppm), but most crops cannot tolerate salinity exceeding 1000 ppm. A loaf of bread contains little of the more than a ton of water necessary to grow the wheat therein. The water content of vegetables comes from the 1000–2000 times their weight of it that is needed to grow them. Some of the main losses in agriculture may be reduced by agronomists and plant physiologists. Increasingly, the balance sheet for irrigation plant technology now makes use of agronomic WUE (water-use efficiency), which is a ratio of the amount of harvestable or economic biomass to the water consumed by evapotranspiration. Studies are being made to determine whether it is possible to supply less water to the roots, with better absorption there and smaller losses by transpiration through the leaves.

Waterproofing sandy soils to prevent drain-through has been successful in increasing crops as much as 400% with the same rainfall. A special plow lifts the soil to allow melted asphalt to be layered in overlapping impermeable strips 82 cm wide and 50 cm below the surface (9). Waterproofing the surface in Israel by compacting with chemicals increases runoff to basins or fields on slopes below. In many places, barren slopes have been coated with asphalt or concrete in layers as thin as 0.3 cm to catch rain, which is conducted to catch basins for irrigation or other uses. In St. Thomas, U.S. Virgin Islands, such catchment basins have, however, now been abandoned in favor of desalination.

The one-seventh of the world's crop lands that are irrigated produce one-quarter of the world's crops. Irrigation's main losses result directly from seepage and evaporation from the open water-carrying channels and the soil. Only a small fraction of the water withdrawn from the irrigation ditch or pipe is absorbed by the plants. Plastic films, as ground covers through which the plants protrude, prevent some losses, but at great expense for film and labor. Cheaper systems are necessary to assure better water utilization by plants. Other possible goals would be food plants with membranes capable of separating freshwater from brackish water, to give a non-salty crop. Progress has been made in both of these directions, and some plants have been developed that accumulate salt from the ground.

Vegetables and fruits such as melons are profitably grown in four or five crops per year by hydroponics, ie, the growth of plants in large, shallow concrete tanks containing no soil but gravel and water with added nutrients. Such installations can even be found in areas known for their water scarcity such as Aruba and St. Croix (U.S. Virgin Islands) where both desalinated seawater and brackish water are used for the purpose. Much water is still necessary per unit weight of crops, but the largest losses of ordinary irrigation are prevented, as indeed they must be because of the comparatively high cost of the water. Such concentrated agriculture is very expensive in preparation of land area, but economical for water and labor requirements. Production is high in the tropics, and hydroponics offers a major opportunity to many developing countries.

Other locations where desalinated water of either brackish or sea origin is used for agriculture are the Channel Island Guernsey, Israel, Libya, and countries in the Arabian Gulf including Saudi Arabia. The cost of such agricultural use is at best marginal and at worst exorbitant, depending on the amounts of water required.

3. Desalination: Manufactured Freshwater

Desalination has been used for providing drinking water on seafaring ships since ancient times (using solar or fuel heat), and an early reference to the scientific or miraculous conversion by Moses of bitter groundwater to fresh, viz, "... and the Lord showed him a wood and he put it into the water and the water became sweet," is made in the Old Testament. The possibility of producing freshwater from seawater or brackish water by separation of the salts opens a new dimension in the supply of freshwater. Areas bordering the sea would have an available raw material without limit or cost of transportation to the water facility. Figure 2 shows the remarkable growth of the cumulative contracted capacity of large desalination plants since 1959, when large-scale, land-based desalination began. As of the end of 2004, the world wide desalination capacity was 55.4 million m^3/ (7). This output can easily satisfy the domestic and public (excluding industrial and irrigation) water needs of at least 150 million people. About 60% of these plants are for desalting seawater, and 25% for desalting brackish water. About 75% of the desalted water is for municipal use, 20% for industrial use, and the remainder is used in the power industry, in discharge treatment, and by the military. There are more than 1700 companies and institutions involved in water desalination (14).

In many places, the need for desalination is even more urgent than the production of food, which is limited by water shortages. These shortages exist both

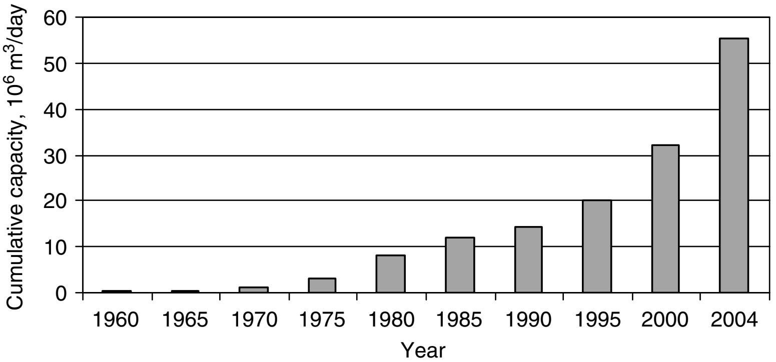


Fig. 2. History of cumulative capacity of land based desalination plants (7).

in the petroleum-rich countries and in many of the least-developed and poorest countries of the world. In the case of the former, the improved water supply obtained through desalination has already had stunning effects; by contrast, little hope for progress is on the horizon for the poor countries.

Water desalination is a process that separates the water from a saline water solution. The natural water cycle is the prevalent, and best, example of water desalination. Ocean waters evaporate as a result of solar heating and atmospheric influences, the vapor, consisting mostly of freshwater because of the negligible volatility of the salts at these temperatures, rises buoyantly and condenses into clouds in the cooler atmospheric regions, is transported across the sky by cloud motion, and is eventually deposited back on the earth's surface as freshwater rain, snow, and hail. The global freshwater supply from this natural cycle is ample, but, as mentioned before, many regions on earth do not receive an adequate share. Population growth, rapidly increasing demand for freshwater, and increasing contamination of the available natural freshwater resources render water desalination increasingly attractive.

Many ways are available for separating water from a saline water solution. The oldest desalination process is distillation. The evaporation of the solution is effected by the addition of heat or by lowering of its vapor pressure, and condensation of these vapors on a cold surface produces freshwater. The three dominant distillation processes are multistage flash (MSF), multieffect (ME), and vapor compression (VC). Until the early 1980s, the MSF process was prevalent for desalination. Now membrane processes, especially reverse osmosis (RO), are economical enough to have gained about half of the desalination market. In all membrane processes, separation occurs as a result of the selective nature of a membrane's permeability, which, under the influence of an external driving force, permits the preferential passage of either water or salt ions, but not both. The force driving the process may be pressure (as in RO), electric potential (as in electrodialysis, ED), or heat (as in membrane distillation, MD). A process used for low salinity solutions is the well-known Ion exchange (IE), in which salt ions are preferentially adsorbed onto a material that has the required selective adsorption property, and thus reduce the salinity of the water in the solution.

The cost of desalted water is composed of the capital cost of the plant, the cost of the energy needed for the process, and the costs of operation, maintenance staff, and supplies. In recent contracts for large seawater desalination plants the combined project cost is about \$1000/m^3-day. In brackish RO plants the combined cost of the project is about \$300–\$500/m^3-day. In large distillation seawater desalination plants the cost of water is about \$1.4–2/m^3. In recent large RO seawater desalination projects, the cost of desalted water is less than \$1/m^3. In large RO brackish water desalting plants total water cost below \$0.5/m^3 is achievable. A methodology for assessing the economic viability of desalination in comparison with other water supply methods is described in Reference 15. Desalination plants are relatively simple to operate, and progress toward advanced controls and automation is gradually reducing operator expenses.

3.1. Minimal Energy Requirements. The relative effect of the cost of the energy on the cost of the freshwater produced depends on local conditions, and is up to one-half of the total. In attempting to reduce this cost, it is of interest to determine the minimal energy amount thermodynamically needed for separating the water from the saline solution. The physical background to this will be introduced in a simple example. Because of the negligible volatility of the salts, the vapor of an aqueous saline solution is practically pure H_2O. It is well known that the vapor pressure of a saline water solution at a constant temperature decreases as the solution salinity increases. Thus, the vapor pressure of such a solution is lower than the vapor pressure of pure water at the same temperature. Withdrawal of this pure water (here in vapor form) from the saline solution to a pure water product storage vessel which is maintained at the same temperature thus requires pumping of the vapor from the vapor under lower pressure above the saline solution to the vapor under higher pressure above the pure water. The work needed for that vapor pumping, assuming a 100%-efficient pump (compressor) and perfectly insulated vessels, is the minimal energy needed for separating the water from the saline solution.

The effect of salt concentration on the vapor pressure of the solution has its well-known temperature exposition, called the boiling point elevation. Because the vapor pressure of saline solutions is lower than that of pure water, their boiling point temperature is higher. The difference between the boiling points is called the boiling point elevation, which rises with the concentration, and it can thus be seen as the measure of the extra energy needed for separation, ie, here the required raising of the temperature of the solution by this amount to attain boiling. Thermodynamically reversible separation defines the minimal energy requirement for that process. The minimal energy of separation $W_{\min}$ in such a process is the change in the Gibbs Free Energy between the beginning and end of the process, ΔG. The minimal work when the number of moles of the solution changes from n_1 to n_2 is thus

$$W_{\min} = \int_{n_1}^{n_2} (\Delta G)\, \mathrm{d}n_W \tag{1}$$

Bromley and co-workers (16) have calculated the minimal energy of separation of water from seawater containing 3.45 wt% salt, at 25°C, to be 2.55 kJ/(kg fresh water) for the case of zero fresh water recovery (infinitesimal concentration change) and 2.91 kJ/(kg fresh water) for the case of 25% fresh water recovery. $W_{\min}$ is, however, several fold smaller than the energy necessary for water desalination in practice.

Improved energy economy can be obtained when desalination plants are integrated with power generation plants (16,17). If the power generation part of such a dual-purpose plant is of the Rankine (steam) type, a back-pressure or extraction turbine is typically used to supply the low pressure and temperature steam as the heat source for the desalination plant. If the power plant is of the gas-turbine type, the turbine exhaust gas is typically used as the heat source for a boiler which provides heating steam for the desalination plant. Diesel engine reject heat, both from the engine coolant and the exhaust can also be used as the heat source for desalination. Dual-purpose plants lead to important energy savings, but also to increases in capital cost and complexity of operation. It is important to be able to deal effectively with possible transient mismatches in power and water demand. Practically all of the operating distillation-type desalination plants with a capacity above 1000 m^3/d are of the dual-purpose type.

3.2. The Major Desalination Processes. The major water desalination processes that are currently in use or in advanced research stages are described in this article. Information on detailed modeling can be found in the literature cited. The major texts on water desalination written since the 1980s are those by Spiegler and Laird (18), Khan (19), which contains many practical design aspects, Lior (20) on the measurements and control aspects, Heitman (21) on pretreatment and chemistry aspects, and Spiegler and El-Sayed (22) an overview primer. Extensive data sources are provided in References 17, 23–25.

4. Distillation Processes

4.1. Multistage Flash Evaporation (MSF). Almost all of the large desalination plants use the MSF process shown schematically in Figure 3. A photograph of a modern operating plant is shown in Figure 4. The seawater feed is preheated by internal heat recovery from condensing water vapor during passage through a series of stages, and then heated to its top temperature by steam generated by an external heat source. The hot seawater then flows as a horizontal free-surface stream through a series of stages, created by vertical walls which separate the vapor space of each stage from the others. These walls allow the vapor space of each stage to be maintained at a different pressure, which is gradually decreased along the flow path as a result of the gradually decreasing temperature in the condenser/seawater-preheater installed above the free stream. The seawater is superheated by a few °C relative to the vapor pressure in each stage it enters, and consequently evaporates in each stage along its flow path. The latent heat of the evaporation is supplied by equivalent reduction of the sensible heat of the evaporating water, resulting in a gradual lowering of the stream temperature. The evaporation is vigorous, causing intensive bubble generation and growth with accompanying stream turbulence, a process known as flash evaporation (23). One of the primary advantages of the MSF process is the fact that evaporation occurs from the saline water stream and not, as in other distillation processes such as submerged tube and multiple-effect evaporation, on heated surfaces, where evaporation typically causes scale deposition and thus gradual impairment of heat-transfer rates. Also, the fact that the sensible heat of water is much smaller than its latent heat of evaporation, the specific heat $c_p = 4.182$ kJ/kg per °C change of water temperature vs $h_{fg} = 2378$ kJ/kg, respectively, and that the top temperature is limited by considerations of scaling and corrosion, dictate the requirement for a very large flow rate of the evaporating stream. For example (in the following subscripts b, d, and s refer to distillate, brine, and steam, respectively) in operating between a typical top temperature $T_{b,t}$ of 90°C at the inlet to the evaporator and an exit temperature $T_{b,e}$ of 40°C corresponding to the ambient conditions, the overall temperature drop of the evaporating stream is 50°C. Using these values, the heat balance between the sensible heat of the water stream, flowing at a mass flow rate m_b, and the latent heat needed for generating water vapor (distillate)

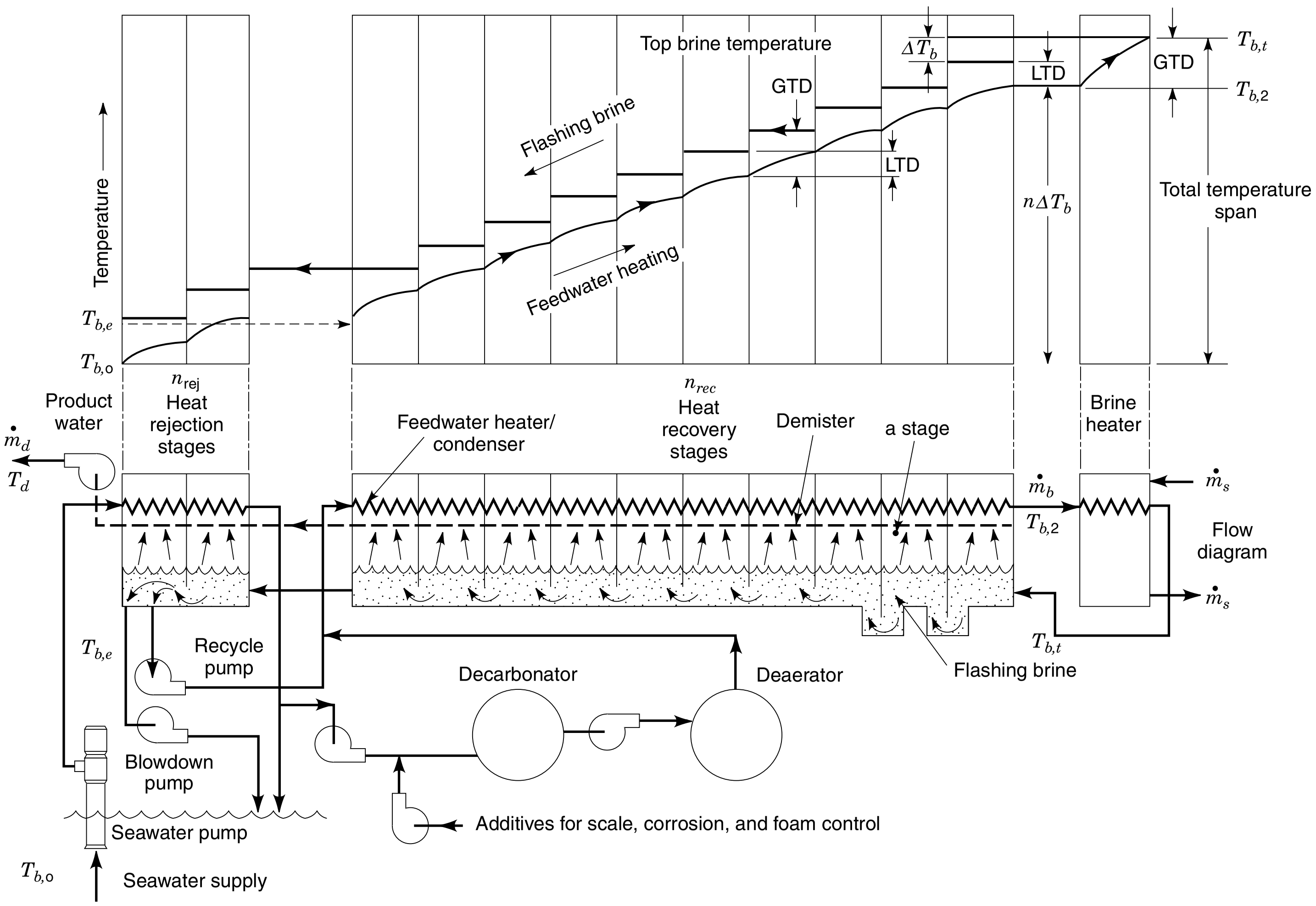


Fig. 3. Schematic flow and temperature diagram of the multistage flash (MSF) process for a recirculation type plant.

