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Thermodynamic and economic evaluation of co-production plants for electricity and potable water

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FOREWORD

Within the framework of the IAEA's activities related to seawater desalination using nuclear energy, a need was identified for developing criteria and methodologies in order to facilitate comparative economic evaluations of nuclear and fossil fuelled energy sources for desalination and generation of electricity. The aspect of costing of electricity and potable water from co-production plants is of particular interest.

In response to these needs, the IAEA carried out a study to establish methodologies for allocating costs to the two final products of co-production plants based on thermodynamic criteria and to enable economic ranking of co-production plant alternatives. This publication describes the methodologies and presents the results obtained from analysing a reference case, taken as an example.

This publication has been discussed and reviewed at a consultants meeting convened by the IAEA in September 1996 in Vienna. The methodologies have been incorporated in an EXCEL spreadsheet routine which is available upon request from the IAEA. The IAEA staff member responsible for this publication is L. Breidenbach of the Division of Nuclear Power and the Fuel Cycle.

It is hoped that the information contained in this report will be of value to decision-makers and the technical community in Member States interested in seawater desalination and considering the use of nuclear reactors as a potential energy source.

EDITORIAL NOTE

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1. INTRODUCTION
FRESH WATER SCARCITY AND DESALINATION MARKET

1.1. GENERAL

Clean fresh water is a basic need for human life, food production and economic development. Although the current usage of water in the industrialized countries may give an impression that fresh water is inexhaustible, about a quarter of the world's population lack the basic human supplies of sufficient food, clean fresh water supply and hygienic means of sanitation [1]. The vital importance of clean fresh water necessitates prudent water management, including efficient water use, recycling of wastewater (reclamation) and making available additional sustainable water resources. In coastal areas, seawater desalination offers a realistic alternative to cope with potable water shortage problems.

Seawater desalination has become a proven and reliable industrial process. By the end of 1995, about 13.6 million m^3/d seawater desalination capacity was installed or contracted worldwide [2]. According to world market projections, the demand for seawater desalination will continue to increase.

Most of the installed large-scale seawater desalination plants are distillation plants, which require mainly low-pressure saturated steam as heat source and some electricity for ancillary equipment (e.g. pumps). From the thermodynamic and economic points of view, it is useful to combine seawater distillation plants with electric power plants in integrated plants in which high-pressure steam is used to produce electricity in the turbogenerator, and the low-pressure exhaust steam from the turbine serves as heat source for the distillation. The construction of such integrated plant, particularly with large size units, which justify consideration of nuclear energy sources, involves high capital costs. The decisionmaker has to take all relevant factors into account to ensure that the best technical and economical plant configuration is selected, and that an adequate method of costing water and electricity is applied.

1.2. SCARCITY OF FRESH WATER

In some regions - especially in the Middle East and in Africa - fresh water is no longer available in sufficient quantity and quality. Reasons for fresh water scarcity are the world population increase from 2.3 billion to about 5.7 billion between 1940 and 1995, the per capita use of water from about 400 to more than 800 cubic meters per person per year in the same period, and the development or extension of large population and industry areas. It is estimated that the world population will grow to values between 8 to 9 billion by the year 2025. This is a clear indication that the scarcity of fresh water will become more critical and will cover additional areas in Asia, in Latin America and even in Europe [3].

To understand the limits of fresh water availability, it's useful to divide the 1.4 billion km^3 water on earth into its categories (Table I). Only 2.5% is fresh water, fit for drinking, agricultural purposes and most industrial uses. Moreover, about 69% of that is locked in polar ice caps and mountain glaciers or stored in underground aquifers too deep to tap with current technology [4].

In calculating how much fresh water is available for human use, what counts is not the sum of total fresh water available, but the rate at which fresh water resources are renewed or

TABLE I WORLDWIDE SALT WATER AND FRESH WATER RESERVES [4]

Salt water	1 365 million km^3 (97 5%)		
Fresh water	35 million km^3 (2 5%)	glaciers and permanent snow cover	24 million km^3 (68 7%)
		fresh groundwater	10 5 million km^3 (30 1%)
		fresh water lakes and river flows	0 1 million km^3 (0 3%)
		other including soil moisture, ground ice, etc	0 3 million km^3 (0 9%)

replenished by the global hydrologic cycle (**renewable fresh water**) Powered by the sun, this cycle each year deposits 113 thousand km^3 of water on the world's continents and islands as rain and snow Of that, about 72 thousand km^3 evaporates back into atmosphere Of the remaining 41 thousand km^3, more than half flows unused to the sea in floodwaters and about an eighth falls in areas too far from human habitation to be captured for use Some water experts estimate the practical limit of the world's available renewable fresh water at 9 to 14 thousand km^3 per year, however, a substantial proportion of this amount is needed to sustain natural ecosystems [3]

The critical limits in available fresh water, of course, are not at the global level but at regional and national levels In measuring a region's or country's (territory's) available renewable fresh water, the annual precipitation that falls on this territory and the water that flows into this territory from rivers and aquifers originating in neighbouring territories are to be added up From this, the losses through evaporation and the fresh water flows which run out of the territory are to be subtracted However, the available renewable fresh water calculated can only be used under ideal conditions Taking into account the technical and natural suitability of the territory to store water, the available renewable fresh water is much less Some developing countries can currently use not more than 20% of their potential renewable fresh water

The renewable fresh water of each country or region is to be compared with the fresh water consumption World wide, agriculture is the biggest user on water supplies, accounting for about 69% of all use About 23% of water withdrawals go to meet the demands of industry and energy, and just 8% to domestic or household use The division into the individual three categories varies greatly from country to country, depending on the economic development, climate and population size In Africa and in the Middle East, for instance, the demand on water consumed in agriculture is very high, while in highly industrialised European countries more than half of the water is used by industry and energy production Domestic and household water use - including drinking, washing, cleaning, food preparation, etc - accounts for only a small portion of total use in most countries, unless the industrial and agricultural sectors are not well developed

According to Table II, in 1990 the fresh water consumption of 9 countries in the world was higher than their available renewable fresh water [4] The extra demand on fresh water was compensated through withdrawals of non-renewable fresh water resources, reprocessing of waste water and desalination of seawater

TABLE II. COUNTRIES WHOSE WATER USE EXCEEDS 100% OF THEIR RENEWABLE WATER [4]

Country	Renewable fresh water resources [km^3/a]	Water consumption [km^3/a]	Water withdrawals in % of renewable fresh water resources
Libyan Arab Jamahiriya	4.622	17.286	374%
Qatar	0.050	0.087	174%
United Arab Emirates	0.489	0.685	140%
Yemen	5.199	7.019	135%
Jordan	1.311	1.442	110%
Israel	2.148	2.363	110%
Saudi Arabia	4.550	4.823	106%
Kuwait	0.161		> 100%
Bahrain	0.090		> 100%

However, Table II doesn't show the special water problems in large population and industry areas. The large increase in population growth and the growing industrialisation have led to the situation that fresh water has to be transported from water reservoirs that are located far away. Another problem which is not evident from Table II, is the contamination of ground and river water with chemicals and germs.

How can one counteract the increasing global and regional water scarcity? In the following, steps and possible solutions which should be incorporated in proper water management plans are presented:

- water resources and demand assessment,
- increasing the consumption of non-renewable groundwater resources,
- increasing of the efficiency in water use,
- extension of using reclaimed water,
- construction and extension of sewage treatment and disposal systems,
- seawater desalination applications.

Water resources assessment, which is the determination of the quantity, quality and availability of water resources, is a prerequisite for proper water management. Without detailed water resources assessment it is impossible properly to counteract the increasing water scarcity. In addition, adequate information about the future development on water demand is necessary. As a result, more effort should be undertaken to complete, update and validate existing data bases on water resources and expected water demand on regional and global basis.

Increasing the consumption of non-renewable groundwater resources is, of course, not sustainable, and only a temporary solution.

The greatest potential in counteracting the water scarcity is the improvement of efficiency in water use. Just in agriculture, the use of highly efficient irrigation technology could lead to water savings probably sufficient to cover the current demand of drinking and domestic water [3]. Israeli farmers, whose drip-irrigation techniques achieve up to 95% efficiency, have more than doubled their food production in the last 20 years without increasing their water use. Furthermore, increasing the cultivation of agricultural products in wet areas, and transporting them to water scarce areas, could bring further savings. By the improvement of freshwater transportation and distribution systems, estimated water leakage losses of 20 to 40% could be reduced. The use of water efficient technologies in industry and household forms a further potential in saving water.

The contamination of ground and river water could be counteracted by the construction and extension of sewage treatment and disposal systems.

In industrial countries, the potential of water savings by the extension of water reclamation of industrial and domestic waste water is estimated to be more than 50% [1]. In developing countries, where such installations are hardly available, the potential is much higher.

Supply of fresh water by seawater desalination plants is a capital intensive option. Since drinking water and water for sanitation service, however, form the basis of human life, seawater desalination is the best alternative to supply potable water in coastal areas when all less expensive options have already been exhausted. For industrial purposes, seawater desalination can be practical as long as the amount of water needed for industrial production is small, so that its contribution to production cost is small. For agricultural purposes, seawater desalination could only be viable in combination with highly efficient irrigation technologies.

1.3. WORLD MARKET OF SEAWATER DESALINATION

Figure 1 shows the historical development of seawater desalination plants in terms of the total capacity installed or contracted worldwide [2]. The most important users are in the Middle East with about 70% of the world capacity. Europe has 9.9%, the USA 7.4% (mainly California and Florida), Africa 6.3% and the remaining countries in Asia 5.8%. Currently, the most dominant seawater desalination processes are:

- Multi-stage flash (MSF) distillation with about 80% of the world market,
- Reverse osmosis (RO) with about 10.6%,
- Multiple effect distillation (MED) with about 9.0%.

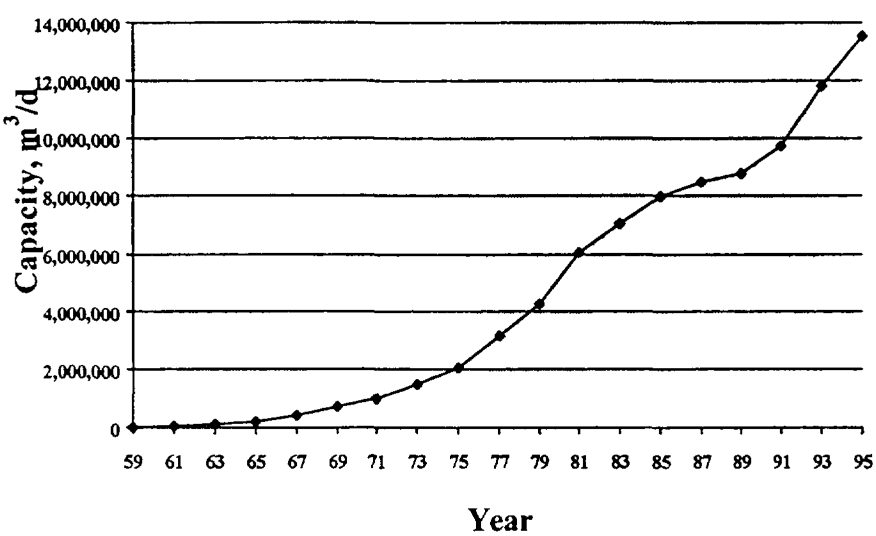


FIG. 1. Worldwide contracted cumulative seawater desalination capacity [2].

TABLE III. WORLD MARKET PROJECTIONS FOR SEAWATER DESALINATION [5]

	Capacity 1995 m³/d	Incremental installed capacity in the years m³/d				Estimated capacity 2015 m³/d
		1996-2000	2001-2005	2006-2010	2011-2015	
USA	183 400	322 971	302 783	483 831	773 136	2 066 122
Mexico	32 864	135 506	104 568	169 510	274 786	717 234
N. Antilles	73 481	28 198	27 991	35 696	45 523	210 889
Cyprus	8 681	44 850	32 531	52 301	84 085	222 448
Italy	126 370	84 073	149 919	256 721	439 609	1 056 692
Malta	122 117	66 716	102 265	157 648	243 025	691 771
Spain	249 315	306 769	197 321	267 338	362 201	1 382 945
Former USSR	136 924	64 356	60 416	78 551	102 128	442 376
Egypt	30 069	27 263	40 041	68 005	115 500	280 878
Libyan Arab Jamahiriya	393 842	195 511	152 999	192 718	242 748	1 177 818
Bahrain	92 717	131 556	71 017	93 505	123 114	511 909
India	13 415	69 817	34 803	49 355	69 992	237 382
Islamic Republic of Iran	319 397	268 716	424 297	730 408	1 257 365	3 000 184
Israel	45 468	145 124	37 432	44 784	53 579	326 387
Kuwait	1 195 895	245 999	214 820	246 825	283 598	2 187 138
Oman	145 343	141 757	96 577	129 065	172 481	685 222
Qatar	513 214	133 818	172 607	218 652	276 982	1 315 273
Saudi Arabia	3 733 747	1 060 526	1 680 828	2 270 110	3 065 990	11 811 202
United Arab Emirates	1 851 166	572 314	724 402	940 932	1 222186	5 311 000
Japan	17 898	49 489	35 671	54 553	83 430	241 040
Total:	9 285 323	4 356 041	4 851 007	6 818 536	9 706 577	35 189 078

It is expected that the proportion of RO and MED processes in the world market will increase as a result of the lower cost of water production compared to MSF processes.

An assessment of the current and projected seawater desalination capacities for municipal supply was carried out by the Agency in 1995 [5]. Based on historical records of installed seawater desalination capacities from 1973 to 1993 and on known orders for new capacities to be installed over the next several years, growth rates on projections of seawater desalination capacities were calculated and adjusted by correction factors as necessary. Table III shows the country-wise expected global incremental installed capacity for seawater desalination up to the year 2015. The following conclusions may be drawn based on the above assessment:

- **USA, Islamic Republic of Iran, Saudi Arabia and United Arab Emirates**
  The expected market for seawater desalination plants is in the order of 200 000 m³/d to 500 000 m³/d.

- **Mexico, Italy, Malta, Libyan Arab Jamahiriya, Bahrain, Oman and Kuwait**
  Expected incremental installed capacity of seawater desalination is in the order of 100 000 to 200 000 m³/d.

1.4. ECONOMIC EVALUATION

In this document, the economic evaluation of co-production plants for electricity and potable water is performed considering thermodynamic as well as economic aspects:

a) to enable an economic comparison of different co-production plants and
b) to establish an appropriate method of costing electricity and potable water.

The economic comparison of different co-production plants for electricity and potable water (integrated plants) is more difficult than for single purpose plants since they have simultaneously two final products, electricity and potable water. The plant with the least annual overall expenditures (capital charge, fuel cost, operation and maintenance costs related to both electricity generation and potable water production) is not necessarily the most economic solution, since it is unlikely that all alternatives will have exactly the same net outputs. Furthermore, both the potable water production cost and the electricity generation cost will vary from one alternative to the other, which makes a comparison difficult.

Appropriate methodology to compare the economics and rank different integrated plants has to be developed, in which the annual overall expenditures of the plant as well as the outputs of electricity and potable water are considered. In this study, a methodology is presented which enables the comparison of different plants with the *same potable water output* and with similar net base load power capacities. This methodology is called "calculation of the ***equivalent electricity generation cost***" (see Section 3.2).

After selecting the most economic integrated plant, a cost basis for the sale of potable water and electricity have to be established. There are several techniques for allocating the overall expenditures of the integrated plant to the two final products [6]. The selection of the most suitable method will depend on the objectives and the environment in which the plant is built.

The annual overall expenditure C_0 of the integrated plant can be expressed as a function of the annual electricity output E_a and the annual potable water output W. Cost allocation methods aim at expressing linearly the unit costs c_E and c_W for a particular integrated plant:

$$C_0 = c_E \cdot E_a + c_W \cdot W \qquad \text{in \$/a} \qquad (1)$$

The line representing Eq. (1) is shown in Figure 2. Its slope depends only on the water-to-electricity ratio. A modification of the economic assumptions used for calculating C_0 would result in moving this line up or down parallel to itself.

Two boundary points can be determined on this line as follows. If the whole economic benefit of combined production is assigned to the cost of potable water without penalizing electricity (i.e. power credit method using the electricity generation cost of a least-cost single purpose power plant), the value of electricity is known and point A can be placed on the curve. Point B is determined in the same manner, but with the entire benefit being assigned to the cost of electricity by using a water credit, the value of which would be equal to the cost of water produced in an alternative least-cost single purpose water scheme. The points on the curve which lie outside the segment AB correspond to subsidizing either the potable water or the electricity.

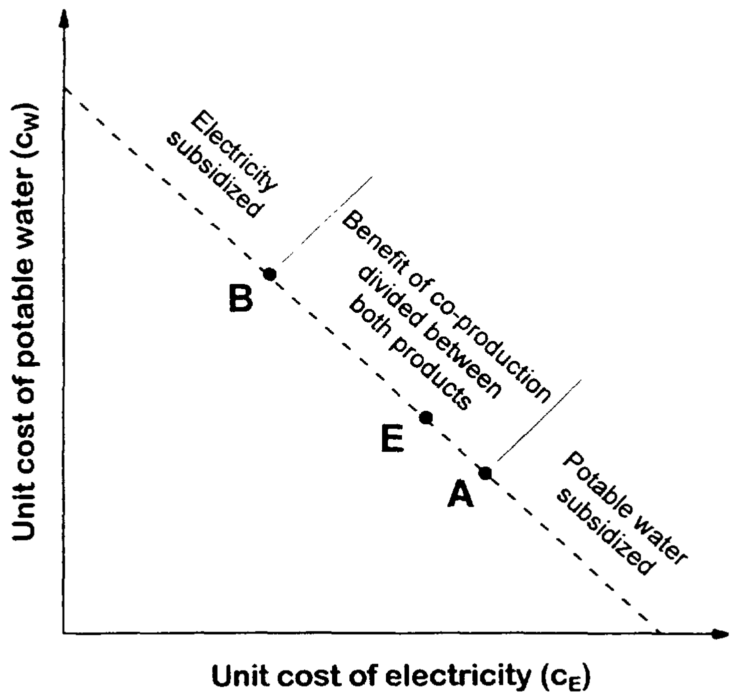


FIG. 2. Allocation of overall annual expenditures of an integrated plant for electricity and potable water production.

An appropriate cost allocation method should enable the distribution of the combined production benefit to both potable water and electricity, resulting in a point inside the segment *AB* in Figure 2. Furthermore, such a cost allocation method should enable an equitable and generally applicable breakdown of this benefit, preferably from a thermodynamic viewpoint. In such a method, the thermodynamic value (exergy) of the energy streams to produce potable water and electricity should be assessed to define the formula for cost allocation (point E in Figure 2).

Exergy or maximum achievable mechanical energy is a measure of the quality of energy; it is the upper limit of the share of energy which is transferable to mechanical work given a certain thermodynamic environment. It is assessed that the value of mechanical and electrical energy streams is higher than the value of heat, and that the heat of a high-temperature heat source is higher in value than the heat of a low-temperature heat source. In this study, an ***exergetic cost allocation method*** for co-production of electricity and potable water is presented, which is valid both for distillation and reverse osmosis processes.

Calculation of the equivalent electricity generation cost as well as the exergetic cost allocation method are applied to different integrated plants with the same net water output, located at a representative site on the Arabian Peninsula. The assumed site conditions, in particular, potable water and electricity demand, are typical for operating and planned integrated plants in this region. Various seawater desalination processes and designs (MSF, MED, RO), which were pre-selected on the basis of their favourable technical and economic characteristics and their commercial availability, are considered. A pressurized water reactor (PWR) power plant as well as a gas fired combined cycle power plant were selected as technically and economically viable energy sources for these site conditions.

2. DESALINATION PROCESSES

2.1. THEORY

Seawater desalination is the processing of seawater to obtain pure water through the separation of dissolved saline components. In general, seawater desalination processes can be classified into two categories:

(a) processes that separate pure water from seawater (saline solution):

- distillation processes,
- membrane processes;

(b) processes that separate salt from seawater:

- electrodialysis,
- organic adsorption,
- ion exchange.

Any desalination process requires energy, either heat and electrical energy (mainly for pumping) or electrical energy only (the use of mechanical energy instead of electrical energy is also possible). For standard seawater (25°C, 34 500 ppm total dissolved solids (TDS)), the theoretical minimum separation work required to produce 1 m^3 pure water is about 0.73 kW·h [7]. However, the energy consumption of currently available commercial processes is much higher due to thermal losses and irreversibilities that occur during the separation process such as transient phenomena and dissipative effects. The lowest energy consumption including that for seawater pumps and water pre-treatment is currently obtained with RO plants. It amounts to 4 to 7 kW(e)·h/m^3 of electrical energy dependant on fresh water quality, seawater salinity and plant configuration. This figure can be directly compared with the theoretical minimum separation work. The achieved ratio of real to theoretical minimum work is 5 to 7, dependant on the seawater salinity and temperature.

After more than 40 years of intensive research and development in seawater desalination technology, only distillation processes and the RO process have achieved commercial large-scale application. These processes are expected to be the leading processes in the near future.

2.2. DISTILLATION PROCESSES

In distillation processes, seawater is heated to evaporate pure water that is subsequently condensed. With the exception of mechanical vapour compression, distillation processes are driven by low-temperature steam as the heat source, which may be taken from a power plant.

From the beginning, distillation processes have been implemented in heat recovery stages placed in series as a result of the high specific heat required to evaporate water. Since the performance of distillation processes increases with increasing number of stages (increasing heat transfer area), it is advantageous to use a large number of stages. However, the overall temperature difference between the heat source and the cooling water sink as well as economic reasons limit the number of stages. Typical temperature differences for commercial distillation plants are 2-5°C per heat recovery stage.

Usually, the thermodynamic efficiency of distillation plants is expressed in kg of water produced per kg of steam used. This ratio is called the gain–output ratio (GOR), which is in the range of 6 to 10 for current commercial multi-stage flash (MSF) distillation plants and up to 20 for multiple effect distillation (MED) plants. However, it is to be noted that the GOR does not account for the steam temperature and therefore does not assign a thermodynamic value (exergy) to the steam. Thus, comparing different distillation plants by means of the GOR is only useful if the temperature difference between the heating steam and the seawater is the same. In this connection, a distillation plant with a GOR of 8 requiring steam of 70°C, for instance, is thermodynamically superior to a distillation plant with the same GOR but requiring steam of 120°C.

A thermodynamically correct method is to value the heating steam according to its exergy content, which is equal to the maximum work achievable by expansion in an ideal steam turbine (100% isentropic efficiency) to ambient temperature. For current commercial distillation plants supplied with low-temperature steam from adjacent power plants, the exergy consumption just for heating steam is in the range of 12 to 18 kW·h/m^3 for MSF, and 7 to 10 kW·h/m^3 for MED.

2.2.1. Multi-stage flash (MSF) distillation

Figure 3 shows the schematic flow diagram of an MSF system.

Seawater feed passes through tubes in each evaporation stage where it is progressively heated. Final seawater heating occurs in the brine heater by the heat source. Subsequently, the heated brine flows through nozzles into the first stage, which is maintained at a pressure slightly lower than the saturation pressure of the incoming stream. As a result, a small fraction of the brine flashes forming pure steam. The heat to flash the vapour comes from cooling of the remaining brine flow, which lowers the brine temperature. Subsequently, the produced vapour passes through a mesh demister in the upper chamber of the evaporation stage where it condenses on the outside of the condensing brine tubes and is collected in a distillate tray. The heat transferred by the condensation warms the incoming seawater feed as it passes through that stage. The remaining brine passes successively through all the stages at progressively lower pressures, where the process is repeated. The hot distillate flows as well from stage to stage and cools itself by flashing a portion into steam which is re-condensed on the outside of the tube bundles.

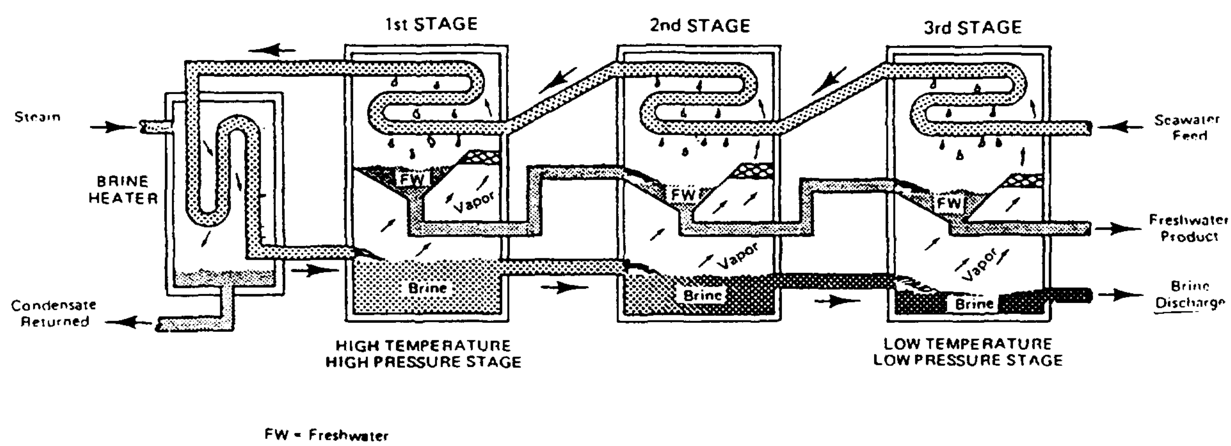


FIG. 3. Schematic flow diagram of an MSF system [8].

MSF plants need pre-treatment of the seawater to avoid scaling by adding acid or advanced scale inhibiting chemicals. If low cost materials are used for construction of the evaporators, a separate deaerator is to be installed. The vent gases from the deaeration together with any non-condensable gases released during the flashing process are removed by steam-jet ejectors and discharged to the atmosphere.

Figure 4 shows the temperature distribution in a given evaporation stage i. There are temperature losses in each stage, which reduce the temperature difference between incoming seawater feed and brine to a real temperature difference $\Delta\vartheta_i$, resulting in higher thermal energy consumption. The temperature losses consist of three components:

- boiling point elevation (BPE) of saline water in contrast to pure water, $\Delta\vartheta_{BPE}$ (0.5-1.2°C according to the operating point of the MSF plant and seawater salinity),

- non-equilibrium temperature loss (NEL) $\Delta\vartheta_{NEL}$, which is caused by thermal and hydrodynamic effects like insufficient time for the superheated brine to evaporate completely, or a greater total static head (vapour plus liquid) on the brine near the bottom of the stage in contrast to the surface (0.2-1.0°C),

- temperature losses as a result of pressure losses of vapour while streaming across the demister and around the tube bundles, $\Delta\vartheta_{RV}$ (<0.2°C).

There are to two principal arrangements used in MSF systems: the brine recycle mode (MSF-BR), and the once-through mode (MSF-OT) (Figure 5). The majority of the MSF plants built use the brine recycle mode. The brine recycle mode was invented in the early years of desalination when seawater corrosion resistant materials and advanced additives were not available or too expensive. In brine recycle systems, the heat of condensation of vapour, produced in the last stages (Heat Rejection Section) is taken by cooling water, a major part of which is rejected back to the sea. Only a small part (about 2.5 times the amount of the product water) is deaerated and chemically treated against scaling, and is fed as make-up water to the subsequent stages. The required amount of feedwater to produce a certain amount of potable water is recirculated and kept below a maximum salinity by constantly

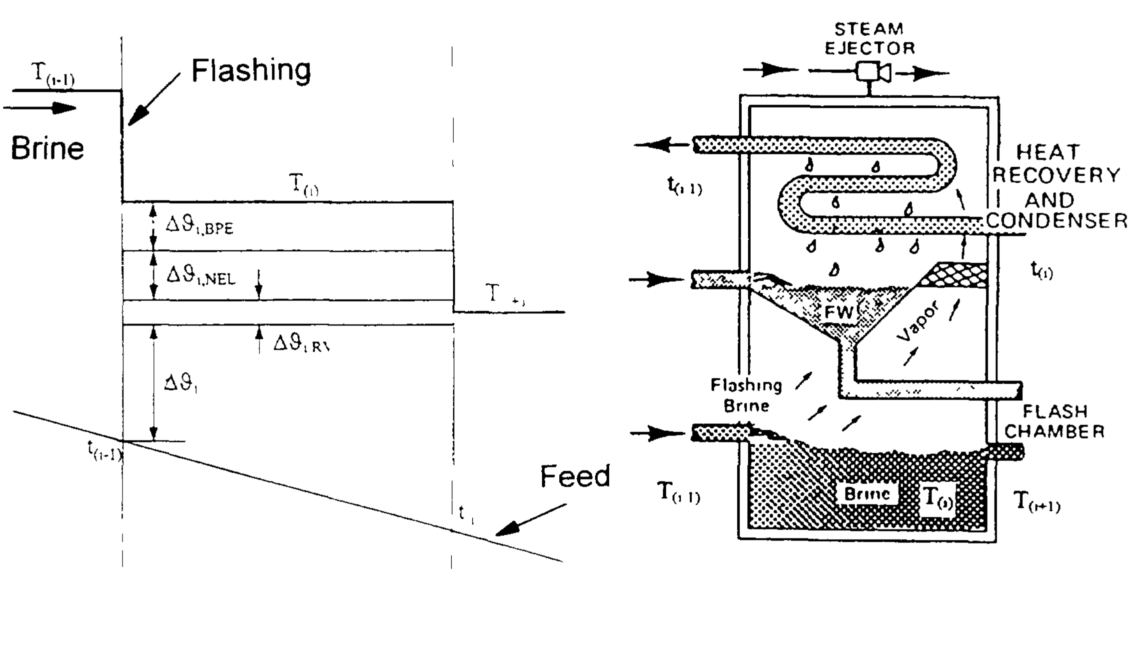


FIG. 4. Temperature distribution in an MSF stage i [9]

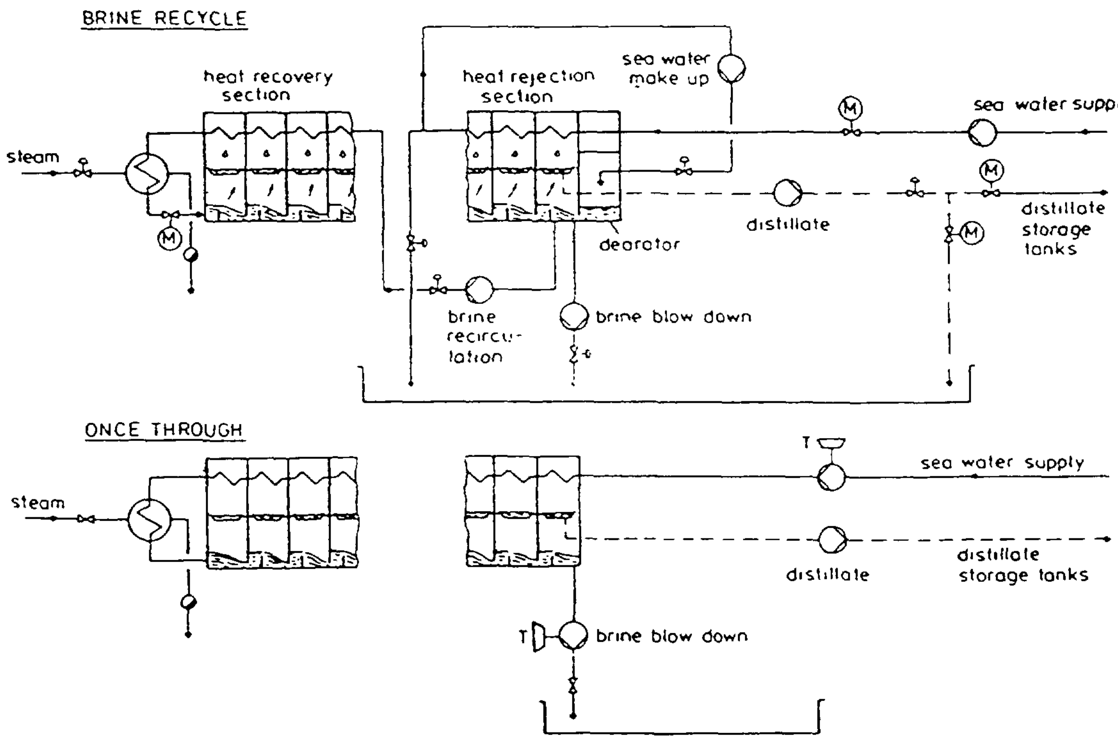


FIG. 5. Comparison of MSF-OT and MSF-BR [10].

removing a certain amount of brine blow-down and adding make-up water. In this way the amount of acid chemicals against scaling can be reduced, and carbon steel with a high corrosion allowance can be used due to the absence of oxygen in the make-up water.

Today, corrosion resistant materials are available at reasonable costs as well as high temperature, cost effective antiscalants. Therefore, MSF-OT systems, in which the feedwater is directly taken from the sea without brine recycling, have already successfully been applied [10]. In MSF systems, the deaeration of the feedwater occurs in the first stage, and additives are injected before the feedwater enters the plant. The main advantages of MSF-OT systems over MSF-BR systems are:

- savings of equipment (pumps, valves and other armatures) and of pumping energy because of leaving out the brine recycle loop and the heat rejection section;

- savings in heat transfer area and/or thermal energy consumption because of the lower boiling point elevation in each stage (lower salinity of the flashing brine);

- reduced risk of calcium sulphate scaling due to the lower salt concentration levels, which also permits a higher maximum brine temperature.

Today, MSF plants have reached a mature and reliable stage of development. Unit sizes up to 60 000 m^3/d have been built. The thermal heat and electricity consumption is in the range of 45 to 120 $kW(th)\cdot h/m^3$ and 3.0 to 6.0 $kW(e)\cdot h/m^3$ respectively. Expressed in exergy units ($kW\cdot h/m^3$), the total consumption is in the range of 15 to 24 $kW\cdot h/m^3$. Using polymeric anti-scaling additives, the maximum brine temperature is limited to 120°C for MSF-BR systems and 135°C for MSF-OT systems due to scaling problems.

2.2.2. Multiple effect distillation (MED)

The MED process is the oldest large scale distillation process. From the thermodynamic point of view, MED processes are superior to MSF processes due to a lower total exergy consumption. This can be illustrated in comparing the GOR of MED plants with the GOR of MSF plants with identical heat transfer area and the same temperature difference between heat source and cooling water sink. The GOR of MED plants is much higher. In spite of this superiority, the MED process could not compete with the MSF process in the past. The main reasons for this may be traced to the components and materials used, as well as the lack of experience in large scale MED plant operation.

Figure 6 shows the schematic flow diagram of MED process using horizontal tube evaporators. In each effect heat is transferred from the condensing water vapour on one side of the tube bundles to the evaporating brine on the other side of the tubes. This process is repeated successively in each of the effects at progressively lower pressure and temperature, driven by the water vapour from the preceding effect. In the last effect at the lowest pressure and temperature the water vapour condenses in the heat rejection heat exchanger, which is cooled by incoming seawater. The condensed distillate is collected from each effect. Some of the heat in the distillate may be recovered by flash evaporation to a lower pressure (not illustrated in Figure 6). As a heat source, low pressure saturated steam is used, which is supplied by steam boilers or dual-purpose plants (co-generation of electricity and steam).

According to the direction of vapour and brine flow, MED plants are subdivided into "forward feed" and "backward feed" arrangements. In forward feed MED plants, vapour and brine move through the evaporators as parallel flows from the first high-pressure evaporator to the last low-pressure one (see Figure 6). The pre-heating of feedwater occurs in separate heat exchangers. In backward feed MED plants, vapour and brine move through the evaporators in opposite directions, whereby, separate feedwater preheating is eliminated.

Currently, the most dominant MED processes with the highest technical and economic potential are the low temperature horizontal tube multi-effect process (LT-HTME) (Figure 7) and the vertical tube evaporation process (VTE) (Figure 8).

The main differences between LT-HTME plants and VTE plants are in the arrangement of the evaporation tubes, the side of the tube where the evaporation takes place and the evaporation tube materials used. In LT-HTME plants, evaporation tubes are arranged horizontally and evaporation occurs by spraying the brine over the outside of the horizontal tubes creating a thin film from which steam evaporates. In VTE plants, evaporation takes

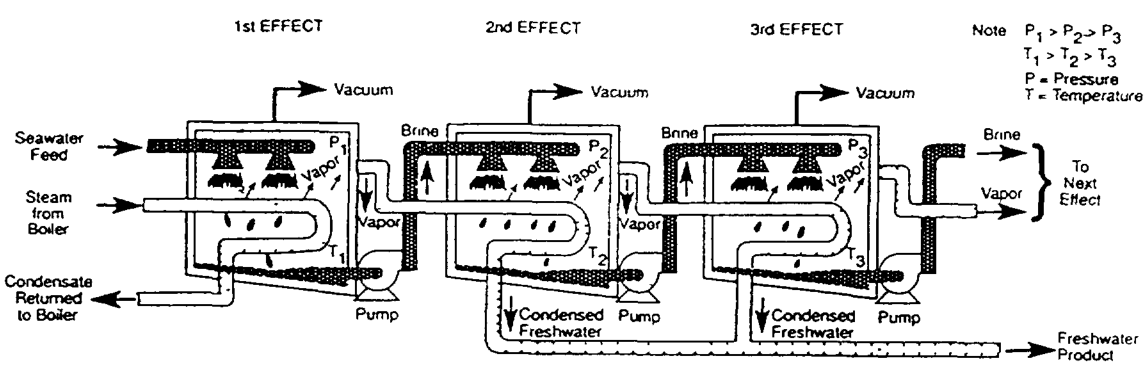


Diagram of a multiple effect plant.

FIG 6 Schematic flow diagram of an LT-HTME plant [8]

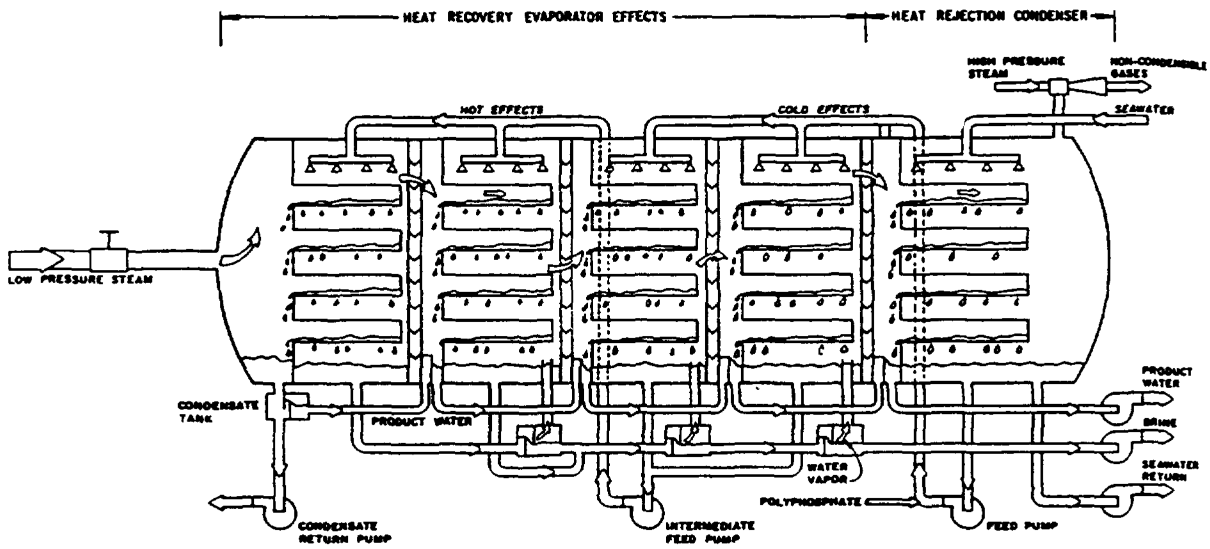


FIG. 7. Principle flow diagram of a LT-HTME unit [8].

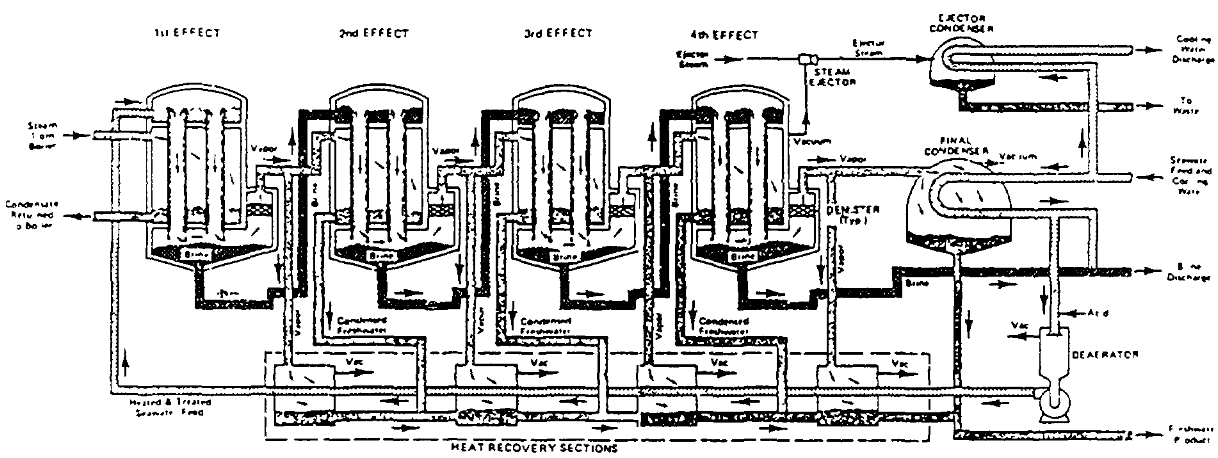


FIG. 8. Principle flow diagram of a HT-VTE unit [8].

place inside vertical tubes. Furthermore, in LT-HTME plants the maximum brine temperature is limited to 70°C, since low cost materials such as aluminium for heat exchanger and carbon steel as shell material are used.