Fig. 4. The 341,000-m^3/d multistage flash (MSF) evaporation desalination plant Al Taweelah B in Abu Dhabi, United Arab Emirates. Courtesy of Italimpianti SpA. It is a dual-purpose plant, composed of six identical power and desalination units. The desalination units at 56,800 m^3/d each are currently the largest in the world. They have 17 recovery and 3 reject stages and a Performance Ratio of 8:1. The plant also produces 732 MWe of power.

at a mass flow rate m_d is

$$(\dot{m}_b - \dot{m}_d)c_p(T_{b,t} - T_{b,e}) \approx \dot{m}_d h_{fg} \tag{2}$$

which yields the brine-to-product mass flow ratio as

$$\frac{\dot{m}_b}{\dot{m}_d} = \frac{h_{fg}}{c_p(T_{b,t} - T_{b,e})} + 1 = \frac{2378}{(4.182)(50)} + 1 = 12.37 \tag{3}$$

Therefore, 12.37 kg saline water are needed in this case to produce 1 kg distillate. This high flow rate incurs corresponding pumping equipment and energy expenses, sluggish system dynamics, and, because the stream level depth is limited to about 0.3–0.5 m for best evaporation rates, also requires large evaporator vessels with their associated expense.

The generated water vapor rises through a screen (demister) placed to remove entrained saline water droplets. Rising further, it then condenses on the condenser tube bank, and internal heat recovery is achieved by transferring its heat of condensation to the seawater feed that is thus being preheated. This internal heat recovery is another of the primary advantages of the MSF process. The energy performance of distillation plants is often evaluated by the performance ratio, *PR*, typically defined as

$$PR \equiv \frac{\dot{m}_d}{\dot{m}_s} \tag{4}$$

where $\dot{m}_s$ is the mass flow rate of heating steam. Since the latent heat of evaporation is almost the same for the distillate and the heating steam, *PR* is also the ratio of the heat energy needed for producing one unit mass of product (distillate) to the external heat actually used for that purpose. Most of the heating of the brine stream to the top temperature $T_{b,t}$ is by internal heat recovery, and as seen in Figure 3, the external heat input is only the amount of heat needed to elevate the temperature of the preheated brine from its exit from the hottest stage at $T_{b,2}$ to $T_{b,t}$.

Assuming for simplification that the temperature drop of the flashing brine, ΔT_b, is the same in each stage and that the specific and latent heat of the brine remains the same throughout the plant, the relationship between the number of stages, n, and the performance ratio is expressed by

$$PR = \frac{1}{\dfrac{LTD}{T_{b,t} - T_{b,e}} + \dfrac{1}{n}} \tag{5}$$

where LTD is the lowest temperature difference between the flashed vapor and the heated feedwater, in each stage (Fig. 3). Equation 5 shows that increasing the number of stages increases the PR. This implies that more heat is then recovered internally, which would thus require a larger condenser/brine-preheater heat-transfer area. The required heat transfer area, A, per unit mass of distillate produced for the entire heat recovery section (composed of n_{rec} stages), and taking average values of the overall vapor-to-feedwater heat transfer coefficient U and $LMTD$ per stage, is

$$A = \frac{h_{b,fg}}{U(LMTD)} \tag{6}$$

where $h_{b,fg}$ is the average latent heat of evaporation of the flashing brine, $LMTD$, the log-mean temperature difference between the vapor condensing on the tubes and the heated brine flowing inside the tubes, for an average stage defined as

$$LMTD = \frac{GTD - LTD}{\ln \dfrac{GTD}{LTD}} = \frac{(T_{b,t} - T_{b,2}) - LTD}{\ln\left(\dfrac{T_{b,t} - T_{b,2}}{LTD}\right)} \tag{7}$$

where GTD is the greatest temperature difference between the flashing brine and the brine heated in the condenser. The size of the heat transfer area per unit mass of distillate produced by the plant is

$$A = \frac{h_{fg,b}}{U} \frac{n_{\text{rec}}}{(T_{b,t} - T_{b,e})} \ln\left(\frac{n_{\text{rec}}}{n_{\text{rec}} - PR}\right) \tag{8}$$

Examination of this equation will show that the required heat transfer area for the heat recovery section per unit mass of distillate produced, A, increases significantly when PR is increased, and decreases slightly as the number of heat recovery stage, n_{rec}, is increased.

The MSF plant shown in Figure 4 is of the recirculation type, where not all of the brine stream emerging from the last evaporation stage is discharged from the plant, as it would have been in a once-through type of plant. A fraction of the emerging brine is mixed with pretreated seawater and recirculated into the condenser of the heat-recovery section of the plant. Because only a fraction of the entire stream in this configuration is new seawater, which needs to be pretreated (removal of air and CO_2, ie, deaeration and decarbonation, and the addition of chemicals that reduce scale deposition, corrosion and foaming), the overall process cost is reduced. The recirculation plant is also easier to control than the once-through type.

Whereas most of the energy exchange in the plant is internal, steady-state operation requires that energy in an amount equal to all external energy input also be discharged from the plant. Consequently, the heat supplied in the brine heater plus any pumping energy is discharged in the heat-rejection stages section of the plant. Assuming, for purposes of estimation, an equal temperature drop in each stage and that the pumping energy can be neglected relative to the heat input in the brine heater indicates that the ratio of the number of the heat-recovery to heat-rejection stages is approximately equal to the performance ratio *PR*.

Further detail about MSF desalination can be found in References 18, 26, 27. A detailed design of an MSF plant producing 9,500 m^3/d gallons of fresh water per day was published by the U.S. government (28).

4.2. Multi-Effect Distillation (ME). The principle of the multi-effect (ME) distillation process is that the latent heat of condensation of the vapor generated in one effect is used to generate vapor in the next effect, thus obtaining internal heat recovery and good energy efficiency. Such plants have been used for many years in the salt, sugar, and other process industries. Several ME plant configurations, most prominently the horizontal tube multi-effect (HTME) (Fig. 5) and the vertical tube evaporator (VTE), shown schematically

Fig. 5. A horizontal-tube multi-effect (HTME) desalination unit, producing 5000 m^3/d in St. Croix, U.S. Virgin Islands. Courtesy of I.D.E. Technologies Ltd.

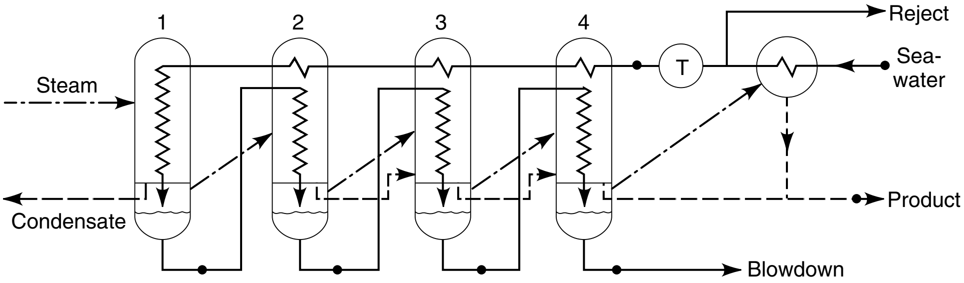


Fig. 6. Simplified schematic flow diagram of a typical 4-effect vertical-tube multi-effect (VTE) desalination plant, where (—) represents brine, (– **u**–) represents vapor, (– – – –) represents condensate, Ⓣ denotes pretreatment, elements enclosed by a cylinder-like shape constitute an effect, and $\mathbf{u}_\mathrm{s}$ represent pumps.

in Figure 6, are in use. In the HTME, vapor is circulated through a horizontal tube bundle, which is subjected to an external spray of somewhat colder saline water. The vapor flowing in these spray-cooled tubes condenses, and the latent heat of condensation is transferred through the tube wall to the saline water spray striking the exterior of the tube, causing it to evaporate. The vapor generated thereby flows into the tubes in the next effect, and the process is repeated from effect to effect. A low temperature multi-effect distillation plant developed to operate at an upper limit of 70°C, illustrated in Figure 5, has a low fuel consumption, estimated by the manufacturer at 2.4–2.8 kg fuel per 1000 m^3, and lower capital cost because the whole plant is built of aluminum (29). Many such plants have been installed and operated successfully.

In the vertical tube multi-effect evaporator (VTE), the saline water typically flows downward inside vertical tubes, and evaporates as a result of condensation of vapor coming from a higher temperature effect, on the tube exterior. While internal heat recovery is a feature common to both MSF and ME processes, there are at least three important differences between them. One is that evaporation in the ME process occurs on the heat-transfer surfaces (tubes), whereas in the MSF process it takes place in the free stream. This makes the ME process much more susceptible to scale formation. At the same time, the heat-transfer coefficient between the vapor and the preheated brine is lower in the MSF process because the heated brine does not boil. In the ME process it does boil, and it is well known that boiling heat-transfer coefficients are significantly higher than those where the heating does not result in boiling. In using direct transfer of latent heat of condensation to latent heat of evaporation, instead of sensible heat reduction to latent heat of evaporation as in MSF, the ME process requires a much smaller brine flow than the MSF. Limiting brine concentration in the last effect to about three times that of the entering seawater, for example, requires a brine flow of only about 1.5 times that of the distillate produced. At the same time, a pump (although much smaller than the two pumps needed in MSF) is needed for each effect.

The performance ratio of ME plants is just slightly lower than the number of effects, which is determined as an optimized compromise between energy efficiency and capital cost. Six effects are typical, although plants with as many as 18 effects have been built. Further detail about ME desalination can be found in References 27, 30.

4.3. Vapor Compression Distillation (VC). The vapor pressure of saline water is lower than that of pure water at the same temperature, the pressure difference being proportional to the boiling point elevation of the saline water. Desalination is attained by evaporating the saline water and condensing the vapor on top of the pure water. Therefore, the pressure of the saline water vapor must be raised by the magnitude of that pressure difference, plus some additional amount to compensate for various losses. This is the principle governing the vapor compression desalination method. Moreover, as shown in the VC plant flow diagram in Figure 7, the heat of condensation of the compressed vapor is recovered internally by using it to evaporate the saline water. Additional heat recovery is obtained by transferring heat from the concentrated brine effluent and the produced fresh water (which need to be cooled down to as close to ambient conditions as possible anyway) to the feed saline water, which is thus preheated. The schematic flow diagram in Figure 7 shows a design in which the preheated seawater is sprayed onto a bank of horizontal tubes carrying condensing compressed vapor at a temperature higher than that of the seawater. The spray thus evaporates on contact with the exterior of the tube and provides the cooling needed for the internal condensation. Considering the fact that the energy required for vapor compression over a typical overall temperature difference of 4°C and a vapor compressor efficiency of 0.8 is 34 kJ/kg (easily calculated from an enthalpy balance), and that the latent heat of condensation is about 2400 kJ/kg, one can see that a small amount of compression energy enables a large amount of heat to be used internally for desalination. One can thus envisage the VC plant as a large flywheel, wheeling a large amount of energy around at the expense of a small amount needed for sustaining its motion.

The compressor can be driven by electric motors, gas or steam turbines, or internal combustion (usually diesel) engines. The compressor can also be a steam-driven ejector (Fig. 7**b**), which improves plant reliability because of its simplicity and absence of moving parts, but also reduces its efficiency because an ejector is less efficient than a mechanical compressor. In all of the thermally driven devices, turbines, engines, and the ejector mentioned herein, the exhaust heat can be used for process efficiency improvement, or for desalination by an additional distillation plant. Figure 8 shows a flow diagram of the vertical-tube vapor compression process.

Figure 9 shows a multi-effect VC plant. Using more than a single effect reduces the vapor volume that needs to be compressed. Moreover, the overall required heat-transfer area is also decreased because much of the single-phase heat-transfer process in the preheater of the single-effect plant is replaced by the high heat-transfer condensation–evaporation processes in the effects. Although the multi-effect feature also increases the required compression ratio, the cost of produced water is reduced overall. An operating two-stage 700-m^3/d VC brine concentration plant is shown in Figure 10.

VC plants are of small (<4,000 m^3/d) capacity and relatively high energy efficiency, the latter depending on the number of effects incorporated, eg, a two-effect plant had an energy use of 57 kJ per kg water produced. Further detail about VC desalination can be found in References 18, 26, 27, 30.