MED plants have a much more efficient evaporation heat transfer process than MSF plants. Due to the thin film evaporation of brine on one side of the tubes and the condensation of vapour on the other side, high heat-transfer coefficients are achieved. Consequently, the number of effects for a certain temperature difference between heat source and cooling water sink can be increased in comparison to MSF plants, thus decreasing the specific heat consumption.

In some MED designs, a part of the vapour produced in the last effect is compressed to a higher temperature level so that the energy efficiency of the MED plant can be improved (Vapour Compression). To compress the vapour, mechanical compressors (isentropic efficiency: about 80%) or steam-jet ejectors (isentropic efficiency: less than 20%) are employed. These designs, however, are usually not applied in integrated plants for electricity and potable water production.

The pre-treatment of seawater for MED plants is similar to that in MSF plants. In general, polyphosphate is introduced into the seawater feed to prevent calcium carbonate scale formation on the heat transfer tubes. A steam jet-ejector vacuum system is used to remove vent gases from the deaerator and non-condensable gases evolving during evaporation from the system. Some LT-HTME designs need a more stringent filtration of the seawater feed, as a result of the small nominal diameters of the brine distribution devices, which do not permit the presence of relatively large suspended particles in seawater.

Table IV shows some technical data of typical commercial MED plants [9, 11]

TABLE IV. TECHNICAL DATA OF MED PLANTS (WITHOUT VAPOUR COMPRESSION)

		LT-HTME	VTE
Maximum brine temperature	°C	70*	135
GOR	1	4-13.5	4-21
Number of effects	1	5-18	5-28
Thermal heat consumption	$kW(th)\cdot h/m^3$	48-160	25-160
Electricity consumption	$kW(e)\cdot h/m^3$	1.2-3.5	0.9-4.5
Total exergy consumption **	$kW\cdot h/m^3$	9-14	9-14

* since low cost materials are used.

** supplied with low-pressure saturated steam of power plants.

2.3. REVERSE OSMOSIS (RO)

Reverse osmosis is a membrane separation process in which pure water is "forced" out of a concentrated saline solution by flowing through a membrane at high static transmembrane pressure differences. These pressure differences have to be higher than the osmotic pressure between the solution and the pure water. In practice, seawater has to be compressed up to 70 to 80 bar since the osmotic pressure of the saline solution is about 60 bar, whereas the osmotic pressure of the permeate is negligible.

The saline feed is pumped into a closed vessel where it is pressurized against the membrane. As a portion of the water passes through the membrane, the salt content in the remaining feed water increases. At the same time, a portion of this feed water is discharged without passing through the membrane.

RO membranes are made in a variety of modular configurations. Two of the commercially successful configurations are spiral-wound modules (Figure 9) and hollow fibre modules (Figure 10). In both of these configurations, module elements are serially connected in pressure vessels (up to 7 in spiral wound modules and up to 2 in hollow fibre modules).

A spiral wound module element, illustrated in Figure 9, consists of two membrane sheets supported by a grooved or porous support sheet. The support sheet provides the pressure support for the membrane sheets as well as providing the flow path for the product water. Each sheet is sealed along three of its edges, and the fourth edge is attached to a central product discharge tube. A plastic spacer sheet is located on each side of the membrane

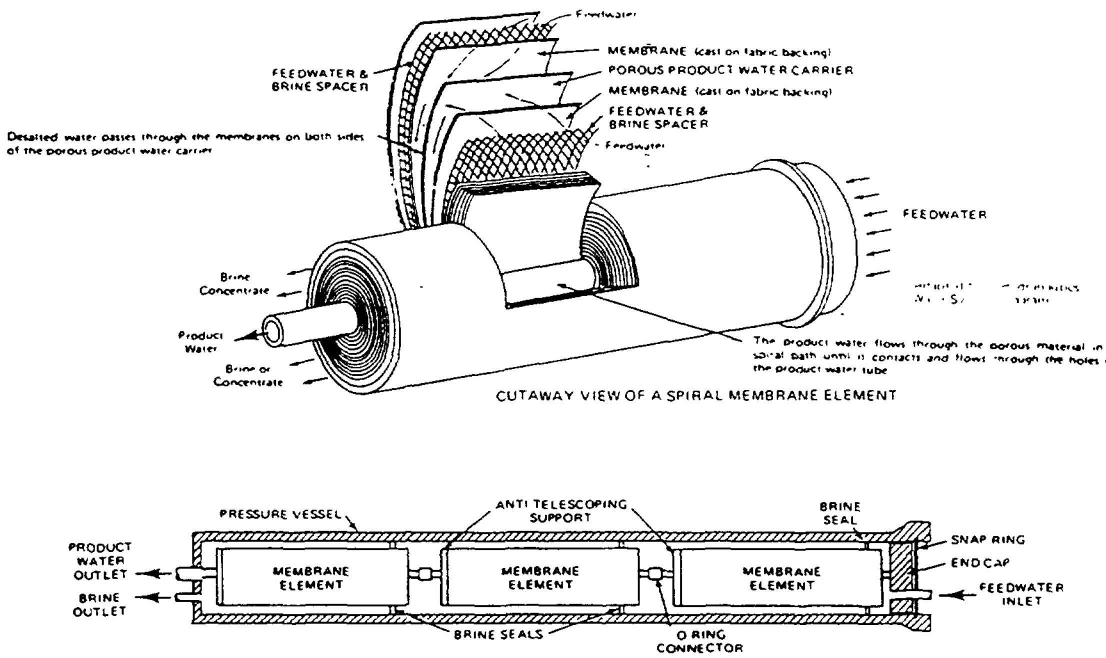


FIG. 9. Schematic diagram of a spiral wound membrane module [8].

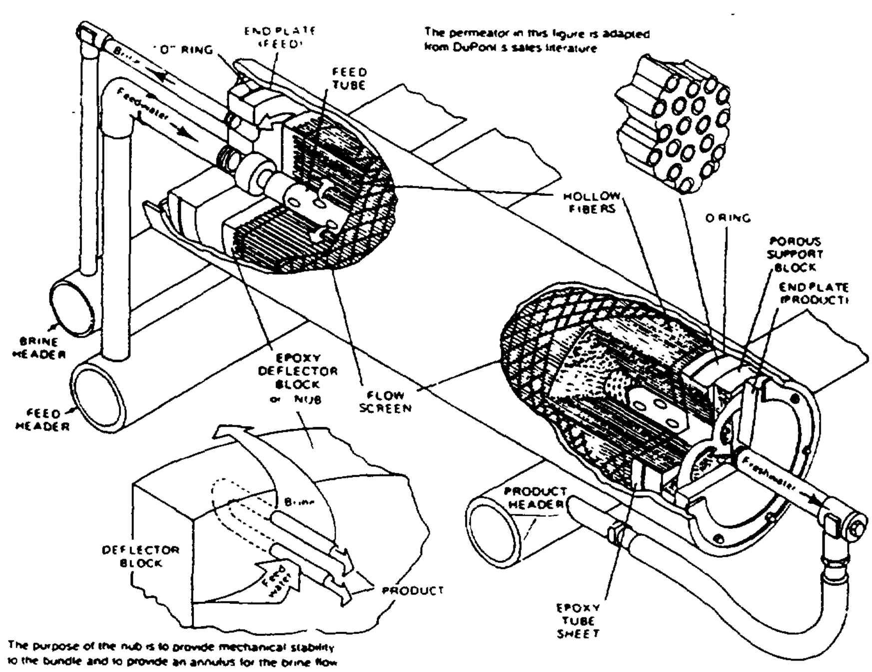


FIG. 10. Schematic diagram of a hollow fibre membrane module [8].

assembly sheets, and the spacer sheets provide the flow channels for the feed flow. The entire assembly is then spirally wrapped around the central discharge tube forming a compact RO module element.

The recovery ratio (permeate flow rate divided by the feed flow rate) of spiral wound module elements is very low so that up to 7 elements are arranged in one module to get a higher overall recovery ratio (see Figure 9). Spiral wound membranes have a simple design (reasonable production costs) with a relatively high resistance to fouling. Spiral wound membranes are currently operated at pressures as high as 69 bar and recovery ratios up to 45%. Spiral wound membranes which can operate at pressures as high as 82.7 bar are already commercially offered [12]. Hollow fibre membranes are made of hair-like fibres which are united in bundles and arranged in pressure vessels. Typical configurations of hollow fibre modules are U-tube bundles, similar to shell and tube heat exchangers. The feed is introduced along a central tube and flows radially outward on the outside of the fibres. The pure water permeates the fibre membranes and flows axially along the inside of the fibres to a "header" at the end of the bundle (Figure 10).

Typical outside diameters of hollow fibres are somewhere in the order of 85 μm to 200 μm. Hollow fibres can withstand pressures as high as 82.7 bar and have high recovery ratios up to 55%.

The following membrane materials are currently used for seawater RO membranes:

- cellulose acetate membranes,
- polyamide membranes
- thin film composite membranes.

The choice of a suitable membrane material is particularly influenced by its resistance to free chlorine, free oxygen, temperature, bacteria and to the index of pH of the saline solution (Table V).

Table V shows why cellulose acetate membranes have been playing an important part in seawater desalination. Although strongly limited in index of pH, the advantages are low material costs and the resistance to chlorine, which is used in feedwater to inhibit biological fouling. Cellulose acetate membranes have a relatively short operating life and suffer pressure compaction (deterioration of permeate water flow because of creep-buckling of the membrane material at high pressure and high temperature).

Polyamide and thin film composite membranes have, in general, higher water fluxes and higher salt rejections than cellulose acetate membranes. However, these types of membranes are subject to chlorine attack. If chlorine is added to feedwater to control biological growth, the feedwater must be dechlorinated before entering the membrane modules.

Thin film composite membranes consist of two layers of different polymers: one relatively thick and porous layer (e.g. polysulfone) which provides the membrane support, and one relatively thin (about 0.05-0.1 μm) and dense layer (e.g. polyamine) which provides the semi-permeable characteristics. The different materials of the layers make it possible to

TABLE V. MEMBRANE DAMAGING CONDITIONS [13]

	Cellulose acetate membrane	Polyamide membrane	Composite membrane
Index of pH	4-6	4-11	3-11
Free chlorine	< 1 mg/l	pH $\leq$ 8: < 0.1 mg/l	unstable
	(shortly up to 5 mg/l)	pH > 8: < 0.25 mg/l	
Bacteria	unstable	unstable	tolerant
Free oxygen	tolerant	tolerant	partly tolerant

optimize each layer separately which results in higher water fluxes and higher salt rejections at high mechanical strength in contrast to membranes consisting of only one material.

A disadvantage of RO is the need for significant pre-conditioning of the feedwater to protect the membranes. The extent of pre-treatment requirements depends on a variety of factors, such as seawater composition and temperature, seawater intake, membrane materials and recovery ratio. RO pre-treatment includes the following steps:

- chlorine disinfection to prevent biological growth in feed water,
- coagulation followed by one of the mechanical separation methods (sedimentation, filtration, flotation) to remove colloidal and suspended matter from the feedwater,
- conditioning with acids to adjust the index of pH for carbonate scale suppression and with inhibitors (polyphosphates) to prevent sulphate scale formation.

For chlorine sensitive membranes, in addition, feed de-chlorination through activated carbon filters and/or sodium bisulphate dosage is required.

Since the overall recovery ratios of current seawater RO plants are only 30 to 50%, and since the pressure of the discharge brine is only slightly less than the feed stream pressure, all large-scale seawater RO plants as well as many smaller plants are equipped with energy recovery turbines, usually Pelton turbines, which recover a part of the pumping energy.

High salt rejection and good high pressure operation qualities of current membranes permit the economical operation of seawater RO plants in single-stage systems, even on the high salt content waters found in the Middle East while producing drinking water in accordance to World Health Organisation (WHO) standards.

Figure 11 illustrates the simplified flow diagram of single-stage RO plant consisting of multiple RO trains arranged in parallel.

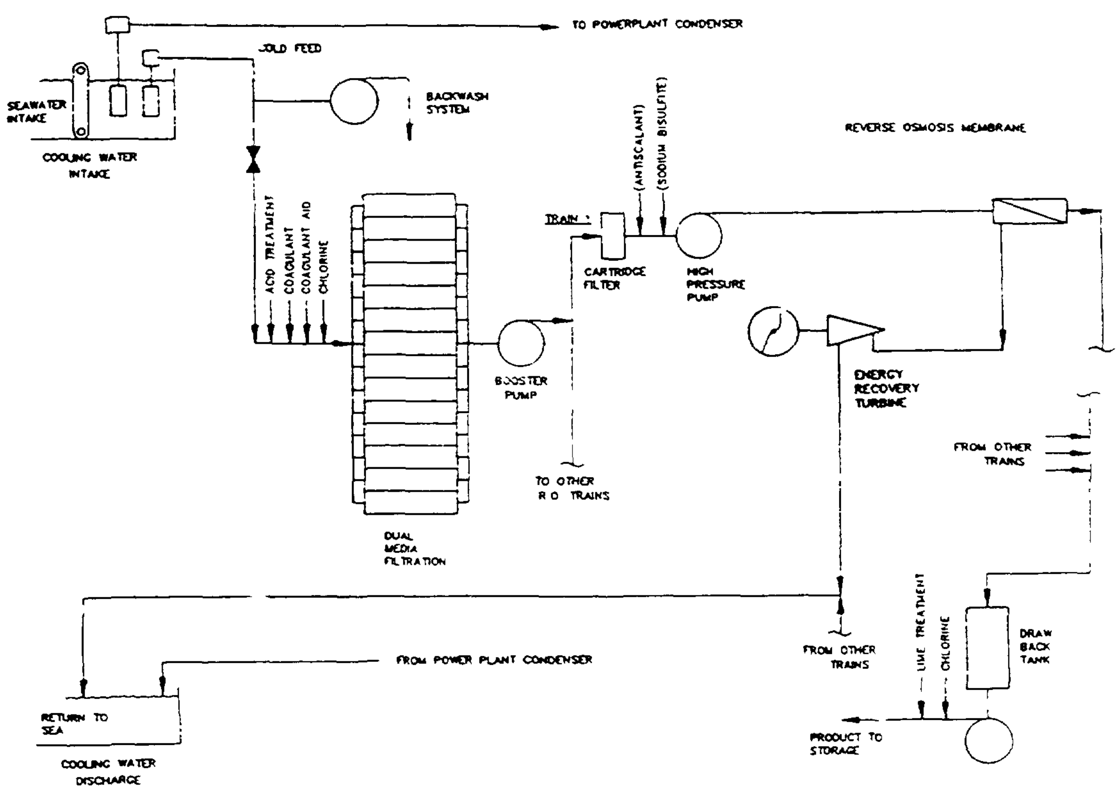


FIG. 11. Principle flow diagram of an RO plant with multiple trains [14].

In recent years, seawater RO has become a reliable and commercial process applicable on a large-scale. A weak point in RO operation is the low tolerance of membranes to operational errors, which has led in the past to high membrane replacement costs in some cases.

Typical electricity consumption of RO plants is in the range of 4 to 7 $kW(e) \cdot h/m^3$ dependent on the seawater salinity, recovery ratio, required permeate quality, plant configuration and implementation of energy recovery in the brine blow down.

3. ECONOMIC RANKING AND COST ALLOCATION FOR INTEGRATED PLANTS

As mentioned in Section 2, seawater distillation plants require low temperature steam as the heat source since the maximum brine temperature cannot be substantially raised above 120 to 130°C (corrosion, scaling). From the thermodynamic point of view, it is thus compelling to use steam from an adjacent power station, in which the high pressure steam is first used to produce electricity, and the exhaust from the turbine serves as the heat source for the distillation plant (integrated plant).

In the following, further economic and technical aspects of electricity and potable water co-production plants are identified and discussed. Subsequently, a methodology for economic comparison and ranking of different integrated plant alternatives is presented. Finally, various cost allocation methods for integrated plants are compared. The exergetic cost allocation method is described in more detail.

3.1. INCENTIVES AND DISINCENTIVES FOR CO-LOCATING POWER AND SEAWATER DESALINATION PLANTS

The following are some of the economic and technical incentives and disincentives for co-locating power and seawater desalination plants which are beyond the direct consequences of co-generation of electricity and heat. These aspects are valid for both distillation plants and RO plants.

(1) Possibility of larger unit sizes

Dependent on the electricity demand, relatively large power plants can be installed, which would benefit from the size effect (economics of scale).

(2) Common use of facilities and infrastructure

Co-locating of power and desalination plants provide the opportunity to share facilities which might otherwise have to be duplicated. In this connection, the largest benefit results from sharing of common seawater intake/outfall structures, which provide cooling water for power plants and seawater feed and brine discharge for desalination plants. Furthermore, access ways, maintenance shops, storage facilities, personnel accommodations, loading and receiving facilities, etc., can be shared.

(3) Common operating staff

Some systems and services in the power and desalination plants require similar types of staff, opening the opportunity for staff sharing and consequently savings in personnel costs. This includes in particular administrative and maintenance staff.

(4) Improved dispersion of the power station and seawater desalination plant effluents

By mixing the warm condenser cooling water with the more saline and higher density brine blow down, a nearly neutral buoyancy with the surrounding seawater is obtained. Consequently, a more rapid dispersion of the effluents can be achieved and the risk to the ecosystem can be reduced.

(5) Mutual effects

Any incident interrupting the output of one of the two products may affect the production of the other. It is possible to minimize the impact by adding devices such as back-up heat boilers to supply the distillation plant with heat if the power plant is out of operation, and modifications in the steam power cycle of the power plant to continue operating if the distillation plant is shut down. However, these devices involve additional investments.

(6) Reduced overall flexibility

The maximum benefit from the combined production of water and electricity is attained when the plant is operating under its rated conditions. Certain designs provide for variation in the water-to-electricity ratio, but to the detriment of efficiency or at the cost of extra investment. In any case, the range of possible variation is rather limited, and some flexibility may be lost. In case of nuclear plants which for economic reasons are intended for base load operation, the loss of flexibility does not have a serious adverse impact.

3.2. COMPARISON AND RANKING OF INTEGRATED PLANTS BY THE EQUIVALENT ELECTRICITY GENERATION COST

The economic objective of single purpose plants for power generation or desalination is to achieve the lowest possible production costs per unit. Single purpose plant alternatives can easily be compared and ranked by calculating the production costs, which are obtained by dividing the annual expenditures (capital charges, fuel cost, operation and maintenance costs) related to production by the annual output. For single purpose plant alternatives of the same net output, the comparison of the annual expenditures is sufficient to select the most economic plant.

For integrated plants, however, which have simultaneously two final products, electricity and potable water, the economic comparison and ranking is more difficult. The plant with the least annual overall expenditures (annual expenditures related to both electricity generation and potable water production) is not necessarily the most economic solution, since it is unlikely that all plant alternatives will have exactly the same net electricity rating and desalination capacity. Furthermore, the potable water production costs and the electricity generation cost vary from one alternative to the other, which makes a comparison of different integrated plants difficult.

A methodology has to be defined to economically assess and compare different integrated plants, in which the annual overall expenditure of the plants as well as the outputs of the two final products, electricity and potable water, are considered.

An appropriate method to compare plants with the ***same potable water output*** and — to make a fair comparison — with similar base power plant capacities (when not supplying heat and /or electricity to the desalination plant) is to calculate the so-called "***equivalent electricity generation cost***" c_{eq} (see Equation (2)) where the annual generated net electricity E_a supplied to the grid is charged with the annual overall expenditure C_0:

$$c_{eq} = \frac{C_0}{E_a} \qquad \text{in \$/kW(e)·h.} \qquad (2)$$

In other words, it is arbitrarily assumed that the potable water production is completely subsidized by the electricity generation. The plant alternative with the lowest resulting equivalent electricity cost will be the economically optimal solution.

3.3. COST ALLOCATION METHODS

After selecting the most economic integrated plant for electricity and potable water production, a cost basis for the sale of both final products have to be established. This is important in cases where separate ownership of seawater desalination plant and power station prevail. For this purpose, it is useful to have some equitable techniques for allocating the overall expenditures of the integrated plant to the two final products. Even if it would be finally decided to adopt a very low (subsidized) potable water tariff in order to promote development of a certain area, it is always necessary to know the amount of the subsidy. There are several techniques of allocating costs to electricity and potable water, or to an intermediate product such as steam delivered to the distillation plant. The selection of the most suitable method will depend on the objectives and the environment in which the plant is built.

The cost allocation methods that have so far been proposed or used for co-production of potable water and electricity can be split into two main groups: "cost prorating methods" and "credit methods" (see Table VI) [6]. The credit methods attribute a value to one of the products and obtain the cost of the other by difference. This value could be based either on market conditions or production costs of single purpose plants. The cost prorating methods divide the overall expenditures of the integrated plant according to a given set of rules entailing, in general, a sharing of the benefit of co-production between the two final products.

The *credit method based on market conditions* allocates a market oriented value to one of the products (electricity or potable water) and determine the cost of the other by subtraction from the overall cost of the integrated plant.

The *power credit method* is based on the concept that the electricity equivalent of steam supply (electricity that could have been generated by the steam supplied to the distillation plant) and/or the electricity provided to the seawater desalination plant, could have been sold to the grid, and that this loss in revenues should be charged to the water cost (power credit). The power credit is calculated by multiplying the reduction in electrical output by the unit electricity generation cost of an equivalent single purpose power plant. Applying the power credit method, the potable water produced is credited with all of the economic benefits associated with co-production.

In the *water credit method*, the whole benefit of co-production is assigned to the cost of electricity by using a water credit, the value of which would be equal to the cost of water produced in an alternative least-cost water scheme.

In the *proportional value method*, either the market values of the two products or the production costs of two single purpose plants are determined, the first producing the same

TABLE VI. COST ALLOCATION METHODS FOR CO-PRODUCTION OF POTABLE WATER AND ELECTRICITY [6]

Cost prorating methods:	Credit methods:
Proportional value method	Credit methods based on market conditions
Caloric method	Power credit method
Exergetic method	Water credit method

quantity of potable water and the other supplying the same net amount of electricity to the grid as the integrated plant. The overall cost of the integrated plant is then divided in proportion to the ratio of the values or costs of the two individual products so defined and then allocated to the electricity and potable water respectively.

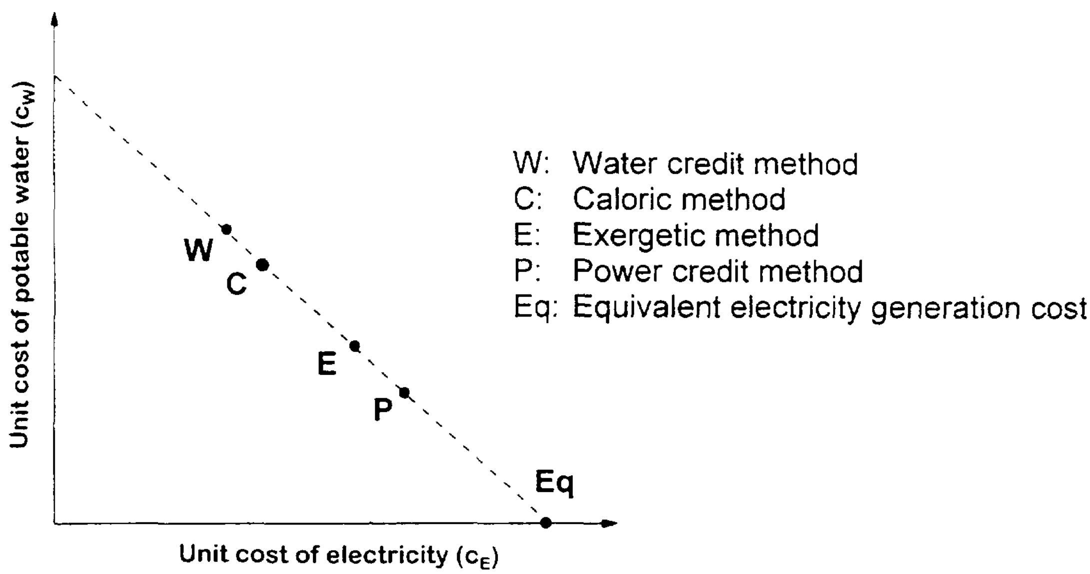
The *caloric method* is based on the First Law of Thermodynamics (law of energy conservation). The method allocates the common production cost of the power station in proportion to the amount of enthalpy used to produce electricity and low temperature steam for the seawater distillation plant respectively.

Figure 12 shows the qualitative relation of the electricity generation and potable water production cost of an integrated plant obtained by the various cost allocation methods described above.

To share the benefit of co-production of electricity and potable water, the cost allocation method chosen should result in points somewhere located inside the line segment W-P in Figure 12. That is, the water credit method and the power credit method are not an equitable cost allocation method, since one of the final products have no share in the benefit.

With regard to the proportional value method and the credit method based on market conditions, the disadvantage is that only market oriented criteria are considered. Therefore, the thermodynamic capability of the integrated plant in producing electricity and potable water is not covered adequately.

The caloric method covers some process-specific thermodynamic criteria of the integrated plant. However, there is no adequate assessment of the thermodynamic value (exergy) to be assigned to the energy flows required to produce electricity and potable water. As a result, from the thermodynamic viewpoint as well as considering the sharing of the benefit, the *exergetic cost allocation method* is the most equitable cost allocation method with a global applicability.



Note: points obtained with the proportional value method and the credit method based on market conditions could be anywhere on the line, dependent on market values

FIG. 12. Qualitative example of the electricity generation and potable water production cost allocation of an integrated plant.

In the following, the exergetic cost allocation method for integrated plants is described and applied to a nuclear power plant using a pressurized water reactor (PWR) and a gas fired combined cycle power station as alternative energy sources. This allocation method is applicable both for distillation plants and RO plants.

3.4. EXERGETIC COST ALLOCATION METHOD

The method is based on the concept of exergy. By definition, exergy is the part of energy transferable to any other form of energy under given thermodynamic conditions [15]. The remaining part of energy is called "anergy". The exergy method takes into account both the First and the Second Laws of Thermodynamics. It is assessed that the value of mechanical and electrical energy is higher than the value of heat, and that the transformation of heat into any other form of energy is accompanied by losses. Before describing the exergetic cost allocation method, some explanations describing exergy are given below.

3.4.1. Exergy

For a given thermodynamic process:

$$\sum_{process} dE \leq 0 . \tag{3}$$

This definition of exergy E is equivalent to the classical statement that the amount of exergy loss E_l is directly related to the irreversible amount of entropy generation:

$$E_l = T_0 \cdot S_P \ , \tag{4}$$

where T_0 represents the temperature of the reference surroundings and S_P is the entropy production.

Exergy is the maximum mechanical work derivable from a system and its surroundings in bringing the system from its present thermodynamic state to a state of complete, stable equilibrium with the surroundings, mathematically represented by the equation:

$$E = H - T_0 \cdot S - \sum_i \mu_{i0} \cdot m_i \ , \tag{5}$$

where H denotes enthalpy of the system , S entropy, m mass and μ_0 chemical potential at reference surroundings.

Although thermodynamic analyses have been traditionally based on energy and the First Law of Thermodynamics, it is exergy that accurately evaluates a system's performance. Energy can not be produced or destroyed; therefore, it is non-depletable. During all real processes, however, entropy is produced, and hence, some of the exergy of the associated energy is lost.

Exergy is the commodity of value to all energy users. When exergy is converted from one form to another, only part of the exergy is transferred to the new form; the remainder is actually lost in order to cause the change. Thus, an exergy evaluation describes how a fuel's potential of producing mechanical work (exergy of fuel) is being used and where the losses of that potential occur. This description also identifies the subsystems for which improvements should be sought.

For stationary open systems, the exergy balance equation can be written in the following form:

$$\sum_{inlet} \dot{E}_j - \int\left(1-\frac{T_0}{T}\right)d\dot{Q} = \sum_{outlet} \dot{E}_j + \dot{W} + \dot{E}_l \ . \tag{6}$$

That is, the sum of exergy associated with matter entering the system and the exergy associated with the net rate of heat addition (indicated by the second term) is equal to the sum of exergy associated with matter leaving the system, the net rate of mechanical work $\dot{W}$ delivered by the system, and the net rate of exergy losses $\dot{E}_l$ (a measure of process irreversibilities).

In steam power cycles, the exergy $\dot{E}_j$ of a steam/water flow j can be calculated by:

$$\dot{E}_j = \dot{m}_j \cdot \left\{\left(h_j - h_0\right) - T_0 \cdot \left(s_j - s_0\right)\right\} , \tag{7}$$

where

h	is the specific enthalpy in kJ/kg,
s	is the specific entropy in kJ/kg/K,
T	is the absolute temperature in K,
0	is the subscript which denotes the state of surroundings, and
$\dot{m}$	is the mass flow rate in kg/s.

T_0 is usually the ambient seawater temperature.

3.4.2. Exergetic allocation of the overall production cost

The overall annual expenditures of integrated plants are the annual costs ($/a) that arise in producing the two final products, potable water and electricity. The overall expenditures are made up of the fixed expenditures and the variable expenditures. The fixed expenditures include all costs that occur independently of the quantity of the final products, such as capital charge, personnel cost, insurance and preventive maintenance cost. Variable expenditures contain expenses that occur in proportion to producing the final products, such as fuel cost and consumable operating materials cost.

In the exergetic cost allocation method, the overall expenditures $\mathbf{C_0}$ of the integrated plant are divided into the following cost components:

- electricity generation expenditures C_{L_c}, allocated exclusively to the generation of electricity supplied both to the grid and to the seawater desalination plant;

- steam production expenditures for providing heat to the desalination plant C_S, allocated exclusively to the production of potable water;

- common electricity and steam production expenditures C_C; and

- remaining water production expenditures C_{W^*}.

$$C_0 = C_{E_0} + C_{S_0} + C_C + C_{W^*} .\qquad (8)$$

The common electricity and steam production expenditures C_C are allocated to the two forms of energy produced, electricity and steam, proportional to the exergy flows $\dot{E}_E$ and $\dot{E}_S$ (exergy loss flows or exergy consumption flows, see Section 3.4.2.1 and Section 3.4.2.2) that are required to produce these two energy forms. Hence, the electricity generation expenditures C_{E^*} to generate electricity and the steam production expenditures C_S are calculated by Eqs (9) and (10):

$$C_{E^*} = C_{E_0} + \frac{\dot{E}_E}{\dot{E}_E + \dot{E}_S} \cdot C_C . \qquad \text{(electricity)} \qquad (9)$$

$$C_S = C_{S_0} + \frac{\dot{E}_S}{\dot{E}_E + \dot{E}_S} \cdot C_C . \qquad \text{(steam)} \qquad (10)$$

C_{E^*} is further divided into expenditures for generation of saleable power C_E and generation of electricity supplied to the seawater desalination plant C_{E_W}, proportional to the saleable electricity P_E supplied to the grid and the electricity P_W supplied to the seawater desalination plant (see Eqs (11) and (12), where P_{net} is the electrical output of the power plant):

$$C_E = C_{E^*} \cdot \frac{P_E}{P_{net}} , \qquad (11)$$

$$C_{E_W} = C_{E^*} \cdot \frac{P_W}{P_{net}} . \qquad (12)$$

Finally, the water production expenditures C_W are calculated by:

$$C_W = C_{W^*} + C_{E_W} + C_S . \qquad (13)$$

Dividing C_E and C_W by the respective units produced, leads to the electricity generation cost c_E and the potable water production cost c_W expressed in \$ per kW(e)·h and \$ per m^3 respectively.

3.4.2.1 Exergetic cost allocation method using a PWR power plant as energy source

In the following, the exergetic cost allocation method is illustrated using a PWR power plant as energy source.

The composition of the four individual cost components defined in Eq. (8) is listed in Table VII.

The exergy flows $\dot{E}_E$ and $\dot{E}_S$ are the shares of the exergy of fuel $\dot{E}_F$ supplied to the power plant, which are required to produce electricity and steam respectively. They consist of the exergy flows of the two products themselves (net electrical output and exergy flow of steam respectively) and a share of the exergy loss flows occurring in the power plant, which are allocated to the two products according to a given set of rules described below.

TABLE VII. COMPOSITION OF THE INDIVIDUAL COST COMPONENTS OF AN INTEGRATED PLANT WITH A PWR AS ENERGY SOURCE

C_T	capital charge of turbogenerator equipment
C_S	capital charge of incremental equipment for providing steam to distillation plant
C_C	remaining capital charge of PWR power plant, fuel cost of PWR, decommissioning cost of PWR power plant, fixed and variable operation & maintenance (O&M) cost of PWR power plant
C_{W^*}	capital charge of desalination plant and backup heat source, fixed and variable O&M cost of desalination plant and backup heat source, fuel cost of backup heat source

TABLE VIII. EXERGY ANALYSIS OF A PWR POWER PLANT

	$\dot{E}_F$	exergy of fuel
	=	
E_C	$\dot{E}_{SG}$	exergy losses in primary circuit [a] including reactor, steam generator and reactor coolant pumps
	+	
	$\dot{E}_{MSR}$	exergy losses in moisture separators and steam reheaters
	+	
	P_{Aux}	electrical auxiliary loads [b]
	+	
	$\dot{E}_{FH}$	exergy losses in feedwater heaters
	+	
	$\dot{E}_{FP}$	exergy losses in feedwater pumps
	+	
$\dot{E}_{E}$	E_T	exergy losses in turbines
	+	
	$\dot{E}_{Con}$	exergy losses in condenser
	+	
	$\dot{E}_G$	exergy loss in generator and mechanical losses
	+	
	P_{net}	net electrical output
	+	
	$\dot{E}_S$	exergy of steam provided for distillation plants

a) mainly associated with the fission process and heat transport
b) with the exception of feedwater pumps and reactor coolant pumps

$\dot{E}_T$ and $\dot{E}_S$ are determined by analysing the PWR power plant presented in Table VIII. In this table, the exergy flows summarised in $\dot{E}_E$ ($\dot{E}_T$, $\dot{E}_{Con}$, $\dot{E}_G$, P_{net}) are allocated exclusively to the generation of electricity; $\dot{E}_S$ is allocated exclusively to the production of steam. The exergy flows summarised in E_C (E_{SG}, $\dot{E}_{MSR}$, P_{Aux}, E_{FH}, E_{FP}), which can be assigned to the generation of electricity as well as to the production of steam, are allocated to the two products proportional to $\dot{E}_{L_e}$ and $\dot{E}_{S_e}$. Based on these considerations, $\dot{E}_L$ and $\dot{E}_S$ are calculated by Eqs (14) and (15) respectively:

$$\dot{E}_L = \dot{E}_{L_e} + \dot{E}_C \cdot \frac{\dot{E}_{L_e}}{\dot{E}_{L_e} + \dot{E}_{S_e}}, \tag{14}$$

$$\dot{E}_S = \dot{E}_{S_e} + \dot{E}_C \cdot \frac{\dot{E}_{S_e}}{\dot{E}_{L_e} + \dot{E}_{S_e}}. \tag{15}$$

The electrical power requirements of feedwater pumps and reactor coolant pumps are not separately listed in Table VIII, since they are proportionally covered in the individual exergy flows.

3.4.2.2. Exergetic cost allocation method using a combined cycle power plant as energy source

The composition of the four individual cost components of an integrated plant with a combined cycle power plant as energy source is listed in Table IX.

The determination of $\dot{E}_L$ and $\dot{E}_S$ of combined cycle power plants is more complicated than for PWR power plants. Separate exergy analyses of the gas turbine cycle and the steam cycle are necessary.

In the first step, exergy of fuel $\dot{E}_F$ is allocated to the exergy flows $\dot{E}_{L_{GT}}$ and $\dot{E}_{HRSG}$ that are required to generate electricity in the gas turbine and to provide heat in the heat recovery steam generator respectively (see Table X). The exergy flows summarized in $\dot{E}_{L_{e,GT}}$ ($\dot{E}_{T_{GT}}$, $\dot{E}_{CO}$, $\dot{E}_{C_{GT}}$, $P_{net_{GT}}$) are allocated exclusively to $\dot{E}_{L_{GT}}$; $\dot{E}_{HRSG_e}$ is allocated exclusively to $\dot{E}_{HRSG}$. The exergy flows summarized in $\dot{E}_{C_{GT}}$ ($\dot{E}_{CC}$, $\dot{E}_{EG}$, $P_{Aux_{GT}}$), which can be assigned to $\dot{E}_{L_{GT}}$ as well as to $\dot{E}_{HRSG}$ are allocated to the two products proportional to $\dot{E}_{L_{e,GT}}$ and $\dot{E}_{HRSG_e}$. Based on these considerations, $\dot{E}_{L_{GT}}$ and $\dot{E}_{HRSG}$ are calculated by Eqs (16) and (17) respectively:

$$\dot{E}_{L_{GT}} = \dot{E}_{L_{e,GT}} + \dot{E}_{C_{GT}} \cdot \frac{\dot{E}_{L_{e,GT}}}{\dot{E}_{L_{e,GT}} + \dot{E}_{HRSG_e}}, \tag{16}$$

$$\dot{E}_{HRSG} = \dot{E}_{HRSG_e} + \dot{E}_{C_{GT}} \cdot \frac{\dot{E}_{HRSG_e}}{\dot{E}_{L_{e,GT}} + \dot{E}_{HRSG_e}}. \tag{17}$$

In the second step, the exergy flows of the steam cycle are analysed (see Table XI). The allocation of $\dot{E}_{HRSG_e}$ to the production of electricity in the steam turbines $\dot{E}_{L_{ST}}$, and to the production of steam $\dot{E}_S$, occur in the same way as in Eqs (14) and (15). Only the exergy of fuel $\dot{E}_F$ is substituted by $\dot{E}_{HRSG_e}$, and the exergy losses in the feedwater heaters are included in the exergy losses in the heat recovery steam generator:

TABLE IX. COMPOSITION OF THE INDIVIDUAL COST COMPONENTS OF AN INTEGRATED PLANT WITH A GAS FIRED COMBINED CYCLE POWER PLANT AS ENERGY SOURCE

C_I	capital charge of compressor, gas turbine and generator of gas turbine cycle, capital charge of turbogenerator equipment of steam cycle
C_S	capital charge of incremental equipment for providing steam to distillation plant
C_C	remaining capital charge of combined cycle power plant, fuel cost of combined cycle power plant, fixed and variable O&M cost of combined cycle power plant
C_{W*}	capital charge of desalination plant and backup heat source, fixed and variable O&M cost of desalination plant and backup heat source, fuel cost of backup heat source

TABLE X. EXERGY ANALYSIS OF THE GAS TURBINE CYCLE OF A COMBINED CYCLE POWER PLANT

	$\dot{E}_F$	exergy of fuel
	$=$	
	$\dot{E}_{CC}$	exergy losses in combustion chamber
	$+$	
$\dot{E}_{C_{GT}}$	$\dot{E}_{EG}$	exergy of exhaust gas leaving the heat recovery steam generator
	$+$	
	$P_{Aux_{GT}}$	electrical auxiliary loads of gas turbine
	$+$	
	$\dot{E}_{T_{GT}}$	exergy losses in gas turbine (including cooling losses)
	$+$	
	$\dot{E}_{Co}$	exergy losses in compressor
$\dot{E}_{L_{GT}}$	$+$	
	$\dot{E}_{G_{GT}}$	exergy loss in generator and mechanical losses of gas turbine
	$+$	
	$P_{net_{GT}}$	net electrical output of gas turbine
	$+$	
	$\dot{E}_{HRSG_c}$	transferred exergy in heat recovery steam generator

$$\dot{E}_{L_{ST}*} = \dot{E}_{L_{ST}} + \dot{E}_{C_{ST}} \cdot \frac{\dot{E}_{L_{ST}}}{\dot{E}_{L_{ST}} + \dot{E}_{S_c}}, \tag{18}$$

$$\dot{E}_{S*} = \dot{E}_{S_c} + \dot{E}_{C_{ST}} \cdot \frac{\dot{E}_{S_c}}{\dot{E}_{L_{ST}} + \dot{E}_{S_c}}. \tag{19}$$

TABLE XI. EXERGY ANALYSIS OF THE STEAM CYCLE OF A COMBINED CYCLE POWER PLANT

	$\dot{E}_{HRSG_C}$	transferred exergy in heat recovery steam generator
	$=$	
	$\dot{E}_{L_{HRSG}}$	exergy losses in heat recovery steam generator
	$+$	
$\dot{E}_{C_{ST}}$	$\dot{E}_{O_{ST}}$	other exergy losses in the steam cycle
	$+$	
	$P_{Aux_{ST}}$	electrical auxiliary loads of steam cycle
	$+$	
	$\dot{E}_{T_{ST}}$	exergy losses in steam turbines
	$+$	
	$\dot{E}_{Con}$	exergy losses in condenser
$\dot{E}_{L_{C,ST}}$	$+$	
	$\dot{E}_{G_{ST}}$	exergy loss in generator and mechanical losses in steam turbines
	$+$	
	$P_{net_{ST}}$	net electrical output of steam turbines
	$+$	
	$\dot{E}_{S_C}$	exergy of steam provided for distillation plants

Subsequently, the allocation of $\dot{E}_{HRSG}$ to $\dot{E}_{L_{ST}}$ and $\dot{E}_S$ is obtained by Eqs (20) and (21):

$$\dot{E}_{L_{ST}} = \dot{E}_{HRSG} \cdot \frac{\dot{E}_{L_{ST^*}}}{\dot{E}_{L_{ST^*}} + \dot{E}_{S^*}}, \tag{20}$$

$$\dot{E}_S = \dot{E}_{HRSG} \cdot \frac{\dot{E}_{S^*}}{\dot{E}_{L_{ST^*}} + \dot{E}_{S^*}}. \tag{21}$$

Finally, both the exergy flows for generating electricity in the gas turbine and the steam turbine are added up:

$$\dot{E}_L = \dot{E}_{L_{GT}} + \dot{E}_{L_{ST}}. \tag{22}$$

4. ECONOMIC ASSESSMENT FOR A REPRESENTATIVE SITE

The specific costs of integrated co-production plants for potable water and electricity depend very much on the size of both the power plant and the seawater desalination plant (economics of scale), the ratio of electricity and water production, their variation throughout the year, and local conditions at the site.