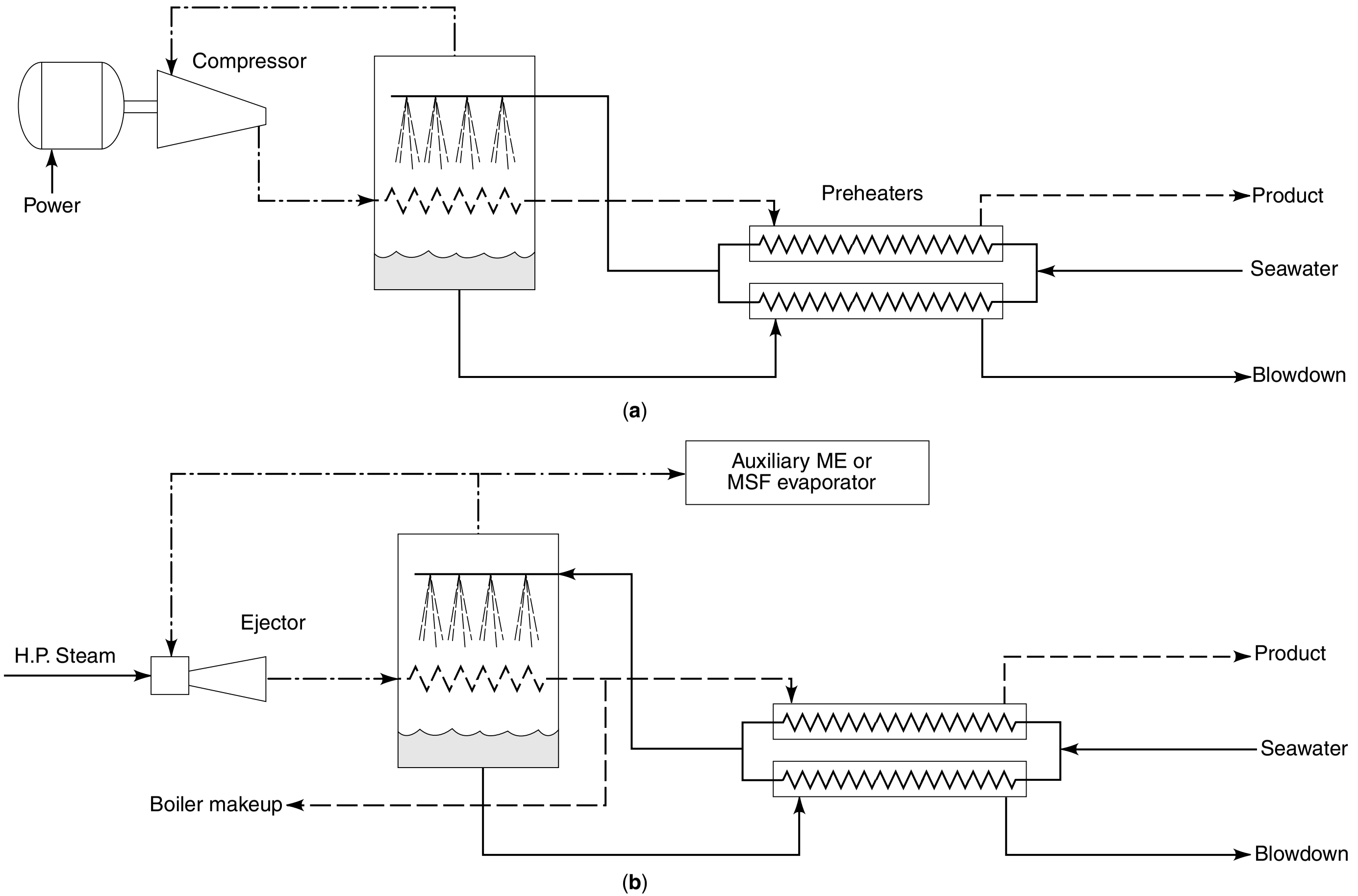


Fig. 7. Schematic flow diagram of a basic horizontal-tube vapor compression (VC) desalination plant, shown (**a**) with a mechanical, motor-driven compressor and (**b**) with a thermocompressor, using an ejector, where (—·) represents vapor; (—), brine; and (– – – –), product.

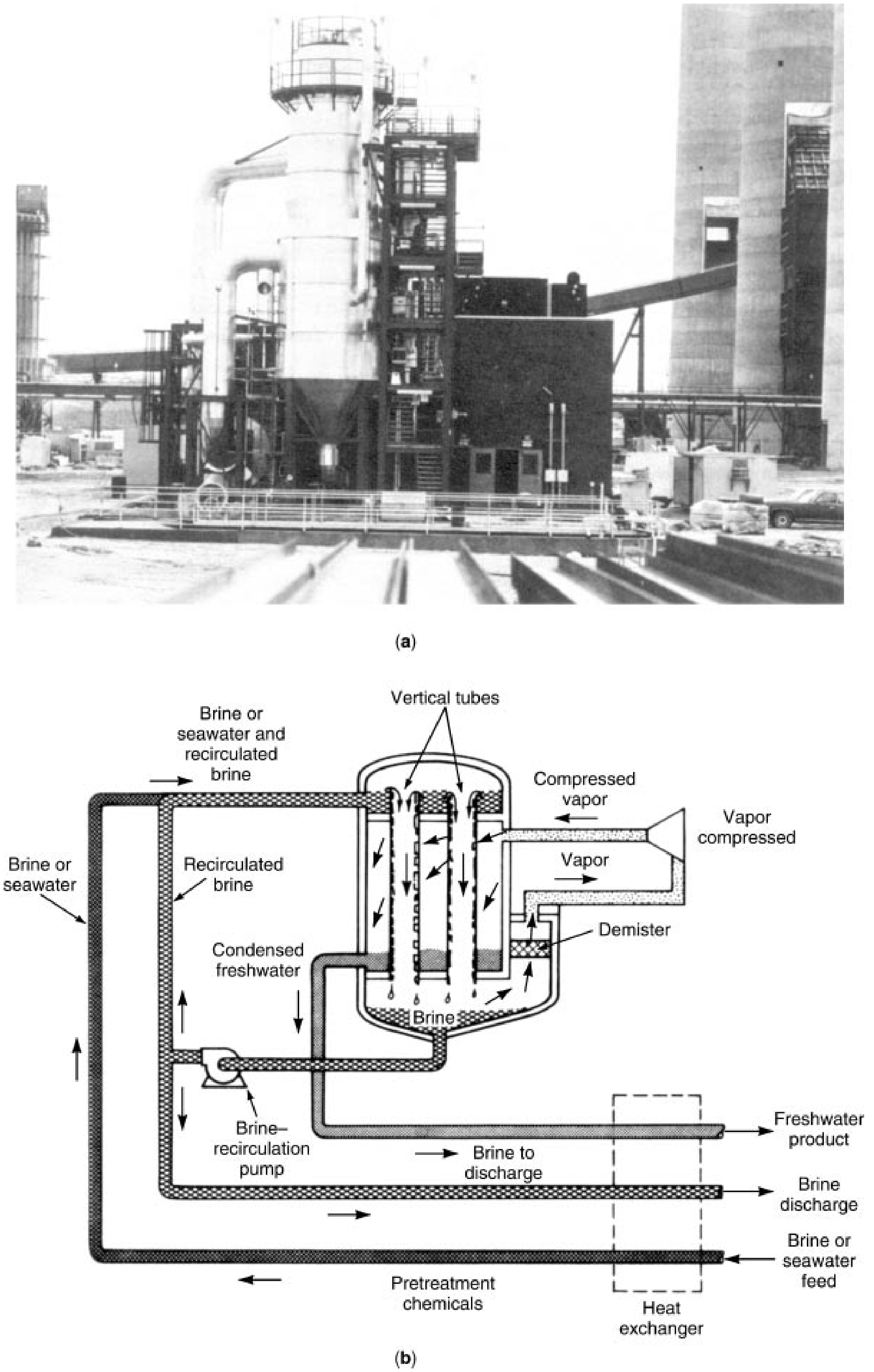


Fig. 8. Pictorial view (**a**) and flow diagram (**b**) for a vertical-tube vapor-compression process. Courtesy of Resources Conservation Co.

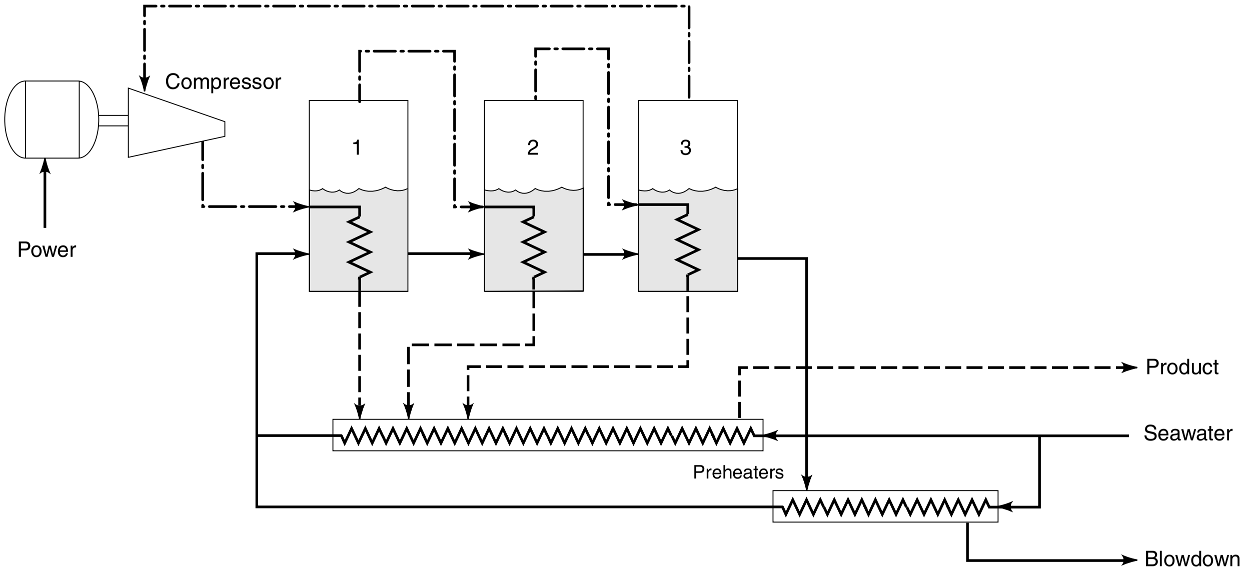


Fig. 9. Schematic flow diagram of a multi-effect vapor-compression submerged-tube desalination plant with three effects, where (—·) represents vapor; (—), brine; and (– – – –), product.

Fig. 10. A two-stage 700-m^3/d VC brine concentration plant. (Courtesy of I.D.E. Technologies Ltd.)

4.4. Freeze-Desalination. It is rather well known that freezing of saline water solutions is an effective separation process in that it generates ice crystals which are essentially salt-free, surrounded by a more salt-concentrated solution. This phenomenon has significant underlying appeal for its use as a water desalination process: (*1*) compared with distillation, which requires sensible heating of the saline water from an ambient temperature of, eg, 25°C to about 100°C, a 75°C rise, freezing needs cooling by only about 25°C to the freezing point; (*2*) in distillation, a latent heat investment of about 2400 kJ/kg is needed for generating the pure water vapor, whereas the latent heat of freezing is only about 333 kJ/kg, almost 8-fold smaller; (*3*) operation at freezing temperatures reduces the problems of corrosion and scaling significantly; and (*4*) the low temperature also allows the use of less expensive construction materials.

In principle, freeze-desalination is perhaps the most energy-efficient desalination process, but it has not yet (ca 2005) reached any significant commercial introduction for several reasons, such as the difficulty in developing efficient and economical compressors for vapor having the extremely high specific volume at the low process pressure, and difficulties in maintaining the vacuum system leak-free and in effecting reliable washing of the ice crystals. A review of freeze desalination processes is given in Reference 31.

4.5. Materials and Scaling Issues in Distillation Processes. Two aspects of the basically simple desalination process require special attention. One is the high corrosivity of seawater, especially pronounced in the higher temperature distillation processes, which requires the use of corrosion-resistant, and therefore expensive, materials. Typical materials in use are copper–nickel alloys, stainless steel, titanium, and, at lower temperatures, fiber-reinforced polymers and special concrete compositions (32). The other aspect is scale formation (33,34), discussed in more detail below.

Obtaining maximum performance from a seawater distillation unit requires minimizing the detrimental effects of *scale* formation. The term scale describes deposits of calcium carbonate, magnesium hydroxide, or calcium sulfate that can form in the brine heater and the heat-recovery condensers. The carbonates and the hydroxide are conventionally called alkaline scales, and the sulfate, nonalkaline scale. The presence of bicarbonate, carbonate, and hydroxide ions, the total concentration of which is referred to as the alkalinity of the seawater, leads to the alkaline scale formation. In seawater, the bicarbonate ions decompose to carbonate and hydroxide ions, giving most of the alkalinity (eq. 12).

$$\begin{aligned} 2\,HCO_3^- &\rightleftharpoons CO_3^{2-} + CO_2 + H_2O \\ H_2O + CO_3^{2-} &\rightleftharpoons 2\,OH^- + CO_2 \end{aligned} \tag{12}$$

The kinetics of the formation of the magnesium hydroxide and calcium carbonate are functions of the concentration of the bicarbonate ions, the temperature, and the rate of release of CO_2 from the solution. At temperatures up to 82°C, $CaCO_3$ predominates, but as the temperature exceeds 93°C, $Mg(OH)_2$ becomes the principal scale. Thus, in seawater, there is a considerable tendency for surfaces to scale with an increase in temperature.

The interrelationship of nonalkaline scales ($CaSO_4$, $CaSO_4{\cdot}1/2\,H_2O$, $CaSO_4{\cdot}2\,H_2O$) depends on temperature and the concentration of $CaSO_4$. To assure that no hemihydrate scale forms, MSF operators must run their plants in such a manner as to assure that the concentration of the total dissolved solids does not exceed 70,000 ppm at temperatures of 120°C. With average-salinity seawater, plants can operate at a concentration factor of 2, but in the Middle East where water salinity can be as high as 50,000 ppm, the concentration factor should not exceed 1.4. Under no circumstances should the total dissolved solids exceed 70,000 ppm, ie, twice the concentration of normal seawater at 120°C.

A number of options for controlling scale formation are used in plant operations around the world. The most common methods are reduction of feed pH by addition of acid or use of scale controlling polymers as a feed water additives. In the feed pH reduction approach, the hydrogen ion from added acid decomposes the bicarbonate ions.

$$HCO_3^- + H^+ \longrightarrow CO_2\uparrow + H_2O \tag{13}$$

About 120 ppm of sulfuric acid must be provided for normal seawater. Control of acid dosing is critical; the amount of acid must be stoichiometric to the alkalinity expressed as $CaCO_3$. In conjunction with acid dosing, the CO_2 formed must be removed and some sodium hydroxide added to maintain ca pH 8 in the system.

Alternatively, less than-stoichiometric amounts of acid can be added to retain some alkalinity in the untreated feed; in either case, CO_2 removal is done with packed columns. Acid-dosed feed is passed through a column with air flow that sweeps the CO_2 from the feed saturated with carbon dioxide. This is usually followed by a deaeration, during which both the air and CO_2 are reduced to the levels needed to minimize, if not eliminate, corrosion. Although acid-dosing does permit higher operating temperatures, it often has adverse effect on plant life.

The second option involves use of a new family of polymers, the so-called high temperature scale-control chemicals. These are compounds that, added in 3–8 ppm, lead to lattice distortion and the formation of a nonadhering scale. Belgard (CIBA-GEIGY) was the first compound exemplifying this type of MSF operation, which is now steadily displacing acid in the operation of MSF plants around the world, with important contribution to plant life (21) (see DISPERSANTS).

A new approach being introduced recently is application of nanofitration technology to soften feed water to the MSF unit (35,36). This approach, known also as a “hybrid technology” nanofltration system processes makeup water, reducing its hardness. Specially developed nanofiltration membranes have high rejection of hardness and sulfate ions and high passage of monovalent ions. Due to this selective permeability, the osmotic pressure difference between feed and permeate is low and the nanofiltration system can operate at low feed pressure. Reduction of hardness in the makeup water enables operation of MSF unit at higher “top” temperature and allows increase unit capacity. This hybrid process is still in the stage if initial commercial introduction and data regarding process reliability and economics is not available.