Local conditions at the site contain factors such as site infrastructure, engineering requirements, local sources of equipment, material and energy, qualification of local construction and operating staff, composition and temperature of seawater, and the financial situation of the country/utility.

The demand for electricity could vary greatly throughout the year as air-conditioning may be the dominant load in the summer months as it occurs in the Gulf, whereas water demand is usually much more stable. When seawater distillation is applied to produce the required potable water, the integrated plant is expected to produce baseload electricity. The peak electricity demand will usually be supplied by peak load power stations such as gas turbines.

From the thermodynamic and economic points of view, it is useful to drive seawater distillation processes with low temperature and pressure (low grade) exhaust steam. The use of higher grade (higher exergy) steam would lead to a higher water production cost, and also to a substantial reduction in thermal efficiency of the power plant. In this connection, PWR power plants provide much greater quantities of low grade steam for seawater distillation than combined cycle power stations with the same net electrical output.

A comparative assessment of all possible energy sources for co-production of electricity and potable water requires comparing a wide range of available options, including nuclear power, fossil fuels, renewable energies, waste recovery, etc. [16, 17]. Within the limited scope of this study, only one nuclear and one fossil power plant type which seem to be the currently most interesting are considered.

In the following, the economic and thermodynamic considerations presented in Section 3 are illustrated for a representative site on the Arabian Peninsula. The site conditions considered, in particular the potable water and electricity demand, are typical for operating and planned integrated plants in this area. A PWR power plant and a gas fired combined cycle power plant were selected as technically and economically viable energy sources for these site conditions. Various seawater desalination processes and types, all with the same net water output, which are preselected on the basis of their favourable technical and economic characteristics and their commercial availability are considered to be coupled with the energy sources. The plants are assumed to be base-loaded. For calculating and comparing the costs of different plant options, the constant money levelized cost methodology is used (see Section 4.5.1). Costs related to water storage, transport and distribution to the consumer are not covered in the assessment.

The integrated plant configurations with the ***lowest equivalent electricity generation cost*** are determined, which would correspond to the economically optimal plant configuration for the site conditions assumed. Furthermore, the ***potable water production cost*** as well as the ***electricity generation cost*** of each plant configuration applying the exergetic cost allocation method are determined.

4.1. DESCRIPTION OF THE REPRESENTATIVE SITE

As a representative example, it was assumed that the integrated plant would be located on the Arabian Peninsula, with site conditions typical for operating and planned desalination plants in this area.

For the purpose of cost comparison, the operation reference date was assumed to be **January 1, 2005**. However, it must be borne in mind that the actual period required for the planning and implementation of a nuclear power project may be longer.

For the seawater desalination plant, a reference capacity of about **290 000 m^3/d** was chosen. This capacity corresponds to some projects planned on the Arabian Peninsula, and would not lead to a great dependence on a single desalination plant.

The demand for additional baseload power was assumed to be **450 to 550 MW(e)** by the year 2005. This amount appears reasonable, based on plans to connect the electric grids of the Gulf Co-operation Council (GCC) countries by the year 2008 [18].

The technical performance data of the seawater desalination plants are strongly influenced by the temperature and composition of seawater. For the representative site, a seawater temperature of **24 to 35°C (annual average of 28.5°C)** and a seawater TDS of about **43 300 ppm** was taken as a basis. It was assumed that the MED and MSF desalination plants would be conservatively designed for 30°C and 45 000 ppm, and the RO plant for 27°C and 45 000 ppm.

For RO plants, the potable water quality required has great influence on the plant configuration. In this study, the WHO drinking water standards have been applied, which recommend **1000 ppm for TDS** and **250 ppm for chlorides** as the "highest desirable level".

4.2. DESCRIPTION OF THE REFERENCE SEAWATER DESALINATION PLANTS

The following seawater desalination processes were considered as the most interesting for large scale seawater desalination in integrated co-production plants:

- Multi-stage flash once through distillation (MSF-OT);
- Multiple effect distillation (MED) and
- Reverse osmosis (RO).

4.2.1. MSF-OT plants

The MSF-OT units chosen are of modular design. The modules are arranged parallel to each other and connected by U-turns of the brine flow in the condensers and the evaporators. Each module contains several evaporation stages placed in long tube arrangement. The condenser tubes are arranged in 2 parallel bundles per module. Figures 13 and 14 show the module arrangement of a long tube MSF-OT unit with 44 stages, and the cross-sectional view of one of its modules respectively.

As construction materials for the MSF-OT units, stainless steel for evaporation shells and titanium (TiPa) for the tubes are used. The units are operated without separate deaerators and decarbonators.

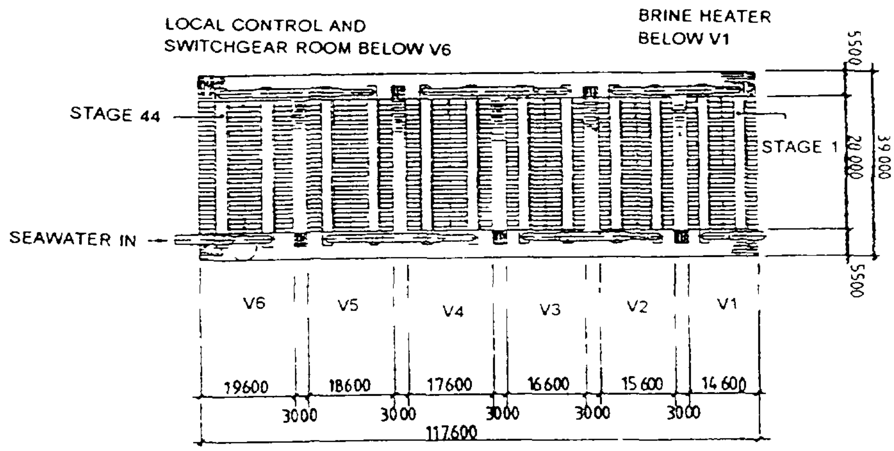


FIG 13 Module arrangement of a 44 stages MSF-OT long tube unit (72 000 m³/d) [10]

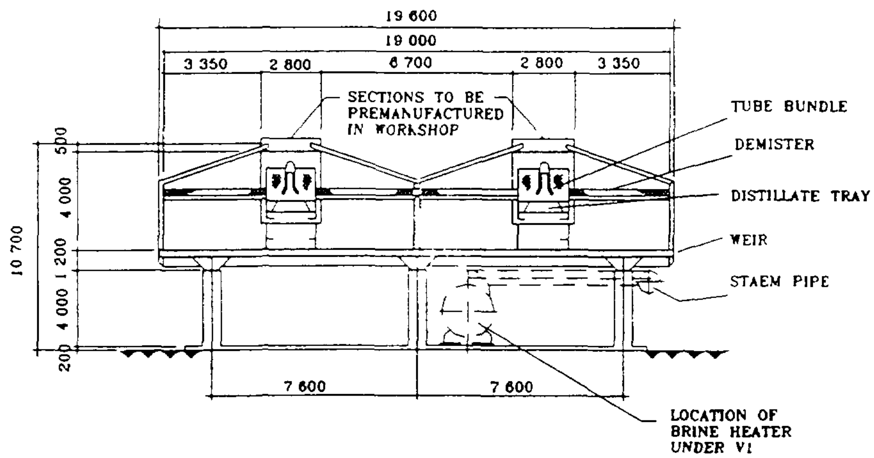


FIG 14 Cross-sectional view of a MSF-OT long tube unit [10]

TABLE XII TECHNICAL PERFORMANCE DATA OF MSF-OT UNITS ALL OF 72 000 m³/d [11]

		MSF-1	MSF-2	MSF-3	MSF-4
GOR		13 5	11 5	9 5	7 5
Number of stages		44	35	27	20
Maximum brine temperature	°C	125	110	98	90
Brine blow-down temperature	°C	35 9	35 8	36 0	36 6
Steam temperature in brine heater	°C	127 5	112 5	100 6	93 0
Thermal heat consumption [a]	kW(th) h/m³	45 1	53 7	65 6	83 5
Electricity consumption	kW(e) h/m³	2 6	2 8	3 0	3 2
Seawater flow [b]	m³/h	21 000	25 000	30 000	35 000

Seawater design parameters 30°C, 45 000 ppm

a) without steam supply for vacuum units
b) including cooling water demand of multi-stage steam ejector vacuum system of barometric type

The nominal net capacity of a single MSF-OT unit is 72 000 m^3/d, taking a seawater temperature of 30°C and a seawater TDS of 45000 ppm as a basis (conservative assumptions) Therefore, 4 MSF-OT units of 72000 m^3/d are required to produce the reference water quantity Four different long tube MSF-OT units with different GOR, all of modular long tube designs, were considered Table XII shows the most important technical performance data of the different MSF-OT units

4.2.2. MED plants

Four different low temperature horizontal tube multi-effect (LT-HTME) processes with a GOR of 7 5 to 13 5, as well as two high temperature vertical tube evaporation (HT-VTE) processes with a GOR of 17 and 21 respectively were chosen At seawater design conditions (30°C, 45 000 ppm), the nominal net capacity of each unit is 36 000 m^3/d Table XIII contains the most important technical performance data of the various MED units

Dependant on the GOR, the LT-HTME plants contain 8 to 18 evaporation effects with brine flash chambers, which are located below the evaporators, and one heat rejection condenser To remove non-condensable gases, a steam jet ejector is assembled at the coolest end of the heat rejection condenser. The feed is treated with a harmless polyphosphate additive to inhibit scaling on the heat transferring outer surface of the tube bundles Low-temperature operation enables the utilization of low cost construction materials such as aluminium tubes, plastic piping and epoxy painted steel shells

The HT-VTE plants contain 28 and 23 evaporation effects respectively, in which the brine runs down as a thin film in the condenser tubes and partly evaporates In contrast to LT-HTME plants, thin-walled titanium or high-grade steel alloys are used as tube bundle material in the top effects, because of the high temperature operation The evaporation effects and the final condenser are arranged in a vacuum-tight concrete or carbon steel shell

TABLE XIII TECHNICAL PERFORMANCE DATA OF MED UNITS ALL OF 36 000 m^3/d [11]

		HT-VTE-1	HT-VTE-2	LT-HTME-1	LT-HTME-2	LT-HTME-3	LT-HTME-4
GOR		21	17	13 5	11 5	9 5	7 5
Number of effects		28	23	18	15	12	9
Maximum brine temperature	°C	120	100	70	65	60	55
Brine blow-down temperature	°C	36 5	36 5	36 0	35 5	35 5	35 5
Steam temperature in brine heater	°C	122 5	102 5	72 5	67 5	62 5	57 5
$\Delta\vartheta$ in final condenser	°C	5	5	4 5	4	4	4
Thermal heat consumption [a)]	kW(th) h/m^3	29 0	36 7	47 9	56 5	68 7	87 5
Electricity consumption	kW(e) h/m^3	0 9	1 0	1 1	1 3	1 4	1 6
Seawater flow [b)]	m^3/h	8 500	10 000	13 000	17 000	20 000	24 000

Seawater design parameters 30°C, 45 000 ppm

a) without steam supply for vacuum units

b) including cooling water demand of multi-stage steam ejector vacuum system of barometric type

4.2.3. RO plant

To produce the reference water quantity, 12 parallel trains are used, each with a net capacity of 24 000 m^3/d As membrane module configuration, hollow fibre membranes are chosen, which are operated in single stage mode The pumping energy in the brine blow-down is partly recovered by Pelton turbines Figure 11 in Section 2 illustrates the simplified flow diagram of the RO plant

The design of the RO trains was performed for a temperature range of 23 to 35°C and a TDS of 45 000 ppm, assuming a five year operation time of the membrane modules (warranty reasons) The recovery ratio of the RO trains is kept constant at 35% by regulating the feed water pressure as a function of temperature (so-called "temperature/pressure guidelines") [19] In Annex I, the chemical analysis of the produced water is given The results show that the WHO drinking water standards can be fulfilled in the single stage module arrangement

In the economic assessment, an average annual seawater temperature of 27°C was assumed (conservative assumption), so that an average feedwater pressure of 72 bar is required to produce the reference water quantity Table XIV shows some relevant technical data of the RO trains, as well as the breakdown of their electricity consumption

No separate assessment of RO plants with spiral wound membrane modules was performed, since similar specific water production cost is to be expected The pre-heating of feedwater in the condenser of the power plant beyond 35°C, which could increase the membrane performance, was not considered This is currently not state-of-the-art, and

TABLE XIV TECHNICAL DATA OF THE 24 000 m^3/d RO TRAINS [17, 19]

Membrane configuration		hollow fibre
Net water capacity	m^3/d	24 000
Seawater design temperature	°C	27
Seawater design TDS	ppm	45 000
Feedwater pressure at seawater design conditions	bar	72
Recovery ratio	%	35
Number of membrane modules		795
Electricity consumption:		
Seawater pumps (Δp=1 7 bar, η_P=0 85, η_M=0 96)	MW	0 17
Booster pumps (Δp=3 3 bar, η_P=0 85, η_M=0 96)	MW	0 33
High pressure pumps (Δp=71 bar, η_P=0 85, η_M=0 96, η_C=0 97)	MW	7 33
Energy recovery (η_{Pel}=0 85)	MW	-3 19
Other power (0 979 kW(e) h/m^3)	MW	1 00
Total specific electricity consumption	kW(e) h/m^3	5 5*

* other plant configurations may reduce the total specific electricity consumption by about 0 5 kW(e)/m^3
Subscripts p=pump M=motor C=hydraulic coupling Pel=pelton turbine

furthermore, the proportion of membrane equipment cost in total RO plant investment cost is low (about 10 to 15%), so that appreciable benefits in specific water cost are not achieved.

4.3. REFERENCE ENERGY SOURCES

The following power plants were considered as energy source for co-production of electricity (in the 600 MW(e) range) and potable water:

- PWR power plant,
- combined cycle power plant with heat recovery steam generators.

4.3.1. Pressurized water reactor (PWR) power plant

To cover the energy demand for producing the reference quantities of potable water and electricity, a medium size pressurized water reactor with a thermal power of 1870 MW(th) was chosen. The schematic flow diagram of the reference PWR and relevant technical parameters are given in Figure 15 and Table XV respectively.

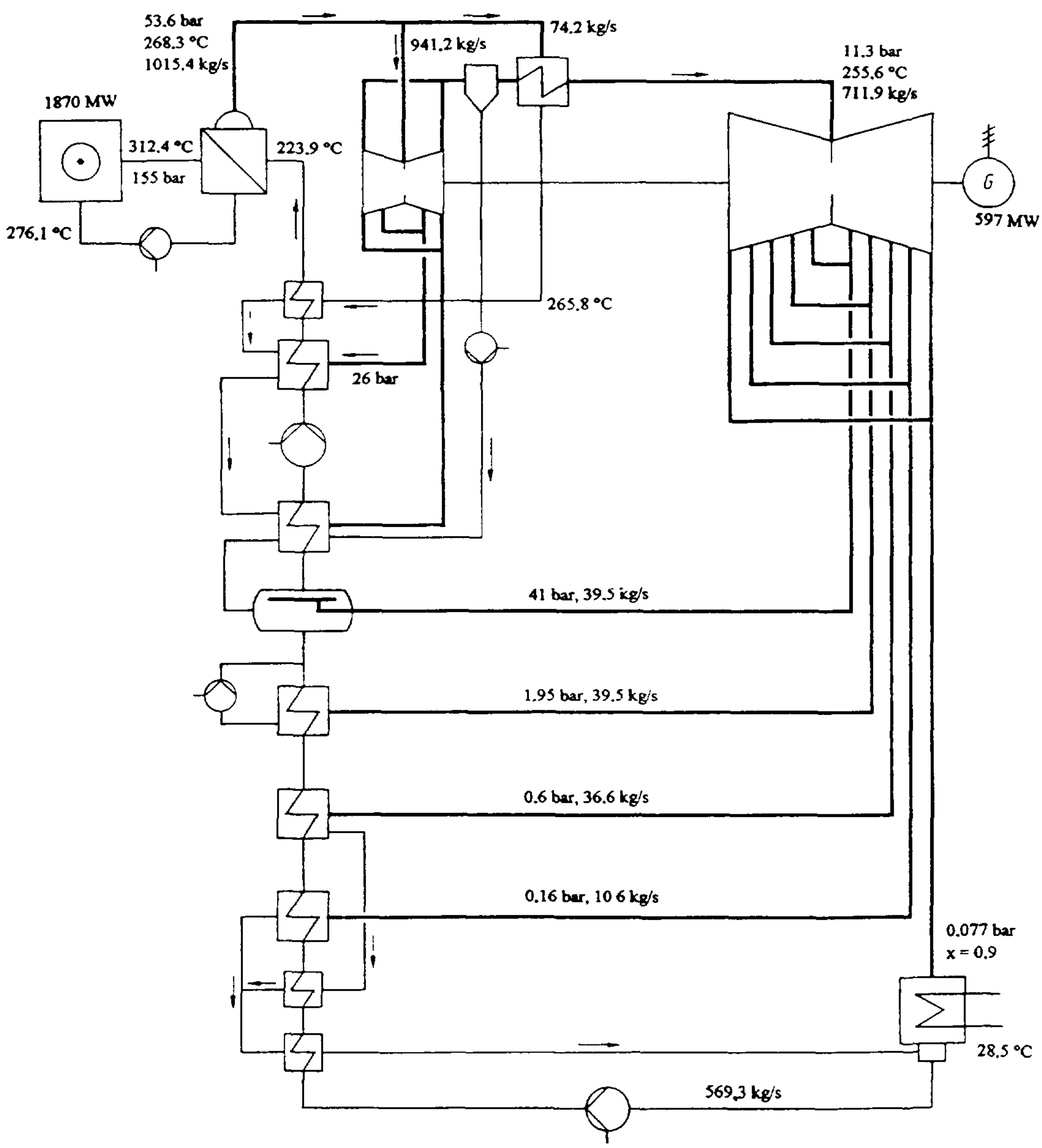


FIG. 15. Schematic flow diagram of the reference PWR power plant.

TABLE XV. TECHNICAL PARAMETERS OF THE REFERENCE PWR POWER PLANT [20, 21]

Core power	MW(th)	1870
Net output	MW(e)	about 600
Net efficiency	%	about 32.0
Auxiliary Loads	MW(e)	38
Primary system:		
Coolant/moderator		H_2O
Coolant cycle		Indirect
Pressure boundary		Pressure vessel
Pressure	bar	155
Temperature (out/in)	°C	312.4/276.1
Loops		2
Steam generators		2
Pumps		4
Fuel Reload:		
Fuel		UO_2
Initial enrichment range	%	2.0 - 3.0
Reload enrichment at the equilibrium	%	3.55
Refuelling frequency	months	18 or 24
Type of refuelling		off power
Number of fuel assemblies		145
Number of fuel rods per assembly		264
Average core power density	kW/litre	78.82
Average discharge burnup	MW·d/t	40 000
Secondary system:		
Pressure	bar	53.6
Temperature (out/in)	°C	268.3/223.9

The reactor is cooled by two 155 bar pressurized water cooling loops, where the thermal energy released during nuclear fission is transmitted to a steam power cycle in two steam generators. In the steam power cycle, there are high and low-pressure turbine stages with two moisture separator reheater units and six stages of feedwater heating. The steam generators produce steam at a pressure of 53.6 bar, yielding a net electrical output of approximately 600 MW(e) at condensing pressure of 0.077 bar (40°C). The turbine unit consists of a double flow, high-pressure turbine and two low-pressure double flow turbines that exhaust to individual condensers.

Water at 223.9°C enters the steam generators of the cooling loops. After evaporating and superheating, the steam leaves the steam generators as slightly superheated steam of 268.3°C and 53.6 bar. This steam flows into the high-pressure turbine, expanding to a pressure near 26 bar. The steam then enters the two moisture separators and flows through two reheating stages (only one moisture separator reheater is shown in Figure 15 for clarity), entering the low-pressure turbines, ultimately expanding to the condenser pressure of 0.077 bar. The condense is pumped through a series of six feedwater heaters back to the steam generators.

4.3.2. Combined cycle power plant (combined cycle)

The reference combined cycle consist of 3 natural gas-fuelled gas turbines with unfired heat recovery steam generators (HRSGs) and a dual pressure reheat steam cycle. The gas turbines are rated at 145 MW(e) net output each and the steam turbines at 205 MW(e) net output taking the average annual ambient conditions (air: 28.5°C, 1 bar, 60%, seawater: 28.5°C) as a basis. The overall net electrical output of the combined cycle is about 640 MW(e) with a net thermal efficiency of 49.7%. Table XVI contains relevant technical parameters of the reference combined cycle based on detailed calculations by a manufacturer (see Annex II) [22]. Figures 16 and 17 show the schematic flow diagram of the combined cycle and the temperature/heat recovery diagram of the HRSGs respectively. To simplify matters, only one HRSG is shown in Figure 16.

The steam parameters of the dual pressure steam cycle are 80 bar/500°C and 5 bar/151°C. Feedwater pre-heating occurs exclusively in the economizer section of the HRSGs. The steam turbine unit consists of a single flow high-pressure turbine and a double flow low-pressure condensing turbine serially placed on one shaft.

TABLE XVI. TECHNICAL PARAMETERS OF THE REFERENCE COMBINED CYCLE POWER PLANT AT AVERAGE ANNUAL AMBIENT CONDITIONS

Gas turbines:		
Net electrical output	MW(e)	3 · 145
Net thermal efficiency	%	33.8
Thermal power	MW(th)	3 · 428.9
Fuel		natural gas
Frequency	Hz	50
Compressor pressure ratio		15:1
ISO turbine inlet temperature	°C	1100
Exhaust gas flow	kg/s	3 · 494
Exhaust gas temperature	°C	541
Steam turbines:		
Net electrical output	MW(e)	204.7
Auxiliary loads	MW(e)	10.8
Steam parameter	bar/°C	80/500
Generator and mechanical efficiency	%	98.5
Condensing pressure	bar	0.077
Isentropic efficiency of high-pressure (low-pressure) turbines	%	85 (75)
Overall combined cycle:		
Gross electrical output	MW(e)	654.9
Auxiliary loads	MW(e)	15.2
Net electrical output	MW(e)	639.7
Net thermal efficiency	%	49.7

average annual air conditions: 28.5°C, 1 bar, 60% relative humidity.

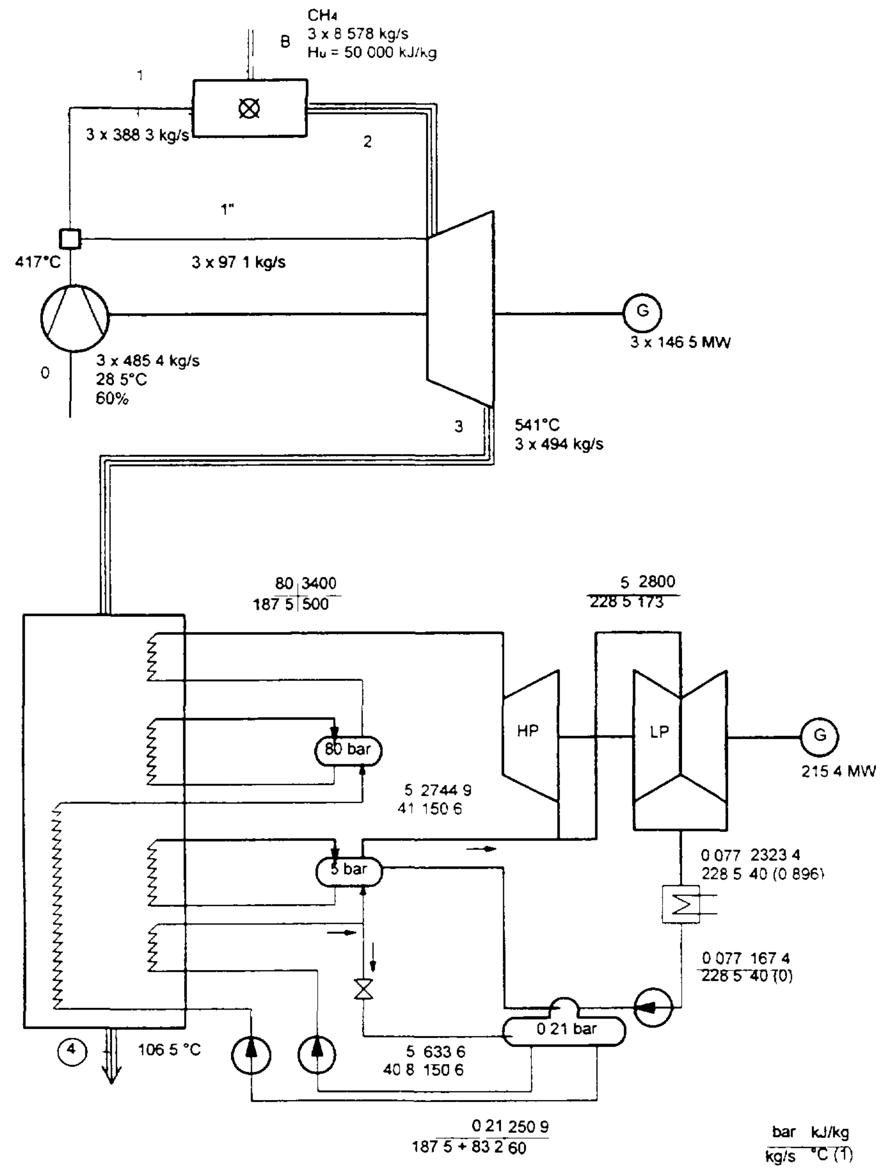


FIG. 16. Schematic flow diagram of the reference combined cycle

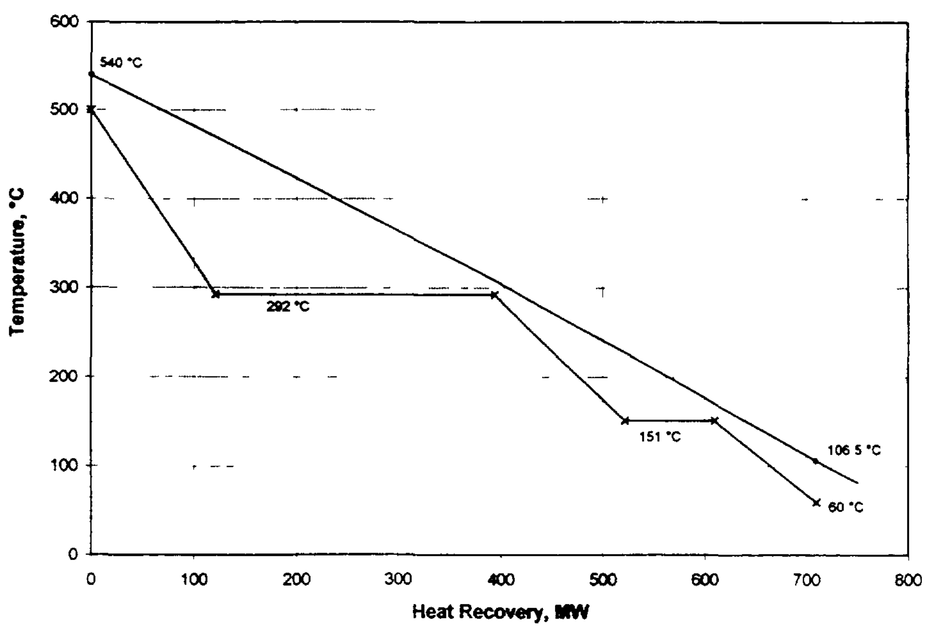


FIG 17 Temperature/heat recovery diagram of the HRSGs

4.4. COUPLING OF THE REFERENCE SEAWATER DESALINATION PLANTS WITH THE ENERGY SOURCES

As discussed in Section 3, seawater desalination plants require different forms of energy input, which are:

- electricity, for the RO plant;
- heat and some electricity, for the MSF-OT and MED plants.

4.4.1. Coupling with the PWR power plant

The coupling of the reference PWR power plant with the reference RO plant is simple, requiring only an electrical connection. Concerning technical aspects, there are no mutual influences between the PWR power plant and the RO plant, except site specific aspects, such as, water intake characteristics which have a substantial influence on site selection.

Technically, there is no need for joint siting of RO plants and PWR power plants. Electricity transport is easy and cheap, even for relatively long distances. Nuclear regulation as well as public acceptance concerns will require siting the nuclear plant at some distance to population centres. The RO plant would be as close as possible to the potable water demand (centre of population or industry), resulting in minimum water transport costs, which are about 0.25 US $\$/m^3$ for 50 km transport distance and large water flow rates ($\geq$200 000 m^3/d) [5].

Joint siting, on the other hand, offers the opportunity of sharing common facilities (see Section 3) between the PWR and the RO plant. Savings in total investment cost would be in the range of 10 to 15% for large-scale RO plants ($\geq$200 000 m^3/d). In addition, sharing of plant staff, and perhaps pre-heating of feedwater would yield further minor savings. The total benefits of joint siting would amount to 4 to 8% (0.03 to 0.08 US $\$/m^3$) in total levelized water cost [5].

The economic impact of joint or separate siting can only be analysed on a case-by-case basis due to the large influence of water transport cost. This, however, is not within the scope of the present study.

For MED and MSF processes, joint siting of the PWR and the distillation plant is necessary because transport of heat over long distances is expensive and involves substantial losses.

The turbine system in the PWR power plant has to satisfy simultaneously the requirements of electricity generation and those of providing low-temperature steam for the seawater distillation system. The latter in turn determines the specific volume of the steam, the volumetric flow rate, the average steam velocity, the cross-section areas and the steam velocity vectors of the turbine(s) supplying heat to the seawater distillation plant.

Using the reference PWR as energy source for the reference seawater distillation plants, the following solutions for providing low temperature steam could be envisaged:

1) using low temperature extraction steam from the low-pressure condensing turbines;
2) diverting steam from the crossover pipe at the inlet of the low pressure turbines;
3) using two back-pressure turbines instead of two low pressure condensing turbines;

4) replacing low-pressure turbines by extraction/condensing turbines with crossover pipes;
5) using a back-pressure turbine and a low-pressure condensing turbine in parallel (not necessarily of the same size) instead of two low-pressure condensing turbines.

Extracting steam from the lowest extraction points of low-pressure condensing turbines has a limitation. The amount of steam extractable is relatively small, not sufficient to produce the reference amount of 290 000 m^3/d.

From the exergetic point of view, diverting steam from the crossover pipe at the inlet of low-pressure turbines is not a good solution. Steam with a relatively high exergy content, synonymous with a relatively high potential to produce electricity, would be used just for low-temperature heating purposes, resulting in unnecessarily high electricity generation losses.

Solution 3 could be applied, either by operating the two low-pressure condensing turbines at higher exhaust pressure (in general limited to less than 0.2 bar), or in exchanging the low-pressure condensing turbines for back-pressure turbines. For this solution, the leeway in optimizing the seawater distillation plant is very low. Taking into account, that the GOR of distillation plants for a certain temperature difference between heating steam and cooling water sink can only be slightly varied for economic reasons, the GOR is nearly determined by the water demand required.

In solution 4, low-pressure steam, which is adjusted to the heating steam requirements of the distillation plant, is taken from the crossover pipes between the two sections of the extraction/condensing turbine. As a result, the full electrical output could come back on line if the distillation plant was shut down. Furthermore, the turbine arrangement has a high flexibility against variable water-electricity-ratios, but would lead to higher investment cost.

In the present study, solution 5 is chosen to couple the reference distillation plants with the PWR using a back-pressure turbine and a low-pressure condensing turbine in parallel (see Figure 18). The exhaust steam condition (mass flow rate, temperature, pressure) of the back-pressure turbine is also adjusted to the heating steam requirements of the distillation plant. Increasing the GOR will decrease the size of the back-pressure turbine, while the size of the low-pressure condensing turbine increases. This turbine arrangement enables the coupling of all reference distillation plants with the PWR power plant, and therefore, an economic ranking (optimization) of the distillation plants while keeping the water output constant.

The question whether solution 4 or solution 5 is better, can only be answered by a detailed and specific case study. In the present study, solution 5 was chosen because of the baseload operation of both power plant and seawater desalination plant.

When coupling seawater desalination plants with nuclear power plants, the risk of possible radioactive contamination of potable water produced must be made as low as achievable. Thus, at least two "barriers" between the reactor and the saline water are required, and the so-called "pressure-reversal" principle should be utilized. For PWR power plants, the steam generators are the first barrier against the transport of radioactive isotopes into the distillation plant.

When coupling the reference MSF plants with the reference PWR power plant, the brine heaters of the MSF units serve as the second barrier. In order to have the pressure-reversal, the brine at the brine heaters is maintained at a pressure sufficiently higher than the pressure of the heating fluid, so that the direction of a potential leakage in the brine heaters will be away **from** the MSF units, into the steam power cycle. In such a case, controlling devices that monitor the salinity of the steam power cycle of the PWR power plant would shut it down. Due to the two barriers and the pressure-reversal, the probability of radioactive contamination of the desalted water is very low. Nevertheless, should it happen, there are further instrumentation devices that monitor radioactivity in the MSF plant, and actuate systems to divert the effluents away from the mains, notify the operators and stop the process.

A more stringent provision against radioactive contamination, which helps also against salination of the steam power cycle of the PWR power station, is a "pressurized water isolation loop" between the condenser of the back-pressure turbine and the brine heaters of the MSF units (see Figure 19). The pressure in this loop would be lower than the brine pressure, but higher than that of the back-pressure steam. This results in an additional barrier and an additional pressure-reversal to prevent radioactive contamination of potable water.

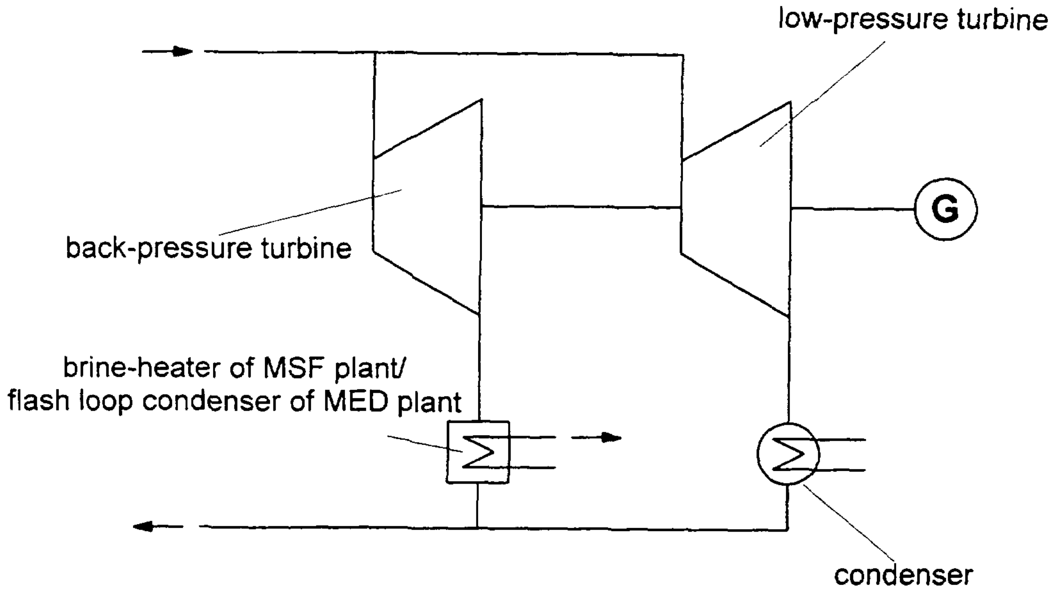


FIG. 18. Coupling of the distillation plant with the PWR power plant (back-pressure turbine and low pressure condensing turbine in parallel).

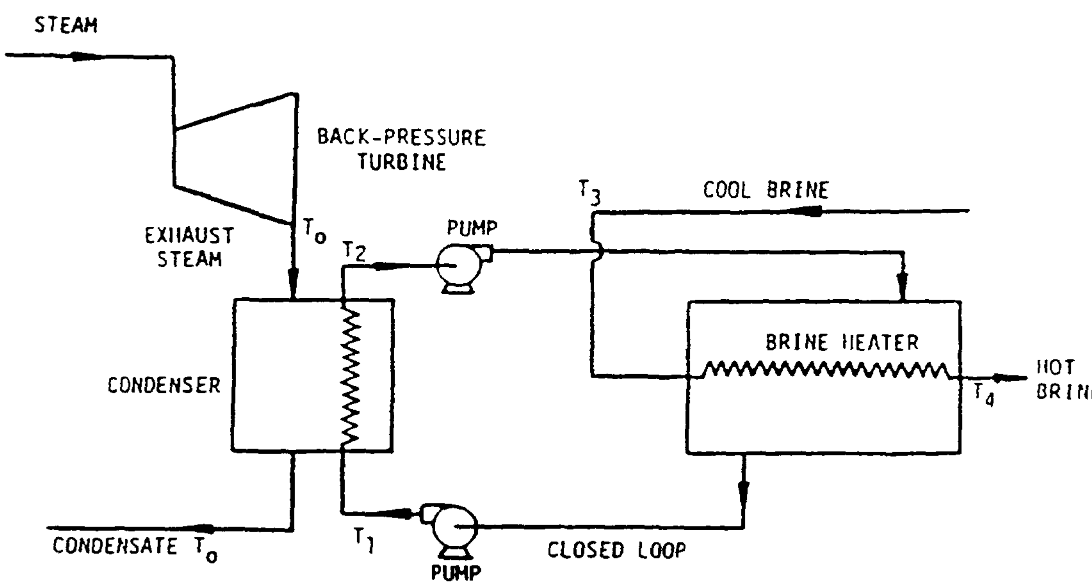


FIG. 19. Pressurized water isolation loop between back-pressure condenser and MSF plant (just one MSF brine heater is shown for clarity) [9].

In the present study, the above described kind of pressurized water isolation loop is chosen to couple the reference MSF units with the steam power cycle of the PWR power plant. However, this results in an additional investment cost for the MSF plant, higher energy demand for pumping, and in an additional loss in electricity generation because of the higher exhaust temperature of the back-pressure turbine. Furthermore, provisions are included for direct seawater (cooling water) supply to and discharge from the back-pressure condenser to allow operation of the PWR power plant when MSF units are out of service.

For MED plants, the thermal coupling with the PWR power station is implemented by open "flash-loops" (see Figure 20). Back-pressure turbine exhaust steam is condensed in the flash-loop condensers of the MED units. The latent heat of condensation is transferred to a circulating saline water stream which is heated by approximately 5°C. A portion of it flashes in the flash chambers forming low-temperature steam for the first MED effect. The condensate of the steam delivered to the first MED effect is already pure distillate and adds to the produced water. As a result, the first effect of the reference MED units listed in Table XIII can be left out, while yielding the same water output. Cooled saline water from the flash chambers is cycled to the flash-loop condensers. A portion of the circulated water is continuously drawn off as brine blow down to prevent salinity build-up. Makeup saline water is supplied from the feed stream to the circulating water to replace the losses through flashing and brine blowdown. When MED units are not in operation, the flash-loop condensers are supplied with cooling water through a bypass line.

In Annex III, the exergy analyses of the PWR power plant coupled with the reference distillation plants are given in the form of exergy flow diagrams. The results are needed to calculate the exergy flows $\dot{E}_E$ and $\dot{E}_S$ which are required to produce the total amount of electricity and steam respectively, and are the basis of the cost allocation. To understand the way of representation of the diagrams, Figure 21 serves as illustrative example, where the exergy analysis of the PWR power plant coupled with the MSF-1 plant is described in more detail.