5. Membrane Desalination Processes

The reverse osmosis (RO) technology is at present the most versatile desalination method. It is effective in treating water of any salinity, staring with low salinity brackish water up to high salinity seawater. RO units are applied to produce ultrapure water for semiconductor industry and potable water from high salinity seawater in a cost effective way. It is also the process with the shortest history among desalination methods, tracking its beginning to scientific experiment performance at University of Florida at 1950’s conducted by Reid and Breton (37,38). Today, RO commercial based systems account for more than 50% of all desalination plants worldwide. The reverse osmosis process was derived from direct osmosis, the major mass transport phenomenon across living cells. The driving force for this water flux is the thermodynamic potential difference, proportional to the concentration difference between the two sides of the membrane, and is exhibited as the so-called osmotic pressure. For typical seawater at 25°C, the osmotic pressure is higher by 2.51 MPa on the freshwater side of the membrane. If a pressure higher than the osmotic pressure is applied to the concentrated solution side of the membrane, the water will move across the membrane in the reverse direction, from the saline solution side to the pure water one. This process is called reverse osmosis (and sometimes hyperfiltration), and is the basic principle underlying reverse-osmosis desalination.

The basic approximate equation for the separation process gives the water flux, $\dot{m}_w''$ (kg/m^2/s) across an RO membrane as

$$\dot{m}_w'' = K_{pe}K_{cf}[(P_f - P_p) - (\pi_f - \pi_p)] \tag{14}$$

where K_{pe} = the water permeability constant of the membrane (in kg/m^2/s · Pa), typically increasing strongly as the temperature rises (about 3% per °C). As membrane desalination plants are design to operate at const permeate flow, with increase of feed water temperature the feed pressure is reduced, P = water or saline solution pressure (in Pa); π = osmotic pressure (in Pa); and the subscripts f and p pertain to the saline feed water and to the desalted product water, respectively.

An approximation for this salt flow is

$$\dot{m}_s = KK_s(C_{fm} - C_p) \tag{15}$$

where $\dot{m}_s$ = salt mass transfer rate across the membrane, kg/s; K = a proportionality constant, dimensionless; K_s = the salt permeation constant, kg/s, which increases with pressure and temperature.

The salinity of the product water, C_p, can be estimated by the formula

$$C_p = K_{cp}(1 - \eta)\bar{C} \tag{16}$$

where K_{cp} = the concentration polarization coefficient, $\equiv C_{fm}/C$, a measure of the increase of the feedwater salinity at the membrane wall beyond that of the bulk solution; C_{fm} = the salt concentration at the membrane wall; C = the bulk salinity of the saline water feed, $\approx (C_f + C_r)/2$, where C_r = the salt concentration of the reject brine; and η = the salt rejection factor, $\equiv$(amount of salts rejected by the membrane)/(amount of salts in the brine feed).

The pressure to be used for reverse osmosis depends on the salinity and temperature of the feedwater, the type of membrane, the design recovery rate (RR) and permeate flux. It ranges from about <1.0 MPa for low feed concentrations softening (nanofiltration) systems through 1.5–2.5 MPa for brackish waters, and to 5.5–8.0 MPa for seawater desalination plants.

$$RR = \frac{\dot{m}_p}{\dot{m}_f} \cong 1 - \frac{C_f}{C_r} \tag{17}$$

The prevalent membrane configurations used in RO plants is the spiral-wound type. The hollow fiber configuration, used frequently in the past for seawater desalting, is in limited use today, due to higher susceptibility of membrane modules in the hollow fiber configuration to fouling and limited commercial offering. The basic spiral-wound-type module (Fig. 11**a**) is made of two sheets placed upon each other and rolled together in a spiral around a cylindrical perforated tube. One of the sheets is in the form of a sandwich typically composed of five layers bonded together along three edges. The two outer layers are the semipermeable membranes. Each of them is backed by a porous material layer for

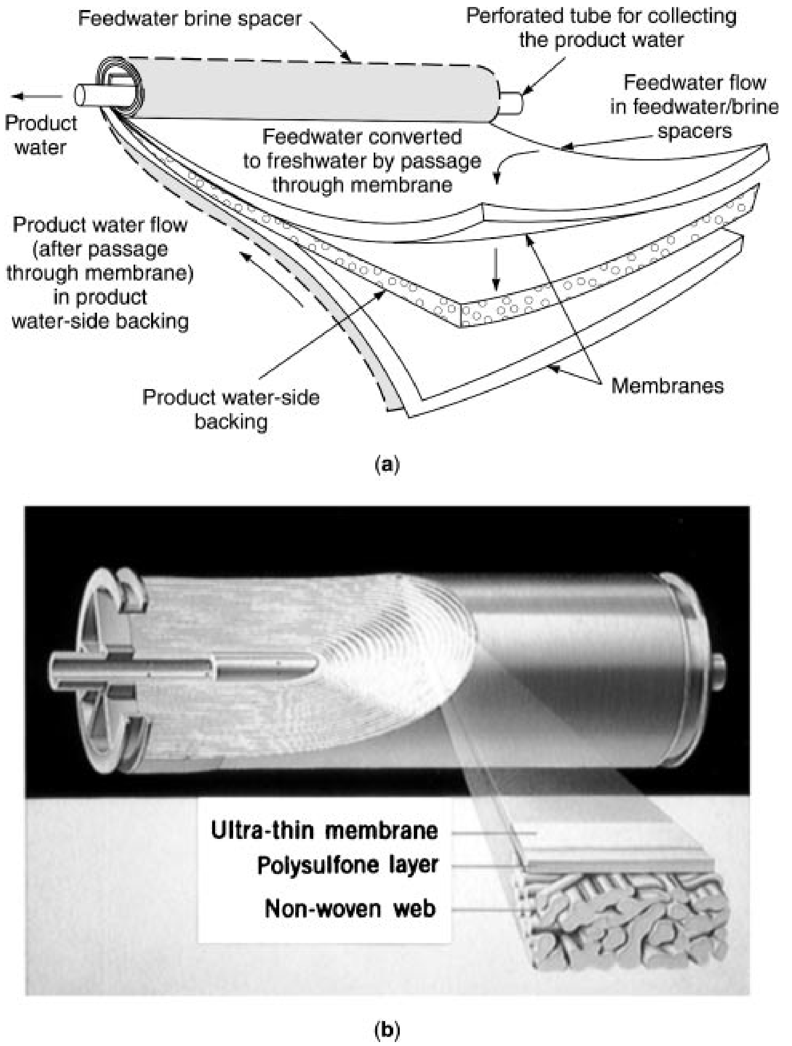


Fig. 11. A spiral-wound reverse osmosis membrane element: (**a**) schematic depiction; (**b**) cross section of a spiral-wound thin-film composite RO Filmtec membrane element (40% conversion). Courtesy of Dow Chemical Co.

mechanical strength, and the central layer is a thicker porous material layer that collects and transports the produced fresh water. The second sheet is a porous mesh spacer through which the high-pressure saline water feed is passed in an axial direction. Product water separates from the saline solution and permeates through the two adjacent semipermeable membranes into the central product-water carrying layer, which conducts it spirally to the unbonded edge of the sandwich to the central perforated tube. The semipermeable membranes are typically made from composites of polyamide polymers on polysulfone support membranes. Figure 11**b** shows a version of a contemporary commercial element of that type.

Table 1. **Nominal Performance and Test Conditions of Representative RO Elements**[a]

Element type	Nanofiltration	Brackish	Seawater
element model	ESNA–LF	ESPA2+	SWC5
membrane area, m2	37.2	40.8	37.2
nominal permeate flow, m^3/day	31.0	45.4	34.0
nominal salt rejection, %	89%	99.6	99.8
nominal test condition			
feed temperature, °C	25	25	25
feed salinity, mg/L	500	1500	35000
feed pressure, mPa	0.52	1.0	5.5
recovery rate, %	15	15	10

[a] Courtesy of Hydranautics Co.

Size of commercial spiral wound membrane elements range from 5 cm in diameter, 25 cm long for undersink application to 20 cm diameter, 100 cm long, used in a large desalination systems. Currently even larger; 40 cm in diameter, 100 cm long membrane elements are being introduced. Membrane elements made by different manufacturers have similar dimensions and configuration and are interchangable. The standard commercial 20 cm diameter by 100 cm long elements have about 40 m2 of membrane area. At field conditions such element will produce, on the average, 25 m^3/day permeate in brackish or softening applications. The same size element used in seawater application will have permeate capacity of about 13 m^3/day. Commercial RO membrane elements belong to one of three basic categories: softening (nanofiltraion), brackish or seawater. Membrane elements in each category are tested at different test conditions, which are uniform across the membrane industry. Table 1 provides representative nominal performance and ex-factory test conditions of elements in each category.

In RO systems membrane elements are operating in pressure vessels (Fig. 12). In large systems a pressure vessel contains between 6 to 8 elements connected in series.

Pressure vessel containing 6 to 8 elements can operate at recovery rate of 40–70%. Pressure vessel has tree ports: feed port, concentrate port and the permeate port. Pressure vessels are connected hydraulically in parallel and form a stage. Number of stages in RO unit ranges from 1 to 3, depending on recovery rate. In a multisatge unit concentrate effluent from one stage flows as a feed water to the next one. The number of pressure vessels in subsequent stages is usually close to the 2:1 ratio. This is to maintain similar rate of feed flow in the subsequent stages. Schematic configuration of a two stage system is shown in Figure 13. Figure 14 is a photograph of a RO train in two stage configuration.

Seawater systems operate at recovery rate of 40–50%, therefore RO trains are usually configured as a single stage array of pressure vessels. Recovery rate in brackish systems ranges from 65–85%. For this recovery rate range, two stage unit configuration is most common. Nanofiltrtaion systems are sometimes design to operate at recovery rate as high as 90%. At the high end of the recovery rate range a three stage configuration array is used. In seawater systems the recovery rate is limited by osmotic pressure. Higher recovery results in higher osmotic

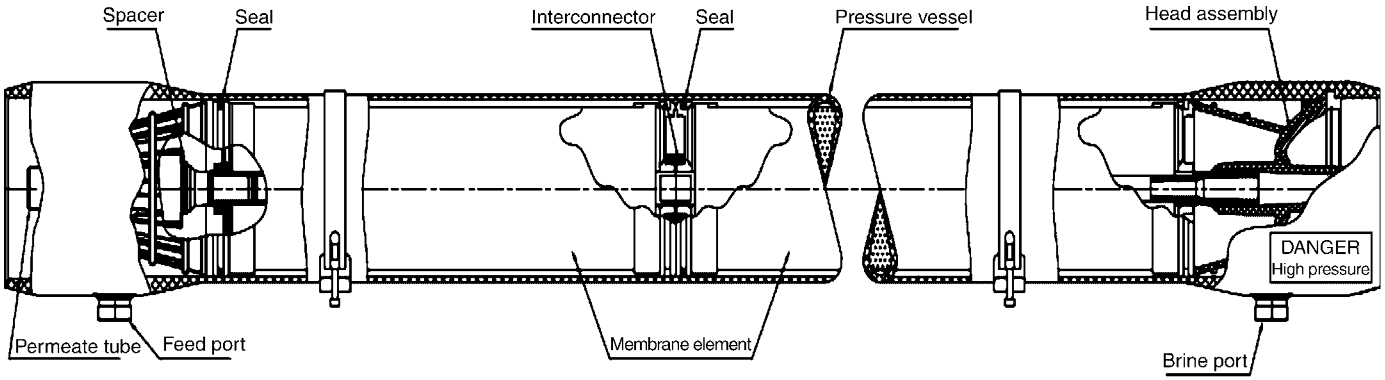


Fig. 12. Configuration of pressure vessel with membrane elements. (Courtesy of Bell Industries.)

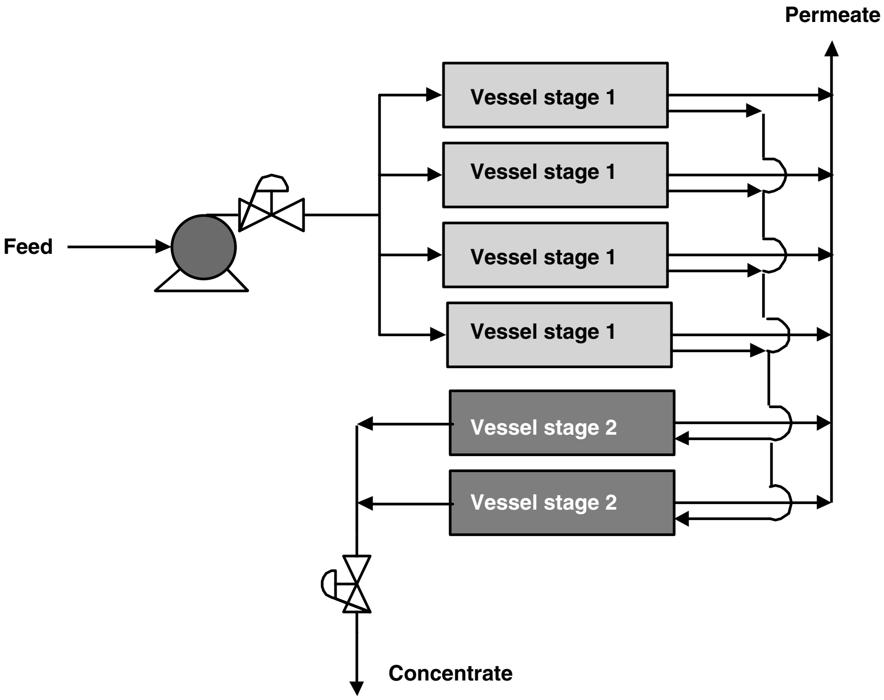


Fig. 13. Configuration of two stage RO system.

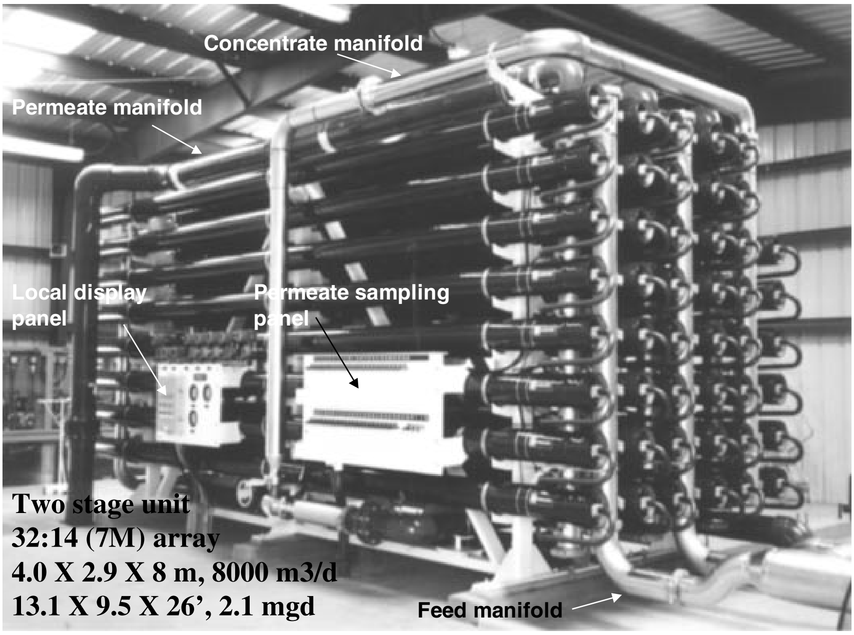


Fig. 14. Two stage RO train. (Courtesy Hydranautics.)

Table 2. **Osmotic Pressure of Feed and Concentrate Streams in Seawater RO Unit**

RO stream	Seawater salinity 35,000 mg/L	Seawater salinity 40,000 mg/L	Seawater salinity 45,000 mg/L
feed water	26 bar	30 bar	34 bar
concentrate at 50% recovery	51 bar	60 bar	67 bar
concentrate at 60% recovery	64 bar	74 bar	83 bar
concentrate at 70% recovery	85 bar	98 bar	111 bar

pressure and therefore in higher feed pressure requirements. Table 2 shows osmotic pressures of feed and RO concentrate for common seawater feed salinities.

In order to produce meaningful permeate flow also from membrane elements located at the concentrate end of the seawater system, the feed pressure has to be about 5 bars higher then the osmotic pressure of the concentrate. At the prevailing electricity cost and efficiency of pumping systems, feed pressure above 70 bar would result in an excessive power cost contribution to the water cost and therefore the recovery rate of seawater RO system is being designed accordingly.

In brackish and nanofiltration systems the tendency is to maximize recovery rate due to limited availability of feed water and limitations with disposal of concentrate. There the recovery rate is limited by the concentration of sparingly soluble salts that can form scale in the tail position membrane elements. The major scaling constituents of concern in brackish systems are calcium carbonate, calcium sulfate, and silica. Very effective scale inhibitors have been developed recently to control scale formation. At a dosing rate of less than 5 ppm to the feed water, scale inhibitors prevent scale formation and enable stable operation of RO systems at recovery rates that result in significant level of oversaturation of scale forming salts in the concentrate stream. Still, potential for scaling is a major concern in determination of recovery rate of brackish and nanofiltration RO systems.

5.1. Reverse Osmosis Applications. RO applications are usually categorized based on the source of feed water and its salinity. Nanofiltration systems treat water of close to potable salinity 300–1000 ppm TDS. The objective of operation of nanofiltration systems is to reduce hardness and constituents that affect tests and esthetic of potable water: color, iron, organic matter. Brackish RO systems treat water of higher salinity: 1000–10000 and the objective is mainly to remove dissolved mineral constituents that are in concentrations above potable water limits. These usually include common ions, such as Cl, Ca, Mg and SO_4. Sometimes the main objective is to remove contaminants that were introduced to aquifer through human activity: nitrates from agricultural run off or toxic compounds from industrial discharge. Nanofiltration and the brackish RO system treat mainly well water.

Wastewater RO reclamation systems treat filtrated secondary effluent, which is usually in the low salinity range, similar to one encountered in the nanofiltration applications. The objective of membrane reclamation system is to reduce salinity of the treated wastewater to the level acceptable for irrigation. Sometimes much lower salinity level of the product water is require if reclaimed product water is intend to be used for industrial applications such as boiler feed make up or process water.

Seawater RO desalination systems treat seawater of high salinity, usually ranging from 34,000 ppm TDS to 47,000 ppm TDS. The objective is produce permeate of potable salinity, usually below 400 ppm TDS. Sometimes additional requirements of permeate quality are imposed, such as low concentration of boron and/or bromides. Some small RO systems treat seawater that originates from beach wells. However, waste majority of the systems treat seawater water supplied from submersible intakes or other surface pumping stations.

The quality and composition of raw water from each of the above mentioned sources are distinctly different, which affects both the configuration of RO systems and their operating parameters.

5.2. Brackish and Nanofiltration RO Systems.

Majority of brackish and nanofiltration systems treat water that originates from wells drilled into an aquifer. Because well water contains low concentration of suspended solids, therefore requires limited pretreatment prior to being feed to the RO unit. Figure 15 shows schematic configuration of RO unit operating on a well water.

The pretreatment is limited to acidification and/or addition o scale inhibitor. From well pump, water flows through cartridge filter into suction of high pressure pump. Sometimes, prior to cartridge filters, a sand trap is installed on the feed line. High pressure pump develops feed pressure required for the membrane separation process and pumps feed water into an array of pressure vessels containing membrane elements. If the well water salinity is relatively low and does not contain any toxic constituents, part of the raw water can be used to blend with the RO permeate.

Blending improves economics of desalination process. In RO systems where blending is possible the following objectives can be achieved:

1. The RO unit can be smaller for a required output capacity.
2. Concentrate flow rate is lower for a give product capacity that would it be without blending.
3. Blending of raw water with RO permeate increases concentration of mineral constituents in the combined stream, therefore reduces corrosion tendency of the desalted water.

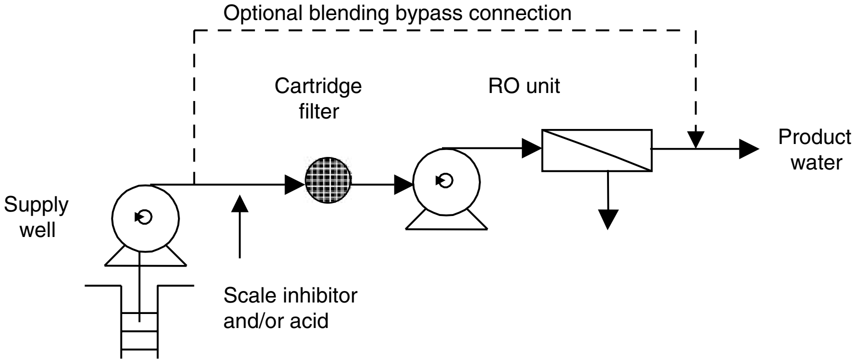


Fig. 15. Configuration of RO unit (brackish or seawater) operating on a well water feed.

RO systems that utilize blending of permeate with raw water, usually do not require any additional post treatment to stabilize product water. Otherwise, a treatment to increase of alkalinity, hardness and pH in product water is required. The treatment of permeate water usually involves removal of CO_2 by aeration and addition of compounds containing bicarbonate and calcium ions.

The important operating parameter of nanofiltration and brackish RO systems is the recovery rate. The usual objective is to design RO systems treating low salinity feed, for a maximum recovery rate possible. In majority of cases the major limitations on recovery rate is the concern of membrane scaling of the membrane elements located at the concentrate end of the RO system.

During the RO process water is removed form the feed stream and converted into permeate. The reduction of volume results in increase of concentration of all feed constituents. For example, in RO system operating at recovery rate of 80%, an equivalent fraction of feed water is removed as permeate. Correspondingly, the volume of feed water at the exit from the system is reduced by factor of 5 and concentrations of dissolved ions are increased roughly by the same factor. In system operating at 90% recovery, the increase of concentrations of individual ions in the concentrate stream would be about 10 fold.

If the scale forming salt is composed of two ions: [A] and [B], then 10 fold volume reduction will result in about 100 times increase of concentration product $[A] \times [B]$ of scale forming ions as compared to feed water. In addition to other factor, such as temperatures and ionic strength, this concentration product determines saturation level in the concentrate stream. Therefore, small increase of recovery rate, at the high recovery range, my result in a very sharp increase of saturation levels in the concentrate stream. The saturation levels of scale forming compounds, that are considered safe in RO applications, are defined by manufacturers and supplies of scale inhibitors. The commonly accepted concentration values are listed in Table 3.

The values listen in Table 2 are based on positive experience with commercial scale inhibitor. It is likely that in the future a more effective scale controlling chemicals will be developed and higher saturation levels can be maintained.

Some of the brackish and nanfiltration sources are locate close to the ocean. At these locations permit, that will allow discharge of concentrate to ocean, is usually granted. However, for inland sources far from the ocean, disposal of concentrate is very serious problem that in some cases may prevent completely an implementation of a desalination programs.

Table 3. **Commonly Acceptable Saturation Levels of Common Scale Forming Compounds in RO Applications**

Scaling constituent	Acceptable concentration level in the concentrate	Controlling agent
$CaCO_3$ in brackish application	Langelier saturation index (LSI) $< 1.8–2.2$	scale inhibitor or acid
$CaCO_3$ in seawater application	Stiff and Davis saturation index (S&DSI) < 0.5	scale inhibitor or acid
$CaSO_4$	$[Ca] \times [SO_4] < 2.4$ Ksp	scale inhibitor
SiO_2	$[SiO_2] < 180–240$ ppm	scale inhibitor

In the U.S., the location with largest number of brackish and nanofiltration systems producing potable water is the state of Florida. Currently the largest, low pressure, RO desalination system is located at Boca Raton. It is nanofiltration system of permeate capacity of 150,000 m^3/day. The feed water is of low salinity, about 400 ppm TDS but contains high hardness, about 400 ppm (as $CaCO_3$) and high color, about 35 color unit (CU). The plant produce permeate with hardness target of 50−80 ppm (as $CaCO_3$) and color less than 2 CU. The RO trains operate at feed pressure of about 5 bar and recovery rate of 85%. Large number of brackish water sources in Florida contain hydrogen sulfide (H_2S). Treatment of such sources, in addition to salinity reduction, also requires complete removal of H_2S, both from permeate and concentrate. This is being accomplished in aerators, operating in similar way as these for CO_2 removal, at water pH below 6.5. The one difference is that H_2S, being toxic, can not be vent off to the atmosphere. The off gases from H_2S aerators are usually adsorbed on iron catalyst that converts H_2S to elemental sulfur that is filtrated and disposed as a solid waste.

An example of brackish RO system, located in inland location is desalination plant located in Arlington California. The objectives of operation of the Arlington plant is to reduce salinity and nitrate concentration in local potable aquifer. The schematic process flow diagram of the Arlington Desalter is given in Figure 16. Feed water from the five local brackish wells is pumped to the plant site where it is split into two streams. Out of the total raw water flow of 26,500 m^3/d (7 MGD), provision exists for passing 7.500 m^3/d (2 MGD) through granular activated carbon (GAC) filters, to remove dissolved organic compounds, mainly dibromochloropropane (DBCP). The remaining flow, 19,000 m^3/d

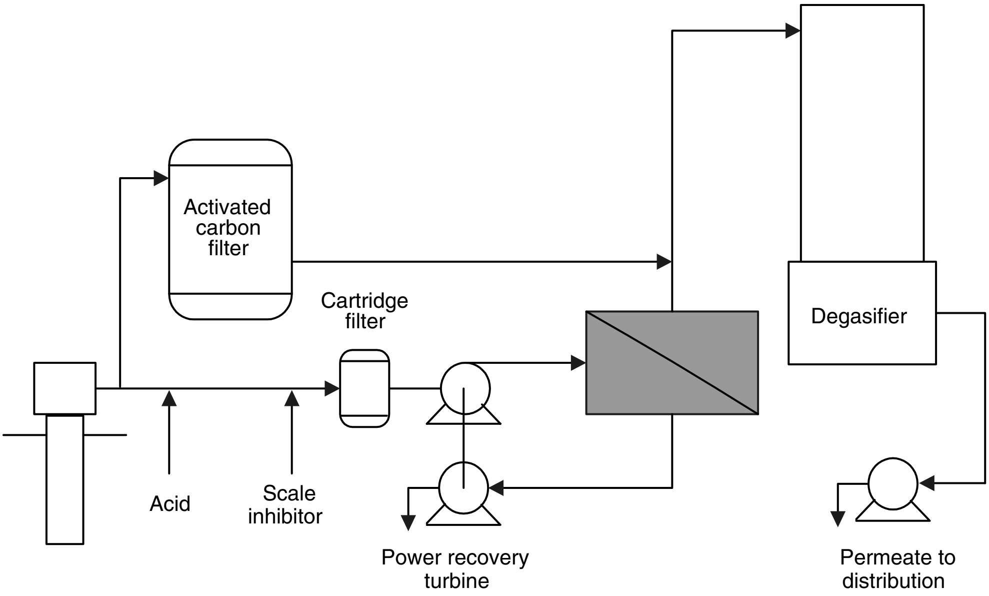


Fig. 16. Flow diagram of brackish RO plant for nitrate reduction and Arlington, Calif. (Courtesy Hydranautics.)

(5 MGD), is used as feed for the RO system. The RO feed water is treated by dosing of scale inhibitor and sulfuric acid to a pH of 6.9 and is filtered through 5-micron cartridge filters.

After the filtration feed water is pressurized to approximately 14.5 bar (210 psi) with vertical turbine pumps, the pressurized feed enters three parallel RO trains operating at 77% permeate recovery. Each train contains 44 pressure vessels, 8″ diameter, in a two pass 33:11 array. The pressure vessels each contain six spiral wound, composite, membrane elements. Permeate flow from the RO trains is combined with the blend stream at the ratio 2:1. The design blend ratio was based on the projected concentration of nitrate in the wells and in the permeate water, with a target concentration corresponding to California drinking water standard of not more than 45 ppm of nitrate in the total plant effluent. This blended effluent is of potable water quality and flows to the storm water channel and eventually recharges the ground water basin.

The concentrate stream from each RO train passes through an energy recovery turbine, which is a reverse running pump mechanically coupled with the high pressure pump. The combined concentrate from the plant is conveyed to the Orange County Sanitation District through the Santa Ana Regional Interceptor (SARI) line. After mixing with municipal sewage, and primary and secondary treatment of the Sanitation District effluent is split for further treatment by OCWD Water Factory 21 RO plant, or direct disposal to the ocean.

5.3. Wastewater Reclamation. The unique differentiation of membrane wastewater reclamation systems from brackish RO systems is quality of the raw water. The raw water treated in wastewater reclamation plants is almost universally secondary effluent form municipal wastewater treatment plants and contains high concentration of suspended particles, organic matters and bacteria. Early efforts to apply conventional pretreatment, that included lime clarification followed by multimedia filtration, were not successful. RO membranes fouled very rapidly and had to cleaned at high frequently. In the last ten years membrane pretreatment technology, mainly capillary microfiltration (MF) and ultrafiltration (UF) has been applied with very good results (38). A schematic diagram of wastewater reclamation system with membrane pretreatment is shown in Figure 17.

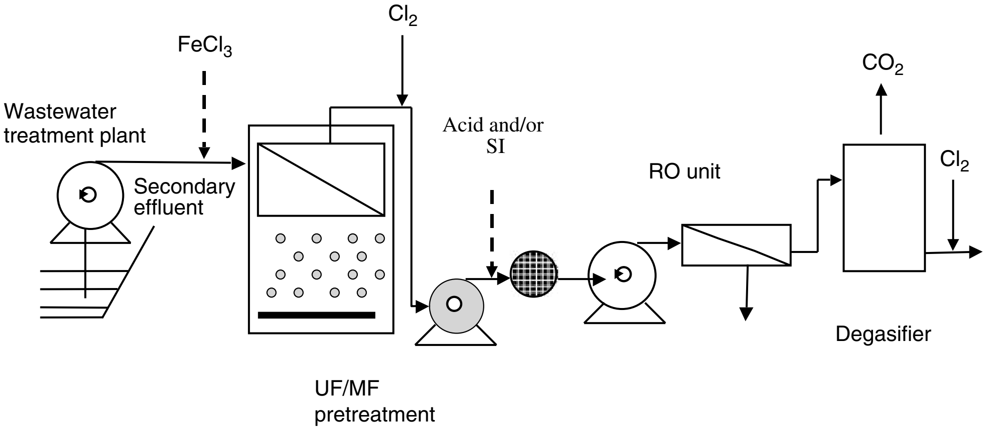


Fig. 17. RO wastewater reclamation system with membrane pretreatment.

The MF/UF technology applied as an RO pretreatment can be either pressure or vacuum driven. These membrane filtration systems operate at low feed pressure or vacuum (<1 bar) and at very high recovery rate (90–95%). The preferred mode of operation of MF/UF systems is direct filtration (15–30 min) followed by backwash (30–60 s). Foulants, that accumulate on MF/UF membrane surfaces during the filtration step, are removed in the backwash step through reverse flow of filtrate and/or air scouring. The effluent after MF or UF treatment still contains high concentration of dissolved organics but suspended particles and bacteria are almost completely removed. Fouling rate of RO membranes, downstream of MF/UF pretreatment, is much lower than experienced with conventional pretreatment, therefore cleaning frequency has been reduced significantly. Also, application of double membrane barrier increases confidence in effectiveness of reduction of pathogens, present in the raw water.