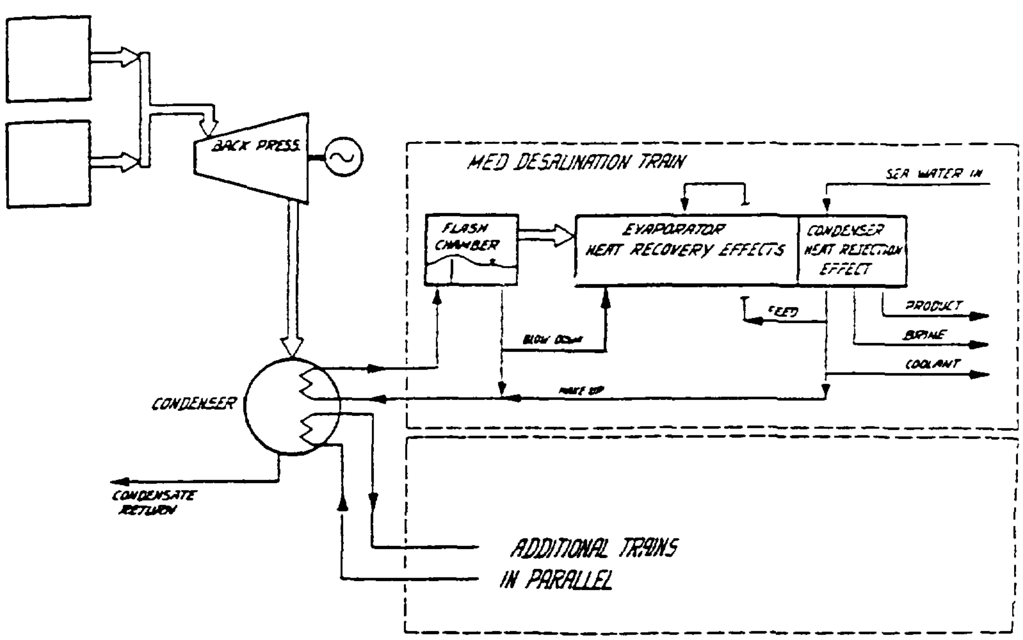


FIG. 20. Flash-loop between back-pressure condenser and MED units [23]

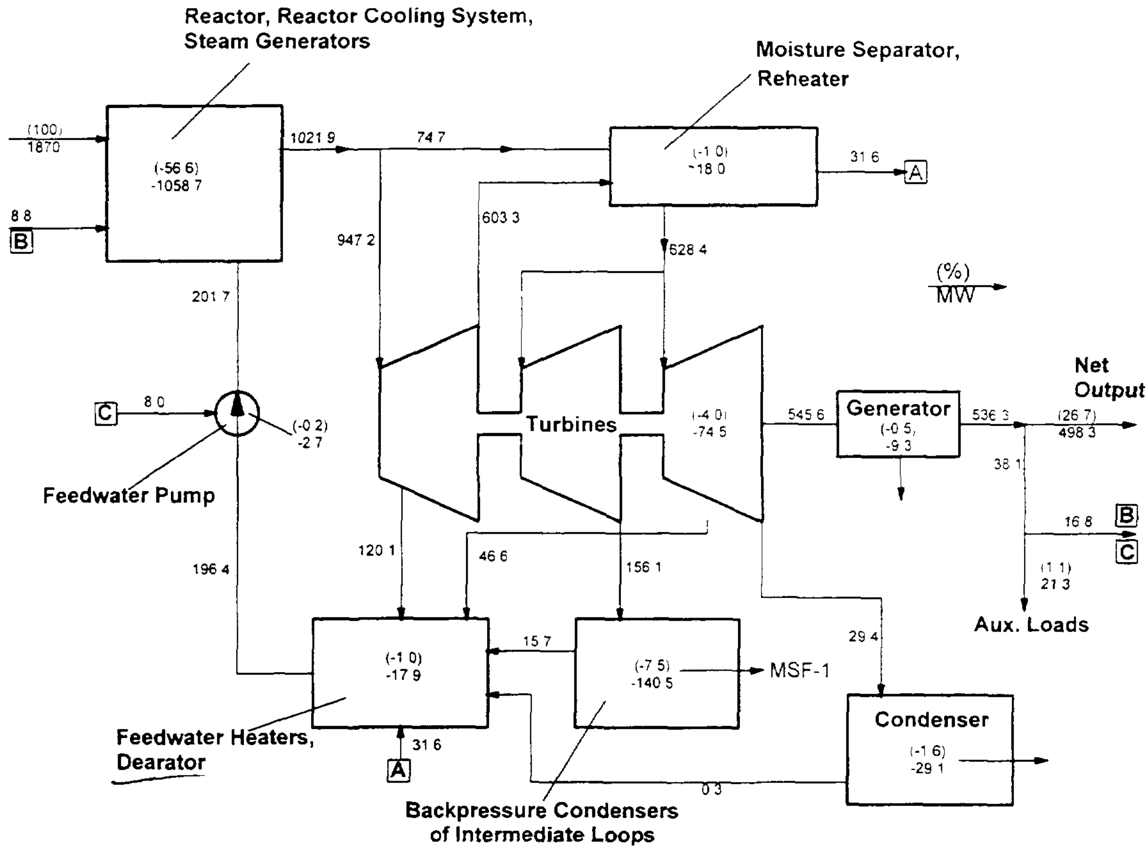


FIG. 21. Exergy flow diagram of the PWR power plant coupled with the MSF-1 plant.

The individual plant components of the power plant are divided into blocks. The numbers shown alongside the flow streamlines between, into and out of blocks represent the amount of exergy in MW flowing past the block. The numbers inside the blocks represent either the exergy losses occurring in this block calculated by Equation (6), or the exergy flows consumed in this block. Numbers shown within parentheses are values of exergy expressed as a percentage of the exergy of fuel supplied to the reactor (1870 MW). The exergy flows of the individual water/steam flows were calculated by Equation (7), using the annual average seawater temperature (28.5°C) as temperature of the reference surroundings T_0 (see Section 3.4.1.). The exergy of the nuclear fuel was equated with the thermal power of the reactor. In this example, 498.3 MW of exergy leave the system as electrical net output, 140.5 MW are needed to supply the MSF-1 plant with heating steam, the balance of the 1870 MW of exergy supplied to the system are destructed somewhere in the system because of irreversibilities.

Table XVII summarises the results of the exergy analyses for each distillation plant coupled to the PWR power plant. Furthermore, the allocation of the exergy of fuel to $\dot{E}_I$ and $\dot{E}_S$ according to Equations (14) and (15) respectively are given.

4.4.2. Coupling with the combined cycle power plant

The coupling of the reference seawater desalination plants is done in the same way as for the PWR power plant. The only difference is that the isolation loops for coupling of distillation plants are left out. That is, the exhaust steam of the back-pressure turbine is directly fed to distillation plants.

TABLE XVII EXERGY ANALYSES OF THE PWR POWER PLANT COUPLED WITH THE REFERENCE DISTILLATION PLANTS

		PWR coupled with										
		-	HT-VTE-1	HT-VTE-2	LT-HTME-1	LT-HTME-2	LT-HTME-3	LT-HTME-4	MSF-1	MSF-2	MSF-3	MSF-4
Net electrical output P_{net}	MW(e)	597	533	535	548	547	546	540	498	497	486	469
Exhaust steam temperature of back-pressure turbine	°C	-	129	109	79	74	69	64	133	118	106	98
Thermal heat to distillation plant	MW(th)	-	348	441	574	678	825	1050	541	644	787	1001
Proportional break down of exergy flows according Table VIII:												
Losses in primary circuit including reactor and steam generator	%	56.6	56.6	56.6	56.6	56.6	56.6	56.6	56.6	56.6	56.6	56.6
Net electrical output	%	31.9	28.5	28.6	29.3	29.3	29.2	28.9	26.7	26.6	26.0	25.1
Steam provided for distillation plant	%	-	4.7	5.0	4.4	4.8	5.2	5.9	7.5	7.9	8.6	10.0
Losses in turbines	%	5.1	4.3	4.4	4.5	4.5	4.5	4.4	4.0	4.0	4.0	3.9
Electrical auxiliary loads (with exception of feedwater pumps and reactor coolant pumps)	%	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1
Losses in condensers	%	2.8	1.9	1.7	1.4	1.1	0.9	0.5	1.6	1.4	1.1	0.7
Losses in moisture separators and steam reheaters	%	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Loss in generator and mechanical losses	%	0.6	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5
Losses in feedwater heaters	%	1.2	1.1	1.0	1.0	0.9	0.9	0.9	1.0	0.9	0.9	0.8
Losses in feedwater pumps	%	0.2	0.2	0.2	0.2	0.2	0.2	0.2	0.2	0.2	0.2	0.2
Exergy of fuel E_f	MW	1870	1870	1870	1870	1870	1870	1870	1870	1870	1870	1870
Exergy flows to produce electricity E_f according Eq. (14)	MW	1870	1649	1638	1664	1649	1629	1595	1522	1505	1470	1404
Exergy flows to provide steam for distillation plants E_S according Eq. (15)	MW	-	221	232	206	221	241	275	348	365	400	466

TABLE XVIII EXERGY ANALYSES OF THE COMBINED CYCLE COUPLED WITH THE HT-VTE DISTILLATION PLANTS

		Combined cycle coupled with		
		-	HT-VTE-1	HT-VTE-2
Net electrical output P_{net}	MW(e)	640	585	586
Exhaust steam temperature of back-pressure turbine	°C	-	123	103
Thermal heat to distillation plant	MW(th)	-	348	441
Proportional break down of exergy flows according to Tables X and XI:				
Losses in the combustion chamber	%	26.4	26.4	26.4
Exhaust gas leaving the HRSGs	%	3.4	3.7	4.4
Electrical auxiliary loads of gas turbine cycles	%	0.3	0.3	0.3
Losses in gas turbines (including cooling losses)	%	8.4	8.4	8.4
Losses in compressors	%	2.6	2.6	2.6
Loss in generator and mechanical losses of gas turbines	%	0.7	0.7	0.7
Net electrical output of gas turbines	%	32.7	32.7	32.7
Losses in HRSGs	%	3.7	4.1	3.3
Other losses in steam cycle	%	0.2	0.2	0.1
Electrical auxiliary loads of steam cycle	%	0.8	0.8	0.8
Losses in steam turbines	%	3.7	2.0	1.9
Losses in condenser	%	1.4	0.5	0.2
Loss in generator and mechanical losses in steam turbines	%	0.3	0.2	0.2
Net electrical output of steam turbines	%	15.4	11.2	11.3
Steam to distillation plant	%	-	6.2	6.6
Exergy of fuel	MW	1331.4	1331.4	1331.4
Exergy flows to produce total electricity E_l according to Eq. (22)	MW	1331.4	1184.1	1178.2
Exergy flows to produce steam E_s according to Eq. (21)	MW	-	147.3	153.2

Only the RO plant and the two HT-VTE plants were coupled with the combined cycle. The coupling with the other distillation plants would require serious modifications of the steam power cycle, since there is not sufficient low-pressure exhaust steam at the outlet of the back-pressure turbine available to cover the heat consumption of these distillation plants. This would result in a substantial reduction in thermal efficiency of the combined cycle and also to higher water production cost.

In Annex IV, the exergy analysis of the combined cycle power plant coupled with both the HT-VTE plants are given. Table XVIII summarises the results of the exergy analyses as well as the allocation of the exergy of fuel to the exergy flows E_l and E_s.

To calculate the individual exergy flows of the gas turbine cycle, the following assumptions were made:

- condition of reference surroundings: 28.5°C, 1 bar;
- exergy content of moist air at reference surroundings is negligible;
- methane (CH_4) is considered as reference fuel for the gas turbines;
- exergy and heating value of CH_4 are approximately equated with 51738 kJ/kg and 50 000 kJ/kg respectively (according to *Baehr* [24]);
- polytropic efficiency of compressor: 91%;
- 20% of the compressed air is needed for cooling the gas turbine rotors, blade carriers and gas turbine stages;
- auxiliary loads of the gas turbines: about 1%;
- generator and mechanical efficiency: 98.5%.

The exergy content of the individual matter flows was calculated by means of tabulated data of the heat capacity and entropy of ideal gases assuming the matter flows to be ideal gas compounds. Table XIX shows the composition of the individual matter flows, their thermodynamic parameters as well as their exergy flows. The numbering of each matter flow corresponds to the numbers in Figure 16.

The flue gas temperature at the outlet of the combustion chamber (No. 2 in Figure 16 and Table XIX) was determined by an energy balance on the combustion chamber.

TABLE XIX. THERMODYNAMIC PARAMETERS OF THE INDIVIDUAL MATTER FLOWS IN THE GAS TURBINE CYCLE

No		0*	1`	1``	B	2	3	4
	matter flow	air	air	air	CH_4	flue gas	flue gas	flue gas
T	°C	28.5	417	417	25	1261	541	106.5
p	bar	1	15	15	1	15	1	1
$\dot{m}$	kg/s	1456.4	1165.0	291.2	25.7	1190.8	1482	1482
h	kJ/kg	29	435.4	435.4	**	1508.1	593.2	112.0
s	kJ/kg/K	6.930	7.007	7.007	**	7.985	7.942	7.102
ζ_{CH_4}		-	-	-	1	-	-	-
ζ_{Air}		0.985	0.985	0.985	-	-	-	-
ζ_{O_2}		-	-	-	-	0.059	0.048	0.048
ζ_{H_2O}		0.015	0.015	0.015	-	0.067	0.053	0.053
ζ_{CO_2}		-	-	-	-	0.138	0.155	0.155
ζ_{N_2}		-	-	-	-	0.737	0.744	0.744
$\dot{E}$	MW	0	446.2	111.6	1331.4	1426.2	383.5	45.8

N_2* nitrogen in air including trace gases such as argon, neon and CO_2.
* was approximately defined as reference surroundings.
** not relevant.

The air temperature at the outlet of the compressor was calculated by the implicit polytropic equation of an ideal gas (see Equation (23)):

$$s_1^0 = s_0^0 + \frac{R_A}{\eta_{v,C}} \cdot \ln \pi \ , \tag{23}$$

with

s_1^0: entropy of air at T_1 and 1 bar (kJ/kg/K),

s_0^0: entropy of air at surroundings (kJ/kg/K),

R_A: gas constant of moist air (0.2896 kJ/kgK),

$\eta_{v,C}$: polytropic efficiency of compressor,

π: compressor pressure ratio.

Finally, the calculation of the exergy loss flows in the components of the gas turbine cycle was performed by the exergy balance Equation (6).

The exergy analysis of the steam cycle was done in the same way as for the steam cycle of the PWR power plant.

4.5. COST COMPARISON

The economic methodology for assessing the integrated plant configurations is based on computing the life-time levelized equivalent electricity generation cost, the life-time levelized potable water production cost and the life-time levelized electricity generation cost for each plant configuration (see Section 4.5.1). To finance the integrated plant, hundred per cent outside financing was assumed to obtain a general comparison of the plant investment, independent from the capital resources of the owner.

4.5.1. Levelized production cost

The levelized production cost of any product is obtained by determining the present value of all the year-by-year expenditures related to its production and dividing that amount by the present value of the product generated over the life of the plant. The term "present value" is the equivalent of all the expenditures/products, transacted/generated over the time frame from start of construction to end of decommissioning, discounted to a reference date by using a predetermined interest or discount rate. The reference date can be any date in time. Usually the date of implementing the economic assessment of the investment, or the date of commissioning is taken as reference date. In the following, the method to calculate the levelized production cost is briefly described.

To determine the present value of expenditures related to the production of any product i, all year-by-year expenditures are discounted to the reference date $\boldsymbol{T_0}$ using an appropriate discount rate $\boldsymbol{r}$ and added up. It is assumed that all expenditures occur at the end of each year, and that the expenditures escalate at an annual escalation rate $\boldsymbol{e}$.

$$PV_{I_i} = \sum_{t=T_{CS}}^{T_L} \frac{C_{i,0}(t) \cdot (1+e)^{t-T_0}}{(1+r)^{t-T_0}} \ , \tag{24}$$

where

PV_i is the present value of expenditures of the product i,

$C_{i,0}(t)$ are the expenditures in the year t to produce i in the value of currency of the year T_0 (payable at the end of the year, not considering the price escalation),

T_{CS} is the starting date of construction, and

T_E is the end of decommissioning date.

The calculation can in principle be performed either in current money terms, with nominal cost escalation and a nominal discount rate, or in constant money terms, with cost escalation relative to general inflation ("real" escalation) and a "real" discount rate. Expenditures that are uniformly distributed over the year can approximately be transformed to expenditures payable at the end of the year by multiplying by the term $(1+e)^{0.5}/(1+r)^{0.5}$.

The annual expenditures can further be split into different categories *j*. Assuming that the expenditures are uniformly distributed over each year, Equation (24) can be converted to Equation (25):

$$PV_i = \sum_j \sum_{t=T_{CS}}^{T_E} \frac{C_{i,0_j}(t)\cdot\left(1+e_j\right)^{t-T_0+0.5}}{\left(1+r\right)^{t-T_0+0.5}}. \qquad (25)$$

The present value of the product i is defined in a similar way:

$$PV_P = \sum_{t=T_C}^{T_l} \frac{P_i(t)}{\left(1+r\right)^{t-T_0+0.5}}, \qquad (26)$$

where

PV_P is the present value of product i,

$P_i(t)$ is the production of i in year t, assumed to be uniformly distributed over the year,

T_C is the date of commissioning, and

T_l is the date of operation end.

The levelized production cost to produce i, for example expressed in $\$/m^3$ or $/kW(e)·h, can be calculated by dividing Equation (25) by Equation (26):

$$c_i = \frac{\displaystyle\sum_j \sum_{t=T_{CS}}^{T_E} \frac{C_{i,0_j}(t)\cdot\left(1+e_j\right)^{t-T_0+0.5}}{\left(1+r\right)^{t-T_0+0.5}}}{\displaystyle\sum_{t=T_C}^{T_l} \frac{P_i(t)}{\left(1+r\right)^{t-T_0+0.5}}}. \qquad (27)$$

Assuming that the annual production of i is constant over the life of the plant, the levelized production cost c_i is equal to the levelized annual expenditures (annuity of expenditures) divided by the amount of annual production. The annuity of expenditures to produce i is the annual cost that, when applied as a uniform series over the life of the plant, results in the same present value of expenditures as the actual lifetime present value of expenditures PV_j calculated by Equation (25), while keeping T_0 as reference date.

For integrated plants for electricity and potable water, the annuity of expenditures $\mathbf{C_{0*}}$ include:

- capital costs $\mathbf{C_{Ca}}$,
- fuel costs $\mathbf{C_F}$ (including the cost of intermediate storage as well as conditioning and final waste disposal for nuclear fuel),
- O&M costs $\mathbf{C_{O\&M}}$,
- decommissioning cost $\mathbf{C_{DC}}$ (only considered for PWR power plant), and
- fuel cost of the backup heat source for the distillation plant $\mathbf{C_{FB}}$ (not applicable for RO plants).

$$C_{0*} = C_{Ca} + C_F + C_{O\&M} + C_{DC} + C_{FB}, \qquad \text{in \$/a} \tag{28}$$

In the following, the equations to calculate the individual annuities of expenditures, which were applied in the economic assessment, are given. A real discount rate as well as real escalation rates were assumed. To simplify matters, the annual expenditures were assumed to occur at the end of the year.

The annuity of capital costs $\mathbf{C_{Ca}}$ was determined from the total investment cost $\boldsymbol{C_{TO}}$ by means of a fixed charge rate $\boldsymbol{a_n}$. The annuity of capital cost is termed fixed capital charge since these yearly expenditures need to be paid regardless of the amount of products generated. The fixed charge rate is a factor which is used to multiply the total investment cost of the plant C_{TO} to obtain C_{Ca}. In the following, the equations to calculate the annuity of capital costs are given. All monetary amounts are referred to the reference date T_0.

$$C_{Ca} = C_{TO} \cdot a_n, \tag{29}$$

$$a_n = \frac{r \cdot (r+1)^n}{(1+r)^n - 1}, \tag{30}$$

$$C_{TO} = (C_{VO} + C_O) \cdot (1 + IDC), \tag{31}$$

$$IDC = (1 + i_{CS})^{\frac{t_{CS}}{2}} - 1, \tag{32}$$

where

C_{Ca}	is the annuity of capital cost in \$/a,
C_{TO}	is the total investment cost of the integrated plant in \$,
a_n	is the fixed charge rate in 1/a,
r	is the real discount rate in 1/a,
n	is the plant economic life in years,
C_{VO}	is the vendor overnight cost in \$,
C_O	is the owner's cost in \$,
IDC	is the factor for interest during construction,
i_{CS}	is the real interest rate during construction in 1/a,

t_{cs} is the construction lead time of the plant in years.

Equation (32) is an approximation; it is valid for costs uniformly distributed over the time of construction.

The nuclear fuel cost and the O&M costs were assumed not to have any costs which vary with time other than by the rate of inflation (real escalation rate is 0). Accordingly, the annuity of nuclear fuel cost $\mathbf{C_F}$ and the annuity of O&M costs $\mathbf{C_{O\&M}}$ are equal to the annual expenditures in the value of currency of the reference date T_0.

The annuity of decommissioning cost of the PWR power plant $\mathbf{C_{DC}}$ is the annual funds that have to be put aside to finance the decommissioning of the PWR power plant. In the present assessment, data from a study about the feasibility of nuclear power plants in Arabian countries are used to determine C_{DC} [25].

For firing the combined cycle power plant and the backup heat boilers, gas was assumed as fuel, which was considered to be subject to a real price escalation. Calculating the annuity of the fossil fuel cost and the fuel cost of the backup heat source $\mathbf{C_F}$ and $\mathbf{C_{FB}}$ respectively, the year-by-year expenditures on fuel cost are discounted to the reference date T_0 and levelized over the economic lifetime of the plant $\boldsymbol{n}$, considering a real fuel escalation rate $\boldsymbol{e_{FF}}$ (see Equation (33) [16]).

$$C_a = C_{a_0} \cdot \frac{\left(1+e_{FF}\right)^{t_c - t_0}}{\dfrac{\left(1+r\right)}{\left(1+e_{FF}\right)} - 1} \cdot \left(1 - \left(\frac{1+e_{FF}}{1+r}\right)^{n}\right) \cdot a_n , \tag{33}$$

where

C_{a_0} are the expenditures on fuel of combined cycle or the backup heat source in the value of currency at the reference date T_0.

4.5.2. Combining cost allocation methods with the levelized production cost

The methodology of calculating the equivalent electricity generation cost and the exergetic cost allocation method presented in Section 3.2 and 3.4 respectively were combined with the method to calculate the levelized production cost in order to calculate the equivalent electricity generation cost, the levelized potable water production cost and the levelized electricity generation cost. The annual production of potable water and electricity was assumed to be constant over the economic life of the integrated plant, so that the overall expenditures of the plant C_0 in Equation (8) corresponds to the annuity of their expenditures C_{0*} of the plant in Equation (28). That is, the four portions of C_0 (C_L, C_S, C_C, C_{W*}) can be determined by the cost components listed in Tables VII and IX, and the equations to calculate the various annuities of the expenditures of the integrated plant specified in Section 4.5.1. As a result, the equivalent electricity generation cost c_{eq} (cent/kW(e)·h), the potable water production cost c_W ($/m^3) and the electricity generation cost c_E (cent/kW(e)·h), can be calculated as follows:

$$c_W = \frac{Annual\ Water\ Production\ Expenditures,\ C_W}{Annual\ Amount\ of\ Water\ Produced,\ W}, \tag{34}$$

$$c_E = \frac{Annual\ Electricity\ Generation\ Expenditures,\ C_E}{Annual\ Amount\ of\ Electricity\ Generated,\ E_a}. \tag{35}$$

$$c_{eq} = \frac{Overall\ Annual\ Expenditures\ of\ the\ Integrated\ Plant,\ C_0}{Annual\ Amount\ of\ Electricity\ Generated,\ E_a}. \tag{36}$$

4.5.3. Economic reference assumptions

In the following, the most important economic assumptions made to calculate the specific levelized production/generation costs of the integrated plant alternatives are explained. A complete listing is included in the EXCEL spreadsheets in Annexes V and VI.