The desalination plants using combined membrane technology: MF/UF + RO are sometimes designated as an integrated membrane system IMS are being used extensively for treatment and salinity reduction of wastewater (39–41). Examples of representative IMS large plants in this category includes: Orange County, California, 265,000 m^3/day; West Basin, California, 87,000 m^3/day; Sulaibiya, Kuwait, 375,000 m^3/day; Bedok, Singapore, 32,000; and Kranji, Singapore, both at 40,000 m^3/day.

Product water from these plants is used for aquifer recharge, irrigation or industrial use. In Singapore, small fraction of RO reclaimed water, designated as "New Water" is pumped to one of the local potable water supply reservoirs and distributed for all potable use. Use of the reclaimed water, produced at the West Basin, California plant, demonstrates all typical usages encountered in this type of systems: 13% for unrestricted irrigation, 28% for ground water injections, 23% for low pressure boiler feed and 8% for high pressure boiler feed. Because of the requirements associated with the high pressure boiler operation, there are very strict requirements of low concentration of silica, hardness and TDS, imposed on permeate water quality at West Basin reclamation plant. The West Basin Reclamation Plant operational history demonstrates that such strict requirements are possible to meet and wastewater reclamation can be a reliable source of water for all application needs.

5.4. Seawater Desalting. A typical configuration of RO seawater system, processing water from an open intake is shown in Figure 18.

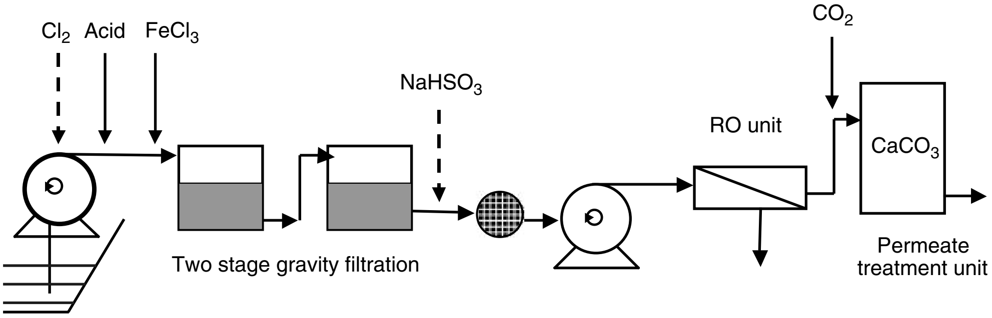


Fig. 18. Seawater RO unit treating surface water feed.

In the RO seawater desalination process seawater, pumped from the submersible intake, usually contain significant quantities of colloidal matter. Periodically, algae and other marine organism can be present as well. Prior to entry to the RO system suspended matter has to be reduced to absolute minimum, otherwise feed channels of the membrane elements will be plugged. The treatment of seawater usually includes rough screening followed by sedimentation, coagulation and media filtration, using either pressure or gravity filters. Filtered effluent is pumped through the cartridge filters that protect high pressure pumps against accidental entry of sand particles or other debris. The outlet of cartridge filters is usually connected directly to the suction of high pressure pumps. High pressure pump the treated seawater feed to the pressure require by the RO process, usually in the range of 5.5–7.0 MPs. The seawater enters RO modules and is separated into permeate and concentrate. The pressure of concentrate is about 0.2–0.3 MPs lower than the feed pressure. The RO systems usually operate at recovery rate of 40–50%, therefore, the concentrate stream contains significant amount of the original energy supplied by the high pressure pump to the feed stream. This energy is partially recovered using power recovery turbines. RO concentrate is returned back to the ocean. Appropriate discharge devices configurations are employed to mix the concentrate with ocean water and fast reduce concentrate salinity to the background level. Permeate water is usually stabilized by addition of calcium and bicarbonate ions to reduce its corrosion potential.

RO seawater desalination process requires membranes with very high salt rejection. In seawater applications average feed salinity could be in the range of 50,000–60,000 ppm TDS. To produce permeate water of 350 ppm TDS the actual membrane salt rejection has to be in the range of 99.3%–99.4%. Considering that in field conditions membrane rejection decreases with time and the required rejection has to be maintained in the whole range of feed water temperature, the initial nominal slat rejection of seawater membrane elements should be in the range of 99.7%–99.8%, which is achievable with the current commercial seawater membranes. This allows design production of potable water in a single pass systems for the lower range of seawater salinity (35,000–40,000 ppn TDS). For high seawater feed salinity and feed water temperature (~45,000 ppm TDS & ~35 C) seawater systems are usually configured as a partial two pass process (Fig. 19).

5.5. Energy Consumption. Because the concentrated reject brine is still at high pressure, it is possible to recover energy by passing this brine through hydraulic turbines and thus reduce the overall energy consumption. The common energy recovery device, used presently in medium and large capacity seawater units, is Pelton wheel. The Pelton wheel turbine is connected on common shaft with motor and high pressure pump. The concentrate stream flowing through the recovery device, turns the Pelton wheel, which transfers its energy to the electric motor and therefore reduces amount of energy required to turn the high pressure pump. Figure 20 shows a diagram of RO unit with a energy recovery device.

In this particular example the energy consumption of the RO process is 2.6 kWh/m^3. Without energy recovery device the same process would consume

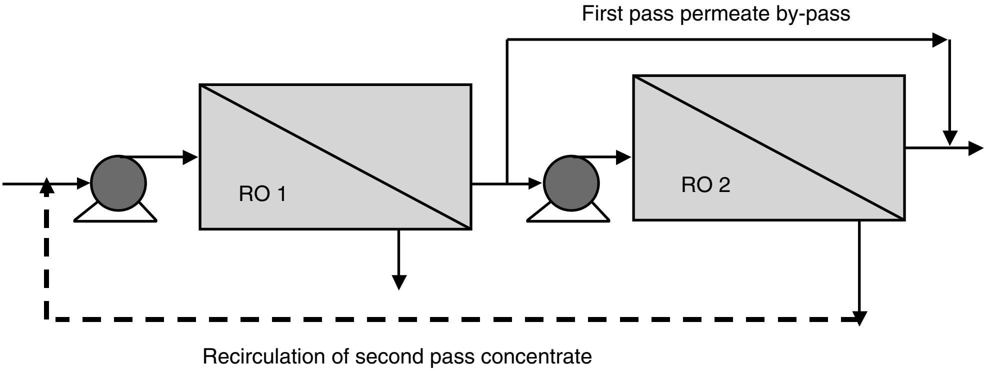


Fig. 19. Configuration of a partial two pass RO unit.

3.96 kwh/m^3. Therefore, use of energy recovery device provide energy reduction of about 35%. The amount of energy recovered from the concentrate stream depends mainly on efficiency of energy recovery devices. Presently, new energy recovery devices called isobaric devices are being introduced that have higher conversion efficiency then the Pelton wheel. Use of isobaric devices can provide additional reduction of about 10% of energy required by high pressure pumps in the RO desalination process. In addition to energy required by high pressure pumping system, the desalination plant uses electric energy for delivery of raw water, to operate pretreatment system and to pump the permeate. The total energy consumption is in the range of RO seawater plant is in the range of 3–4 kWh/m^3 (11–15 MJ/m^3). In comparison, the MSF desalination process requires about 120–280 MJ of heat and about 15 MJ of mechanical/electric power (for pumping and auxiliaries) per m^3. The energy requirement of the RO process is thus smaller than that of the MSF process even if the RO energy requirement is multiplied by the thermal-to-mechanical (or electrical) power conversion factor of 3–4. The specific exergy (or availability, the thermodynamic potential of useful work) consumption of the MSF process using 120°C steam is about 2–3-fold higher than that of the RO process, but the difference becomes smaller if the steam temperature is lowered to 80°C.

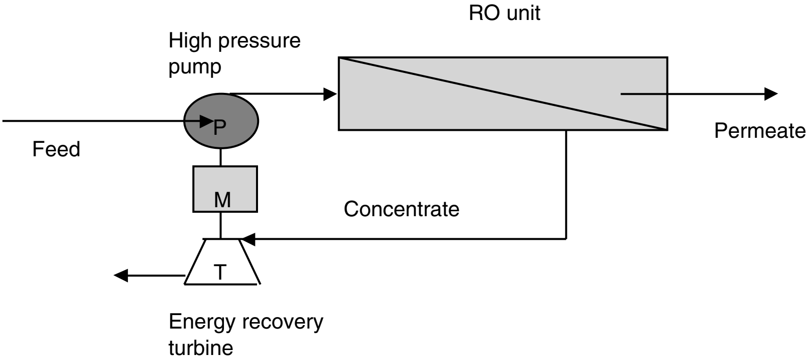


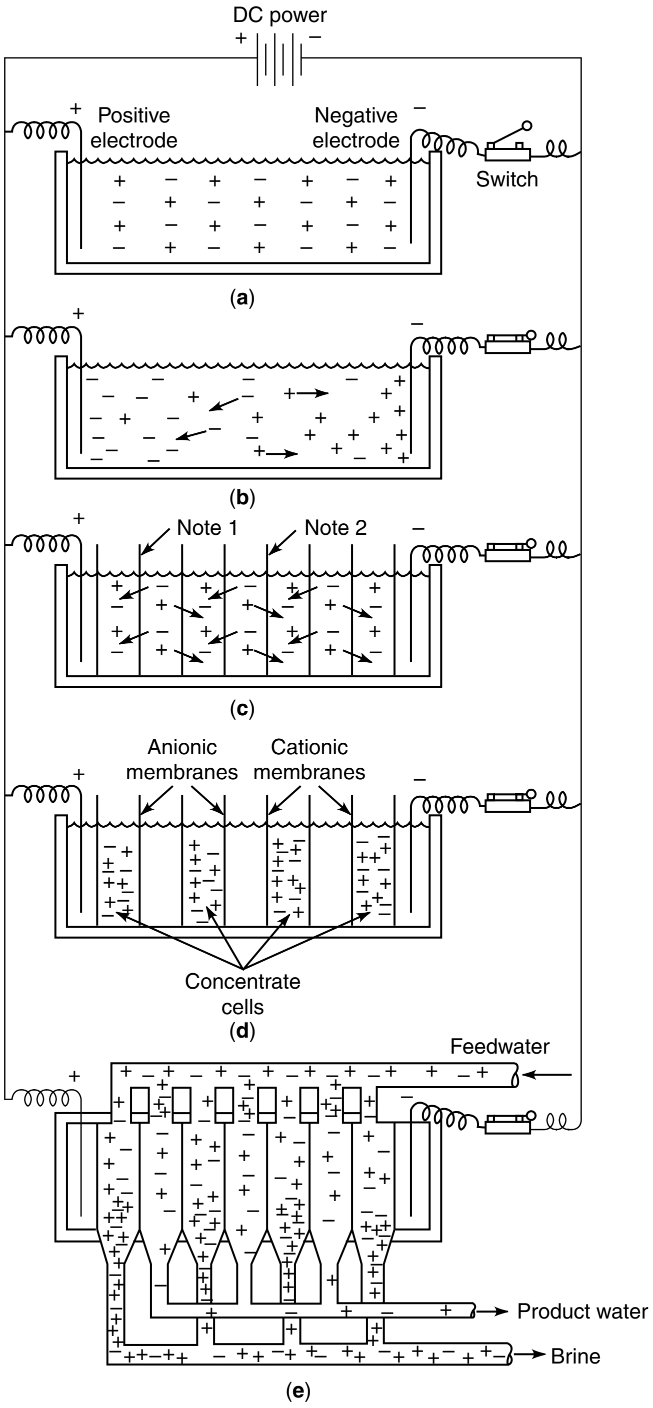
Fig. 20. Configuration of a RO unit with energy recovery turbine.

Energy consumption in nanofiltration and brackish water desalination plants is much lower than in RO seawater systems due to lower feed pressure and higher recovery rate. It is about 0.4–0.5 kwh/m^3 in nanofiltration plants and 0.5–0.8 kwh/m^3 in brackish systems. Energy recovery devices are being used in brackish plant, treating high salinity feed and operating at high feed pressures. In nanofiltration systems due to low pressure an high recovery rates, very little energy is contained in the concentrate stream, and use of energy recovery devices is not cost effective.

5.6. Electrodialysis. In electrodialysis (ED), the saline solution is placed between two membranes, one permeable to cations only and the other to anions only. A direct electrical current is passed across this system by means of two electrodes, causing the cations in the saline solution to move toward the cathode, and the anions to the anode. As shown in Figure 21, the anions can only leave one compartment in their travel to the anode, because a membrane separating them from the anode is permeable to them. Cations are both excluded from one compartment and concentrated in the compartment toward the cathode. This reduces the salt concentration in some compartments, and increases it in others. Tens to hundreds of such compartments are stacked together in practical ED plants, leading to the creation of alternating compartments of fresh and salt-concentrated water. ED is a continuous-flow process, where saline feed is continuously fed into all compartments and the product water and concentrated brine flow out of alternate compartments. The flow along the membranes also improves the mass transport there, and the separators between the membranes are constructed to provide good flow distribution and mixing on the membrane surfaces. Membrane sizes are often about 0.5 × 1 m, spaced about 1 mm apart. Many types of polymers are used to manufacture these ion-exchange-selective membranes, which are often reinforced by strong fabrics made of other polymers or glass fibers.

Since membrane fouling could quickly render the system inefficient, very careful and thorough feedwater pretreatment similar to that described in the section on RO, is required. Some pretreatment needs, and operational problems of scaling are diminished in the electrodialysis reversal (EDR) process, in which the electric current flow direction is periodically (eg, 3–4 times/h) reversed, with simultaneous switching of the water-flow connections. This also reverses the salt concentration buildup at the membrane and electrode surfaces, and prevents concentrations that cause the precipitation of salts and scale deposition. A schematic and photograph of a typical ED plant are shown in Figure 22.

The voltage used for electrodialysis is about 1 V per membrane pair, and the current flux is of the order of 100 A/m^2 of membrane surface. The total power requirement increases with the feedwater salt concentration, amounting to about 10 MW per m^3 product water per 1000 ppm reduction in salinity. About half this power is required for separation and half for pumping. Many plant flow arrangements exist, and their description can be found, along with other details about the process, in References 42 and 43. Many ED plants, as large as 15,000 m^3/d, are in operation, reducing brackish water concentration typically by a factor of 3–4.


DC power
+
−
Positive electrode
Negative electrode
Switch
(a)
(b)
Note 1
Note 2
(c)
Anionic membranes
Cationic membranes
Concentrate cells
(d)
Feedwater
Product water
Brine
(e)

6. Solar Desalination

The benefits of using the nonpolluting and inexhaustible energy of the sun for water desalination are obvious. Furthermore, many water-poor regions also have a relatively high solar flux for a large fraction of the time. The major impediment to the use of solar energy is economical: the diffuse nature of solar energy, even at its highest, dictates the need for constructing a large solar energy collection area. For example, typical annually averaged solar energy inputs in sunny regions range from 17–23 MJ/m^2/d. The latent heat of evaporation of water is about 2.407 MJ/kg at 40°C, and assuming a single-effect solar still efficiency of 50% (which is the upper practical limit for conventional designs), the still would produce at most about 3.5–4.8 (kg fresh water)/m^2/d, or a 208–286-m^2 solar still would be required to produce 1 m^3 of fresh water per day. More realistic still efficiencies reduce the production rate to about 2 (kg fresh water)/m^2/d and increase the still area requirement for producing 1 m^3 of fresh water per day to about 500 m^2. Consequently, whereas the energy input for solar desalination is free, the capital investment may be high and thus the cost of solar-desalted water may be higher than that of water desalted by processes which indeed require non-free energy input (such as fuel) but have lower capital costs. Solar desalination technology is mature and reliable, and thus becomes competitive with other desalination processes in regions or circumstances where the solar energy input and cost of fuel or of environmental penalties of its use are high, and the construction costs of the solar plant are low.

A typical solar still, shown in Figure 23, consists of a saline water container in which the water is exposed to the sun and heated by it. The temperature rise to above ambient causes net evaporation of the saline water, thus separating pure water vapor from the solution. The transparent cover of the still serves

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Fig. 21. Ion movements in the electrodialysis process. Courtesy U.S. Agency for International Development. (**a**) Many of the substances which make up the total dissolved solids in brackish water are strong electrolytes. When dissolved in water, they ionize; ie, the compounds dissociate into ions which carry an electric charge. Typical of the ions in brackish water are Cl^-, Na^+, HCO_3^-, Mg^{2+}, SO_4^{2-} and Ca^{2+}. These ions tend to attract the dipolar water molecules and to be diffused, in time, fairly evenly throughout a solution. (**b**) If two electrodes are placed in a solution of ions, and energized by a battery or other direct-current source, the current is carried through the solution by the charged particles and the ions tend to migrate to the electrode of the opposite charge. (**c**) If, alternatively, fixed-charge membranes, which are selectively permeable to ions of the opposite charge, are placed in the path of the migrating ions, the ions are trapped between the alternate cells formed. A positively fixed-charge (anionic) membrane allows negative ions to pass but repels positive ions. A negatively fixed-charge (cationic) membrane allows positive ions to pass, but repels negative ions. (**d**) If this continues, almost all the ions become trapped in the alternate cells, which lack ions, have a lower level of dissolved constituents, and have a high resistance to current flow. (**e**) The phenomenon illustrated above is used in electrodialysis to remove ions from incoming saline water on a continuous basis. Feedwater enters both the concentrate and product cells. Up to about half of the ions in the product cells migrate and are trapped in the concentrate cells. Two streams emerge from the device: one of concentrated brine and the other with a much lower concentration of dissolved solids (product water).

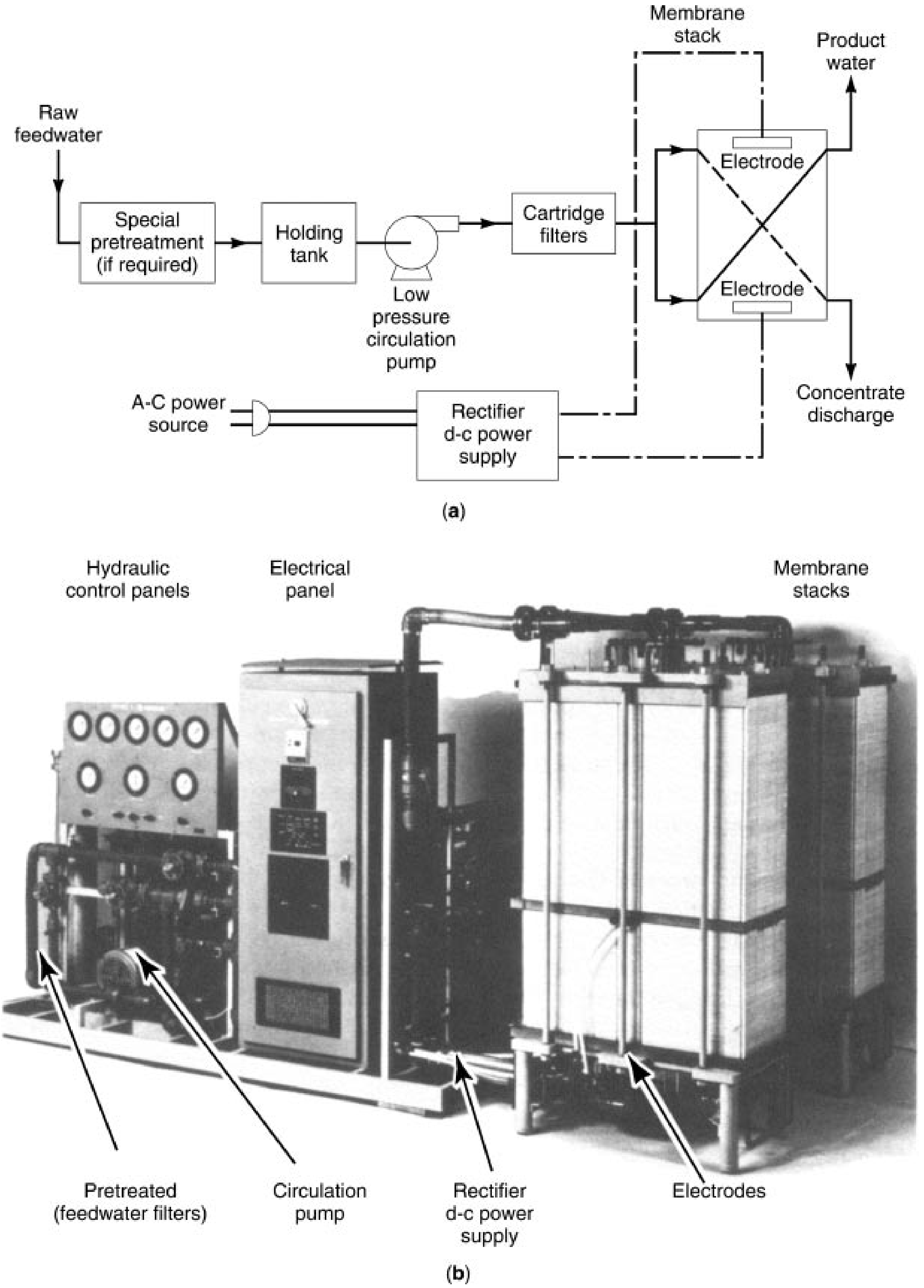

Fig. 22. Schematic (**a**) and basic components (**b**) of an electrodialysis unit. (Courtesy of Ionics.)

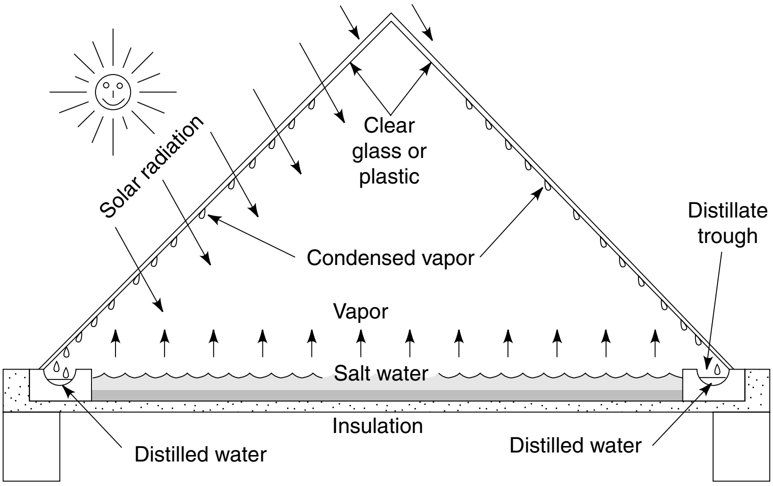

Fig. 23. A typical basin-type solar still.

several important functions: it prevents vapors from escaping the still; exposed to air at temperatures lower than the heated saline water in the basin, it serves as a condenser for the vapor; given proper inclination and geometric configuration, it channels the condensed fresh water to product collection troughs; it reduces heat loss from the warm saline water to the outside; and it prevents dirt from entering the still. However, it also must be as transparent as possible to solar radiation so that maximal heat gain by the still can be accomplished.

Because water of depths below about 2 m does not absorb much solar radiation directly, the radiation is absorbed and converted to heat primarily in the basin floor, which thus should have high radiative absorbance in the solar radiation spectrum. It is also noteworthy that if the still is designed to have low heat losses to the ambient, and if the ambient temperature drops, distillation will continue for some time even in the absence of solar energy input, because the saline water may remain warmer than the condensing glass surface and thus continue evaporating.

Solar stills of the type depicted in Figure 20, in many sizes and constructional variants, have been built and used successfully in many countries in the world. They are simple, easy to construct, reliable, and require very little maintenance.

Because the heat of condensation in single-effect stills of the type shown in Figure 17 is lost to the ambient, it is obvious that more energy-efficient operation would be achieved if a multi-effect design could be implemented, in which the heat of condensation is used to evaporate additional saline water. There are several problems with the design implementation of this concept. The principal one is the rather low available temperature driving force for separation in even a single-effect still: the condensation temperature is the ambient, say at 25°C, whereas the diurnal average temperature of the saline water in the still is around 40°C, only 15°C higher. Another difficulty is configurational, owing to the need for adding effects without at the same time obstructing solar radiation incidence on the saline water. A number of such stills were built and tested successfully, but are not yet commercially competitive.

Solar stills integrate the desalination and solar energy collection processes. Another approach to solar desalination is to use separately a conventional desalination process and a solar energy supply system suitable for it. Any compatible desalination and solar energy collection processes could be used. Distillation, such as MSF or ME, can be used with heat input from solar collectors or concentrators (44) or from solar ponds (45–47). Net average solar energy conversion efficiencies of solar collectors (48,49) are about 25%, and of solar ponds (45–47,50,51), 18%, similar to that of solar stills, but the MSF or ME plants can operate at performance ratios of 10 or more, thus basically increasing the freshwater production rate by at least tenfold, or reducing the required solar collection area by at least tenfold for the same production rate.

Still in the R&D stage, open-cycle ocean−thermal energy conversion (OTEC) plants using surface condensers produce desalted water (52,53). The concept is illustrated in Figure 24. The warmer surface ocean water is pumped

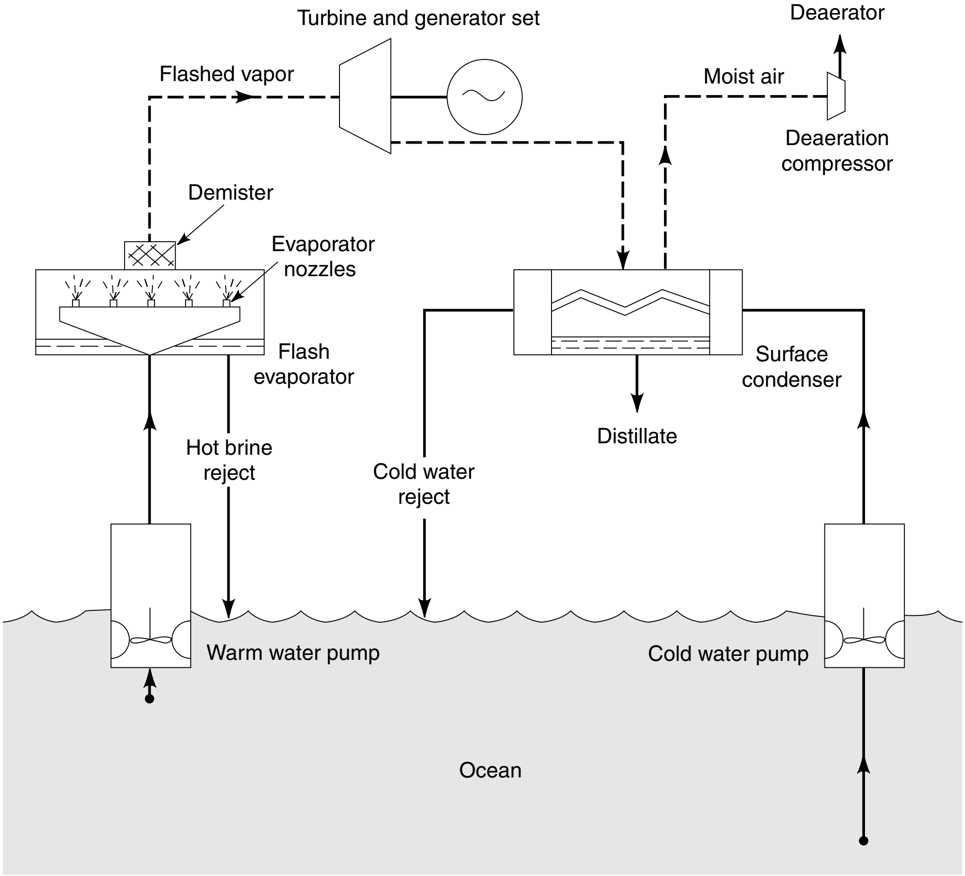

Fig. 24. Flow schematic of an open-cycle ocean−thermal energy conversion (OTEC) and desalination plant (49), where (—) represents liquids; (– – –), vapors or gases.

into a flash evaporation vessel from which the air has been evacuated. The vapor generated by flashing passes through a turbine and thus produces power, and is then condensed in a deaerated surface condenser cooled by colder (typically by about 20°C) seawater pumped to it from the ocean depths. The condensed vapor is freshwater, which can then be pumped out and used. In addition to requiring no fuel or other form of nonrenewable energy, OTEC and solar pond desalination have an advantage in that thermal energy storage is naturally included in the large thermal mass of the ocean or pond, while such storage may need to be constructed where the heat to the desalination process is supplied by solar collectors. It is noteworthy that fresh water can be produced by the OTEC-type cycle even without power production.

Solar or wind energy can also be used for desalination processes which are driven by mechanical or electrical power, such as VC, RO, and ED. The solar heat can be used to generate the required power by a variety of means. Figure 25 shows the flow diagram of a reverse osmosis plant in which the pumps are driven by photovoltaic solar cells, built by the French Atomic Energy Commission at Cadarache, France (54). A similar system, powered by an 8-kW array of photovoltaic modules has also been installed in Jedda, Saudi Arabia (55). A wind-powered vapor compression desalination demonstration plant using a 60-kW (max) electric power output wind turbine and producing up to 48 m^3/d of freshwater was installed on the island of Borkum, and a prototype plant using a 300 kW (max) electric power wind turbine and producing up to 360 m^3/d of

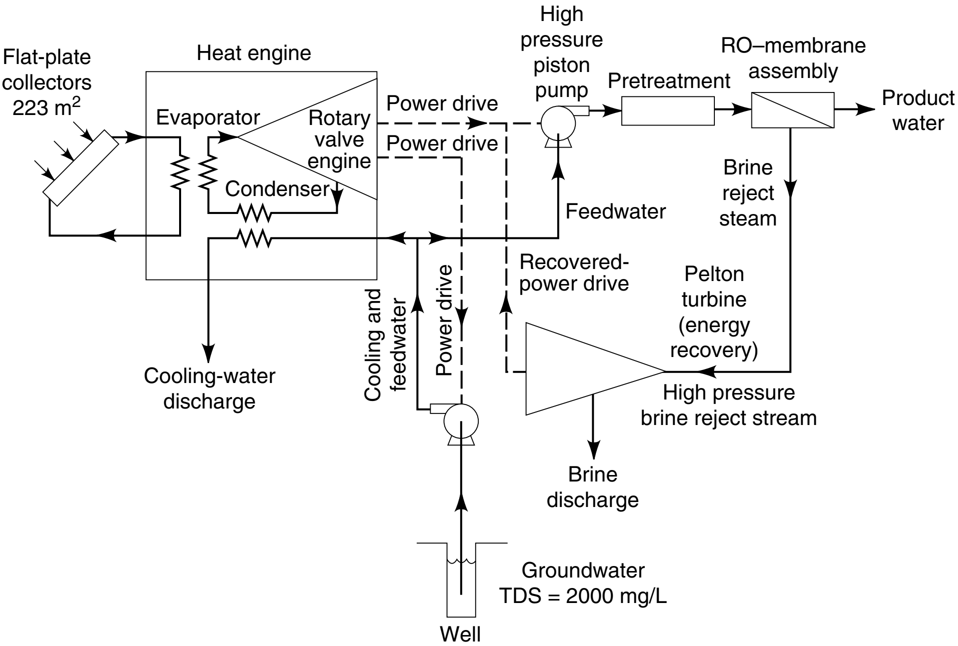

Fig. 25. Flow diagram of the experimental solar RO unit at Cadarache, France. Feedwater flow = 1.38 L/s, product water flow = 0.69 L/s, energy consumption without recovery = 0.89 kW-h/m^3 (48). (Courtesy of Commissariat d'Energie Atomique, France.)

freshwater was installed on the island of Rügen, both in the North Sea of Germany (56). The electricity produced by the wind-turbine–generator system is used to drive the VC plant compressor and to provide resistance heat to the VC system. Variable water output by the plant resulting from variations in wind velocity is accommodated easily by product water storage. The reported initial results have shown efficient and reliable operation.

7. Hybrid Desalination Systems

The hybrid desalination concept is the combination of number of desalination technologies and achieve improved performance and economics as compared to individual processes. Some of the hybrid configurations were discussed already. For example utilization of nanofiltrtaion technology as a pretreatment of makeup water for MSF unit will improve efficiency of operation of the distillation system. This is a serial configuration; seawater is processed in one unit, which produces improved effluent as a feed to the next unit, which is based on a different desalination technology. More common hybrid systems are of parallel configuration. Units based on different technologies treat parallel streams of feed water and product water from both systems is blended together. An example of such systems are desalination plants consisting of distillation plant (usually MSF) and seawater RO unit. Schematic configuration of a hybrid desalination system is shown in Figure 26.

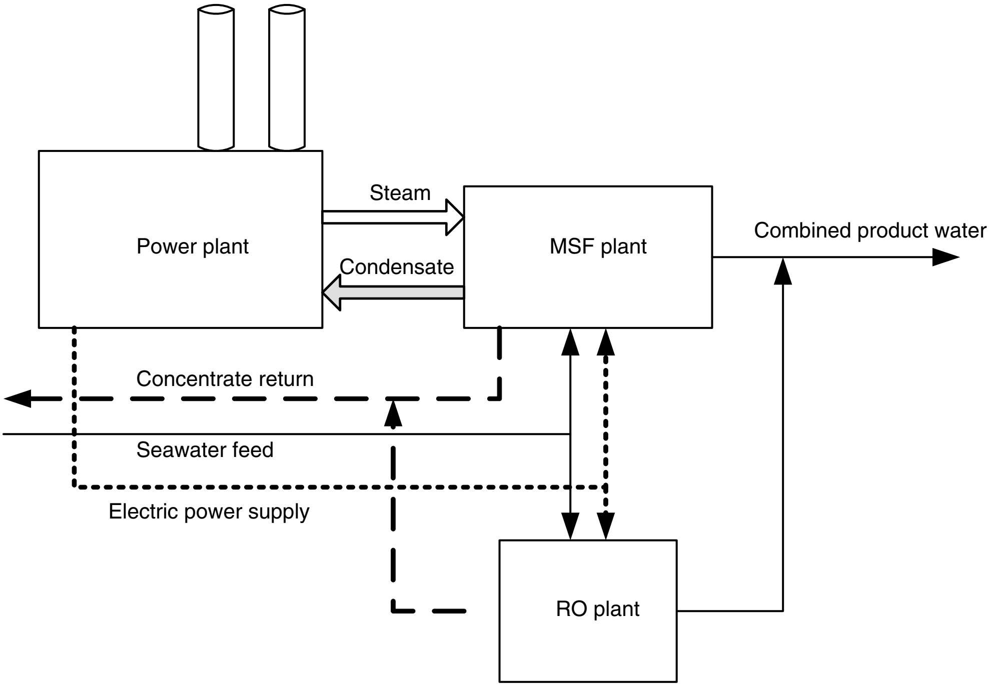

Fig. 26. Configuration of a hybrid desalination plant.

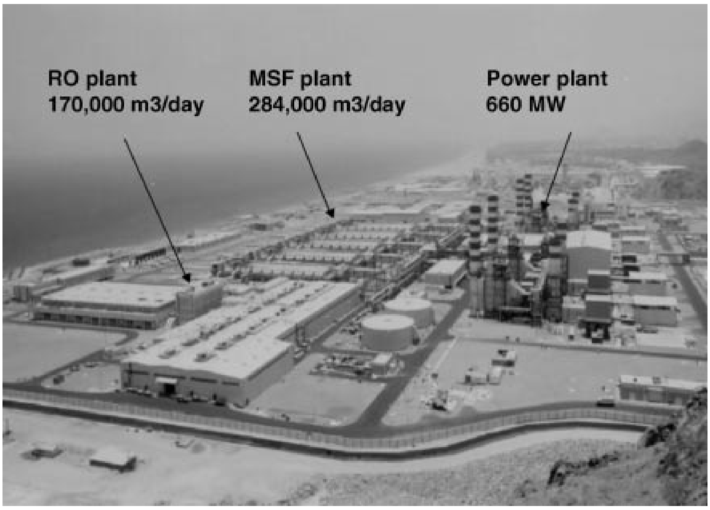

Fig. 27. Fujairah hybrid desalination complex, net output: power 500 MW, water 454,000 m^3/day. (Courtesy Doosan Engineering.)

Number of such systems have been built already in the middle east area. Probably the oldest large plant of hybrid configuration is the Jedda plant, Saudi Arabia, combining MSF unit of capacity 303,000 m^3/day and seawater RO unit of capacity 116,000 m^3/day (57). The latest addition to this category is desalination complex at Fujairah, UAE (58), which started operation in 2003. The Fujairah hybrid complex consists of power plant rated at 660 MW. The net power output is 500 MW. The desalted water capacity is 454,000 m^3/day. It consists of MSF output of 284,000 m^3/day and seawater RO output of 170,000 m^3/day (see Fig. 27).

The hybrid configuration plant which combines distillation plant with seawater RO provides benefits of increased production flexibility and reduced water cost at locations where seawater distillation desalting is prevalent technology (59). As explained in previous chapters dealing with distillation methods, The only practical configuration of large MSF desalination plants is utilization spent steam from electric power plant as a source of energy required for the distillation process. This arrangement requires parallel operation of MSF and the power plant. The fluctuations of water and electricity demands are not always similar. In Middle East countries, where majority of thermal desalination plants are located, demand for electricity is highest in summer and decreasing down to about 50% in winter. However, water demand is relatively constant through the year. Satisfying all potable water demand from MSF plant would require operation of power plant also at periods of low electricity demand. On the other hand, RO plant consumes only electricity as a energy source for the process. Therefore, operation of RO plant increases base load of the power plant, indirectly enables MSF unit to operate. The reverse benefit from the MSF in the hybrid configuration is the MSF produces product water of very low salinity, usually below 25 ppm. This allows RO to produce permeate of much higher salinity, while

Table 4. **Residential Water Costs, U.S. $/1000 ft^3 (28.3 m^3)**[a]

	City					
Year	El Paso, Texas	Albuquerque, New Mexico	Los Angeles, California	Newark, New Jersey	Boston, Massachusetts	Philadelphia, Pennsylvania
1978	3.99	4.48	4.55	7.79	8.90	8.13
1994	7.89	10.54	15.80	15.13	19.92	14.07

[a] Ref. 60.

maintaining blended product at the potable quality required. This situation reflects directly on process economics, reducing the cost of combined operation.

8. Economic Aspects

As shown in Table 4, residential water rates in the U.S. have gone up significantly between 1978 and 1994 (60).

According to the *1994 National Water and Wastewater Rate Survey* (60), the lowest rate per 28.3 m^3 of water in the United States was $5.40 in Saginaw, Michigan, and the highest, $44.88 in Scranton, Pennsylvania. Since the rates for wastewater were higher, by an average of $2.59, increased rapidly from a surfeit of $0.05 in 1986, and since an increasing number of utilities also charge customers for storm water runoff treatment, the actual costs to the customer are higher than those shown in Table 4.

The range of consumer prices for water in the developed countries in 1990 ranged from about $9.90 in Norway to $45.28 in Greece, per 28.3 m^3, or $0.35–1.60/m^3. In comparison, 2000–2005 water costs from large seawater desalination plants were in the $0.6–1.2.0/m^3 range (61).

Paradoxically, even rapid lowering of demand may cause rapid cost escalation; this was the case in New Jersey during a 1981 drought when conservation caused substantial shrinking of demand and private water companies had to double their water prices. The continued increase of demand and reduction of supply portend real and relentless water-cost increases in every part of the-United States in the future. One possible way to assure at least adequate supplies, and possibly to moderate these cost increases, is through water reuse.

9. Summary and Future Prospects

Rising population, standards of living, and water pollution are gradually diminishing the amounts of naturally available freshwater of good quality, whereas the demand is relentlessly increasing. Manufactured water in the form of water desalination is today making a considerable contribution to the world's potable water supply as well as water for industry, ranging from boiler feedwater for the power industry to ultrapure water for the electronics industry. The process industries are also looking at desalination technology either for better process water, or as means of conservation through recycling of wastewaters.

In some industries, it is considered a technique for recovering some process chemicals from water and lowering the cost of purification of polluting discharges. Some desalinated water is also used in agriculture, but only to a limited extent because of cost. The technology is improving both in cost–performance and reliability, as evidenced not only by the rapid growth in worldwide desalination capacity, but also by the encouraging observation that costs of desalted water has declined despite inflation.

Although desalination technologies are diverse, MSF has been for some time, and will remain well into the next century, the main process for thermal desalination of seawater. Inroads are being made by the multi-effect processes and, in particular, by the low temperature ME processes.

Important advances in the membrane field were responsible for the commercialization of seawater RO. This technique both saves energy and offers much-reduced plant construction time. Compared with distillation, RO treatment of seawater has had a meteoric success beginning in the early 1990s, and further growth is expected. Considering nanofiltration, brackish-water RO and to some extend EDR, the membrane processes have become an integral and growing part of desalination. In fact, their combined growth appears to exceed that of distillation.

The steady growth in water consumption around the world without any real growth of available water resources is focusing increasing attention on water reuse. Desalination technology is already an integral part of all water reuse schemes, whenever salinity reduction is required. Water reuse is already being implemented commercially, and within the next ten years there will be water-reuse schemes in virtually all areas of water use, ie, not only in water-scarce regions, but also in water-abundant regions such as the industrial northeast of the United States. Municipal authorities will become intimately involved in conversion of their wastewater in the future.

Much progress has been effected in desalination technologies. The more important recent advances in RO are increase of water permeability of RO membranes and introduction of new, more efficient, energy recovery equipment (42). Both developments contributed to reduction of energy consumption by reduction of required feed pressure and allowing increased recovery of energy of the concentrate stream. It is expected that the current desalination methods arrived close to a plateau of energy requirement and additional reduction of specific energy will be minimal. However, this could change if totally new technologies will be developed.

The projected significant future improvements, that could provide additional cost reduction, are expected introduction of larger diameter RO membrane elements and use of nanofiltration for pretreatment of feed water to MSF desalination systems. According to recent study (43) use of elements with twice diameter then the current 20 cm diameter elements, which translates to about four times current membrane area, could provide 6–10% reduction of product water cost. Process improvement of distillation technology–reduction of hardness in seawater makeup to the MSF plant (35) using nanofiltration membranes will result im more efficient operation of distillation plants, mainly due to increase of unit capacity. This technology could be applied both to existing plants and new installations.

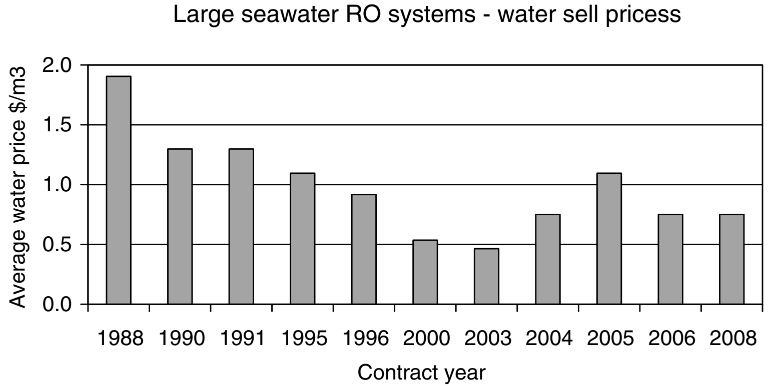

Fig. 28. Product water contract price for a large seawater RO projects (62).

Although energy costs have prompted the consideration of renewable energy resources as sources for desalination, these sources are unlikely to become significant, except in certain limited circumstances, until there are significant unforeseen breakthroughs in either efficiency of energy conversion or desalination technologies.

At present desalination of seawater is the only potential source of large quantity of new water of potable quality. In countries that are net importers of oil and experiencing water shortage, the desalination technology of choice is RO due to significantly lower energy requirement compared to the distillation methods. The major impediment to massive implementation of RO seawater desalting is water cost. During the period of 1990–2000 there was a steady reduction of cost of product water from large RO seawater systems, as shown in Figure 28 (62). The numbers indicate that in the last five years the product water sell prices have stabilized at the level of about \$0.7/m^3.

Despite the cost of desalination technology, it has made water available in places where it was not before. Not only has water become available in these places, but the quantities available have also opened prospects for industrial and commercial development. This has led to important improvements in the standard of living with prospects for even further improvements in countries fortunate enough to be able to meet the cost of the technology.

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DESALINATION ORGANIZATIONS

International Desalination Association. The IDA (formerly IDEA) is the major association specifically devoted to desalination. Beginning in 1976 it began organizing international congresses in desalination and water reuse. The proceedings of its conferences have been published (63). It also publishes a trade magazine (64). Information on IDA is available at: www.idadesal.org

American Membrane Technology Association. The AMTA, formerly the National Water Supply Improvement Association (NWSIA) and latter on renamed as American Desalination Association (ADA), is devoted to water supply improvement and considers desalination as a major technique. The association publishes a newsletter, *ADA News*. Information on AMTA is available at: www.membranes-amta.org

The European Desalination Society. The EDS formed recently, is the successor of the European Federation of Chemical Engineers Working Party on Desalination and Water Technology. The latter organization has held seven symposia in different parts of the world, and their published proceedings contain much valuable material (65). Information on EDS is available at: www.edsoc.com

EDS maintains site with links to large number of recent publications on desalination related subjects. This information can be acessed through: www.desline.com

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