The reference currency is the United States dollar (**US $**) **of January 1995**. For the purpose of economic assessment, the commissioning reference date was assumed to be **January 1, of the year 2005** (see explanations in Section 4.1).

Real discount/interest rates used in many industrialized and developing countries range from 5% to 10% according to IAEA and OECD/NEA studies [26, 27]. In the present economic assessment, 8% was considered as reference value for Arabian countries, and 5% and 10% were used for sensitivity analyses.

Table XX shows the cost data assumed for the reference power plants. The total overnight cost assumed for the PWR power plant is based on information provided to the Agency by a prospective supplier valid for conditions prevailing in industrialized countries [21]. A 10% increase was assumed to consider additional costs resulting from construction in Arabian countries [25]. These additional costs correspond to the net result of taking into account the higher costs resulting from construction outside the suppliers' country and the effects of local participation. The total overnight cost assumed for the combined cycle power plant is based on levelized experience data of similar projects in developing countries provided by manufactures, consultants and feasibility studies [28, 29]. Separate determination of the percentage capital cost of incremental equipment for providing steam to the distillation plants was not considered. This cost portion is included in the percentage capital cost of the remaining power station, and is allocated to the steam production proportional to the exergy flow $\dot{E}_s$ in Equation (10).

The fuel cost and O&M cost of the PWR power plant are based on data generated for a case study on the feasibility of small and medium power plants in Egypt [25]. It was assumed that these cost data are also valid for countries on the Arabian Peninsula. The cost of nuclear fuel has been declining in the last decade and has now been stabilized, so that no real cost escalation of nuclear fuel is currently foreseen for the next decade [27, 29]. For the combined cycle plant, it was assumed that the fuel cost will be governed by the world-market crude oil price by the equivalent in heating value. For the representative assessment, the crude oil price was set to US $17 per barrel, with 2% annual real escalation rate according to OECD/NEA [29]. 0 and 4% annual real escalation rate were assumed for sensitivity analysis.

For both the power plants, an economic life of **30 years** and an average life-time load factor of **80 %** was assumed [17, 26]. In the case of combined cycle, since the technical live is expected not to be longer than 15 years, replacements of the rotors and the hot-gas-path parts of the gas turbines will be necessary to allow for continuing operation.

TABLE XX. COST ASSUMPTIONS FOR THE REFERENCE POWER PLANTS [17, 21, 25, 28, 29]

		PWR power plant	Combined cycle power plant
Net output	MW(e)	600	640
Total overnight cost [a)]	$/kW(e)	1874	600
Cost percentage of turbogenerator equipment	%	20	30 [c)]
Cost percentage of remaining power plant	%	80	70
Construction time	months	60	36
Annual fixed O&M cost	M$/a	39.42	7.41
Annual variable O&M cost	mills/kW(e)·h	0.5	3.85
Annual fuel cost	mills/kW(e)·h	7.47	b)
Annual decommissioning cost	mills/kW(e)·h	1.00	N/A
Life-time average load factor	%	80	80

a) including owner's cost and contingency, excluding cost escalation and interest during construction
b) is calculated separately
c) including cost of air compressors, gas turbines and generators of gas turbine cycles

Gas and/or fuel-oil fired backup heat boilers were considered to supply heat to the distillation units if the power plant is out of operation. The capital cost of these boilers was estimated at 50 000 US $ per MW(th) installed [17]. Furthermore, it was assumed that the fuel cost of the backup heat boilers will also be governed by the world-market crude oil price.

Table XXI shows the unit base cost of the reference seawater desalination units. The capital cost of the MSF and MED units are based on cost estimates by consulting engineers for construction in Arabian countries [11]. The cost data of the MSF units are related to an advanced MSF-OT design (see Section 4.2.1), which have a noticeably lower unit base cost than current MSF-BR units of cross-tube design. The unit base cost of the RO unit is drawn from actual experience in Arabian countries, taking into account the current low cost of membrane equipment [17].

It was assumed that multiple unit construction would reduce the unit base costs of seawater desalination plants in accordance with cost reduction factors contained in Table XXII. These factors consider shared engineering, erection, supervision, infrastructure, as well as better purchasing conditions, learning curve effect during manufacturing and reduced management and insurance costs.

To the adjusted base cost, the following cost items have to be added:

- Intermediate loop cost C_{il} (not applicable for the RO plant), approximately calculated by Equation (37) for pressurized water isolation loop and open flash loop:

$$C_{il} = (11 / GOR)^{0.6} \cdot 100 \quad \text{in US \$ per m}^3\text{/d desalination plant capacity.} \qquad (37)$$

- Interest during construction (see spreadsheets in Annex V and VI).

- Owner's cost and contingency, which were estimated as 5% and 10% respectively.

TABLE XXI. UNIT BASE COST OF THE REFERENCE SEAWATER DESALINATION UNITS [11, 17]

Unit design	Unit size m^3/d	Unit base capital cost [a)] M$
MSF-1	72 000	105.0
MSF-2	72 000	100.0
MSF-3	72 000	95.5
MSF-4	72 000	92.0
HT-VTE-1	36 000	45.3/46.5[b)]
HT-VTE-2	36 000	42.3/43.6[b)]
LT-HTME-1	36 000	39.8/41.4[b)]
LT-HTME-2	36 000	36.4/38.2[b)]
LT-HTME-3	36 000	34.6/36.7[b)]
LT-HTME-4	36 000	32.6/35.3[b)]
RO	24 000	24.0

a) Unit base cost excludes:
- intermediate loop cost (not applicable for RO unit),
- water intake/outfall structures,
- owner's cost (5%),
- contingency (10%),
- interest during construction

b) Unit base cost for coupling with PWR power plant (one evaporation effect is substituted by the intermediate flash-loop)

TABLE XXII. COST REDUCTION FACTORS FOR SEAWATER DESALINATION PLANTS WITH MULTIPLE UNITS [30]

Number of units:	1	2	3	4	5	6	8	10	12
Cost reduction factor:	1.000	0.933	0.896	0.871	0.851	0.836	0.812	0.794	0.780

– Water intake/outfall structures.

Calculation of the capital cost of the seawater intake/outfall structures C_{SIA} was performed by Equation (38), which is based on extrapolated experience data of open seawater intake/outfall structures as a function of the seawater mass flow rate $\dot{m}_{SIA}$ required by the integrated plant [17]:

$$C_{SLA} = 7.4002 \cdot \left(\frac{\dot{m}_{SLA}}{486} \right)^{0.38} \quad \text{in US M\$.} \tag{38}$$

The cost allocation of the seawater intake/outfall structures to the power plant and the seawater desalination plant was done proportional to the seawater mass flow rates, needed as cooling water in the condenser of the power plant, and to provide cooling and/or feed water to the desalination plant respectively. The total overnight cost of the power plants listed in Table XX includes the seawater intake/outfall structure cost of the original base load power stations (only electricity production), so that this cost, calculated by Equation (38), has to be deducted from the total overnight cost (see EXCEL spreadsheets in Annex V and VI).

In the present study, an economic life of the reference seawater desalination plants of **30 years** was considered. For MSF plants, this assumption is justified by many years of experience as well as the use of corrosion-resistant materials. For large-scale MED plants, where insufficient experience exists at this time, an economic life of 30 years is also justifiable, since materials for vessels and piping used are similar to MSF plants, and since the LT-HTME plants must anticipate retubing (aluminium tubes) after 15 years operation (included in the O&M costs). Titanium or copper-nickel tubes may last 30 years. In the RO plant, the shortest life components are the membrane elements and filters which are changed at regular intervals (included in the O&M costs). Pumps, pressure vessels and piping have a minimum 30 year life.

The construction time of the reference seawater desalination plants was assumed to be **24 months**. Table XXIII shows the assumed O&M costs for the reference seawater desalination plants based on pertaining to past experience provided by consultants.

An average life-time load factor of 91% was considered for all seawater desalination plants, without taking into account the availability of the steam and electricity supply to the desalination plants. The availability of electricity supply was assumed to be about 100%, because of the integration of the plant into the electric grid. Therefore, the total load factor of the reference RO plant is also 91%. To simplify matters, it was assumed that the electricity purchased from the grid during the time when the power plant is not in operation, is equal in price with the electricity sold to the grid. This assumption is in general not the case, and has to be examined case by case. In addition, availability of steam supply has to be considered for seawater distillation plants, depending on load factors of the power plant and backup heat boilers. Figure 22 shows the procedure to calculate the combined heat source/distillation plant

TABLE XXIII. O&M COSTS FOR THE REFERENCE SEAWATER DESALINATION PLANTS (288 000 m^3/d)

		MED	MSF	RO
Management	persons	5	5	5
Average management salary	\$/a	66 000	66 000	66 000
Labour	persons	35	35	35
Average labour salary	\$/a	29 700	29 700	29 700
Spare parts	\$/m^3	0.03 [a)]	0.035	0.04
Chemicals for pre-treatment	\$/m^3	0.02	0.02	0.03
Chemicals for post-treatment	\$/m^3	0.04	0.04	0.03
Membrane price	\$/permeator	N/A.	N/A.	4 000
Membrane replacement rate	%/a	N/A.	N/A.	15
Membrane replacement	\$/m^3	N/A.	N/A.	0.06
Insurance cost	% capital cost	0.5	0.5	0.5

a) plus 0.01 \$/m^3 for LT-HTME plants for retubing of aluminium tubes (8% discount rate)

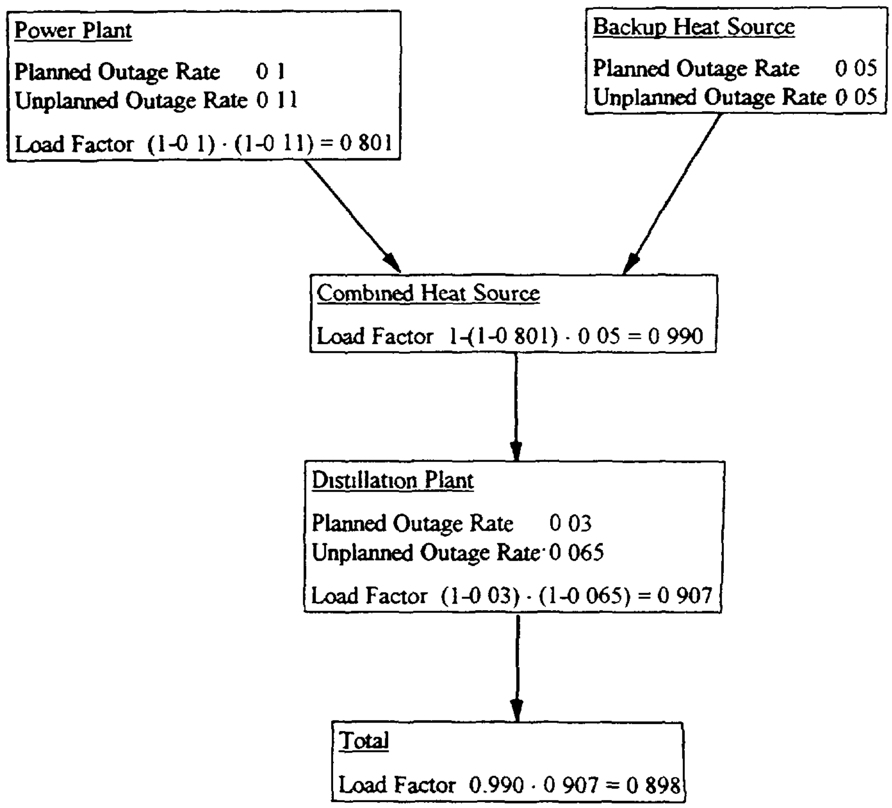


FIG 22. Calculation of the combined heat source/distillation plant load factor

load factor of the integrated plant (**89.8%**). In this calculation, it was assumed, that the planned outages (maintenance) of the backup heat boilers should not coincide with the planned outage of the power plant.

When calculating the total annual steam cost, the load factor of the power plant as heat source for the distillation plants, as well as the load factor of the backup heat source has to be determined. As a hypothetical reference case, it was assumed that the planned outages of the distillation plants coincide with the planned outages of the power plant. Accordingly, the load factor of the power plant as steam source to the distillation plants can be calculated as follows:

$0.801 \cdot (1-0.065) = \mathbf{0.749}$.

That is, the load factor of the backup heat source is **14.9%** (89.8%-74.9%).

It is to be noted, that only individual modules of seawater desalination plants are usually taken out for maintenance activities. In such a case, the load factor of the power plant as heat source for the distillation plant would be only 72.3%.

4.5.4. Results with reference assumptions

The algorithms to calculate the equivalent electricity generation cost and to apply the exergetic cost allocation method were incorporated in an EXCEL spreadsheet routine. This allows for an easy comparison of the input data, as well as the provisional and final results of the economic assessment. Annexes V and VI contain the spreadsheet outputs of the reference integrated plant alternatives analysed. A summary of the results is given in Table XXIV.

In Figures 23 to 26, the following parameters of the reference integrated plants are shown:

- the levelized equivalent electricity generation cost c_{eq},
- the levelized potable water production cost c_W,
- the levelized electricity generation cost c_E, and
- the exergy of fuel required for potable water production e_W.

The ***exergy of fuel required for potable water production*** e_W in an integrated plant, by definition, is the amount of exergy of fuel in kW·h required to produce 1 m^3 potable water. It consists of:

- the exergy of fuel allocated to steam production according to the exergetic cost allocation method (not applicable for RO),
- the exergy of fuel allocated to electricity generation which is supplied to the seawater desalination plant (also allocated according to the exergetic cost allocation method), and
- the exergy of fuel for firing the backup heat boilers (CH_4 was chosen as reference fuel).

TABLE XXIV. SUMMARY OF THE ECONOMIC ASSESSMENT OF THE REFERENCE INTEGRATED PLANT ALTERNATIVES RELATED TO A DISCOUNT RATE OF 8%

		GOR	Max. Brine Temp., °C	c_W $/m^3	c_E cent/kW(e)·h	c_{eq} cent/kW(e)·h	P_E [a] MW(e)	e_W kW·h/m^3
PWR[b]	MSF-1	13.5	125	1.243	4.65	8.30	458	41.6
	MSF-2	11.5	110	1.281	4.62	8.42	453	44.8
	MSF-3	9.5	98	1.363	4.59	8.76	439	50.7
	MSF-4	7.5	90	1.495	4.56	9.38	417	59.3
	HT-VTE-1	21	120	0.881	4.68	6.97	517	24.7
	HT-VTE-2	17	100	0.905	4.63	6.98	517	27.2
	LT-HTME-1	13.5	70	0.918	4.56	6.91	526	28.2
	LT-HTME-2	11.5	65	0.959	4.51	6.99	521	31.9
	LT-HTME-3	9.5	60	1.023	4.46	7.13	517	36.1
	LT-HTME-4	7.5	55	1.130	4.40	7.41	506	42.9
	RO	N/A.	N/A.	0.716	4.70	6.57	522	17.2
CC[b]	HT-VTE-1	21	120	0.867	4.40	6.45	570	17.7
	HT-VTE-2	17	100	0.884	4.37	6.45	570	19.7
	RO	N A.	N A.	0.710	4.52	6.23	565	11.5

a) saleable electricity supplied to the grid related to average life-time load factor (80%)
b) electricity generation cost of base PWR and combined cycle power plant is 4.72 cent/kW(e)·h and 4.54 cent/kW(e)·h respectively (without water production)

Figure 27 contains the cost composition of the levelized potable water production cost for the integrated plant alternatives divided into the following components:

- capital cost,
- steam cost,
- O&M costs,
- electricity cost, and
- fuel cost of the backup heat boilers.

The above "capital cost" component refers only to the desalination plant. The steam and electricity costs also contain respective capital cost components corresponding to the power plant itself.

The composition of the exergy of fuel required for potable water production in the integrated plant alternatives is shown in Figure 28.

The main important results of the economic assessment can be summarized as follows:

Equivalent electricity generation cost

1) The RO process coupled with the combined cycle or PWR yield the lowest equivalent electricity generation cost (6.23 cent/kW(e)·h for the combined cycle and 6.57 cent/kW(e)·h for the PWR).

2) Among the integrated plant alternatives with distillation plants, the lowest equivalent electricity generation cost is attained by the LT-HTME-1 plant (GOR: 13.5, maximum brine temperature: 70°C) coupled with the PWR (6.91 cent/kW(e)·h), and both HT-VTE plants coupled with the combined cycle (6.45 cent/kW(e)·h).

3) For the MSF plant options, the equivalent electricity generation costs are 1.4 to 2.0 cent/kW(e)·h higher than for the MED plant options with the same GOR.

4) For all distillation plant processes considered (MSF, HT-VTE, LT-HTME), the equivalent electricity generation cost decreases either with increasing GOR or increasing maximum brine temperature.

Potable water production cost

1) The RO process coupled with the combined cycle or the PWR yield the lowest potable water production cost (0.71 $\$/m^3$ for the combined cycle and 0.72 $\$/m^3$ for the PWR).

2) The lowest potable water production cost from distillation plants is provided by the HT-VTE-1 plant (GOR: 21, maximum brine temperature: 120°C), which amounts to 0.87 $\$/m^3$ with the combined cycle and 0.88 $\$/m^3$ with the PWR.

3) For all distillation plant processes considered (MSF, HT-VTE, LT-HTME), the potable water production cost decreases either with increasing GOR or increasing maximum brine temperature.

4) For the MSF plants, the potable water production costs are 0.32 to 0.37 $\$/m^3$ higher than for the MED plant options with the same GOR.

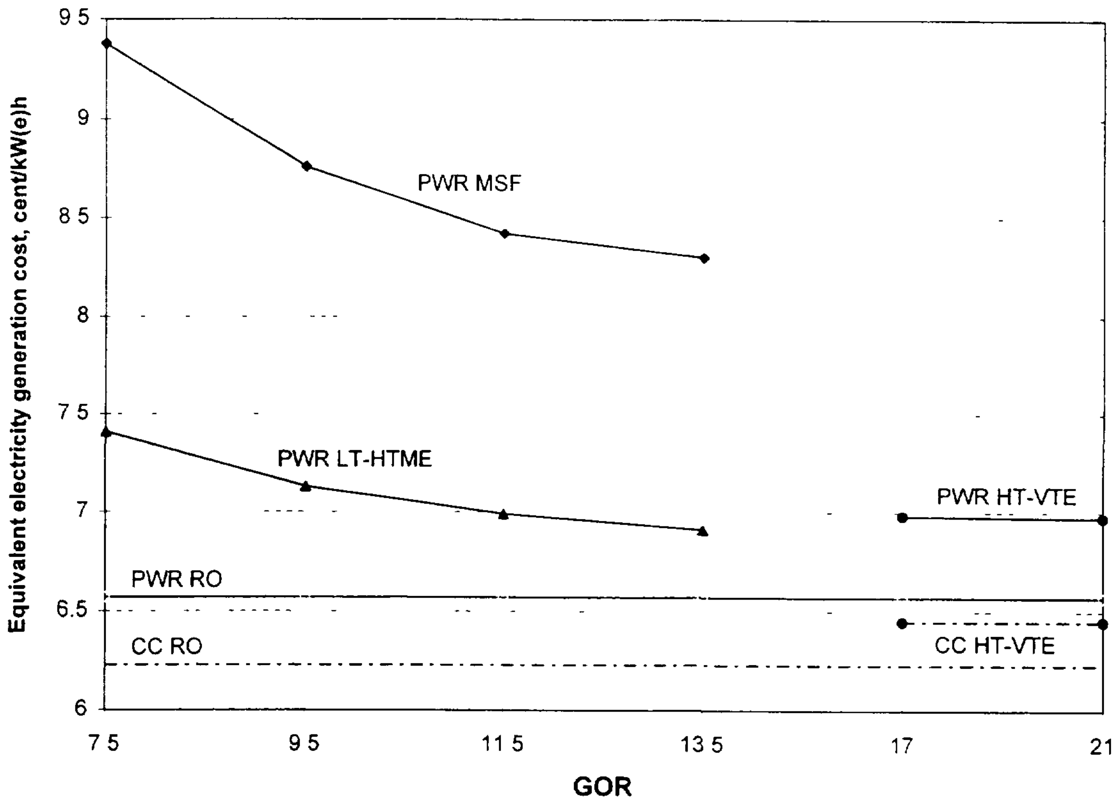


FIG. 23. Levelized equivalent electricity generation cost in cent/kW(e)·h

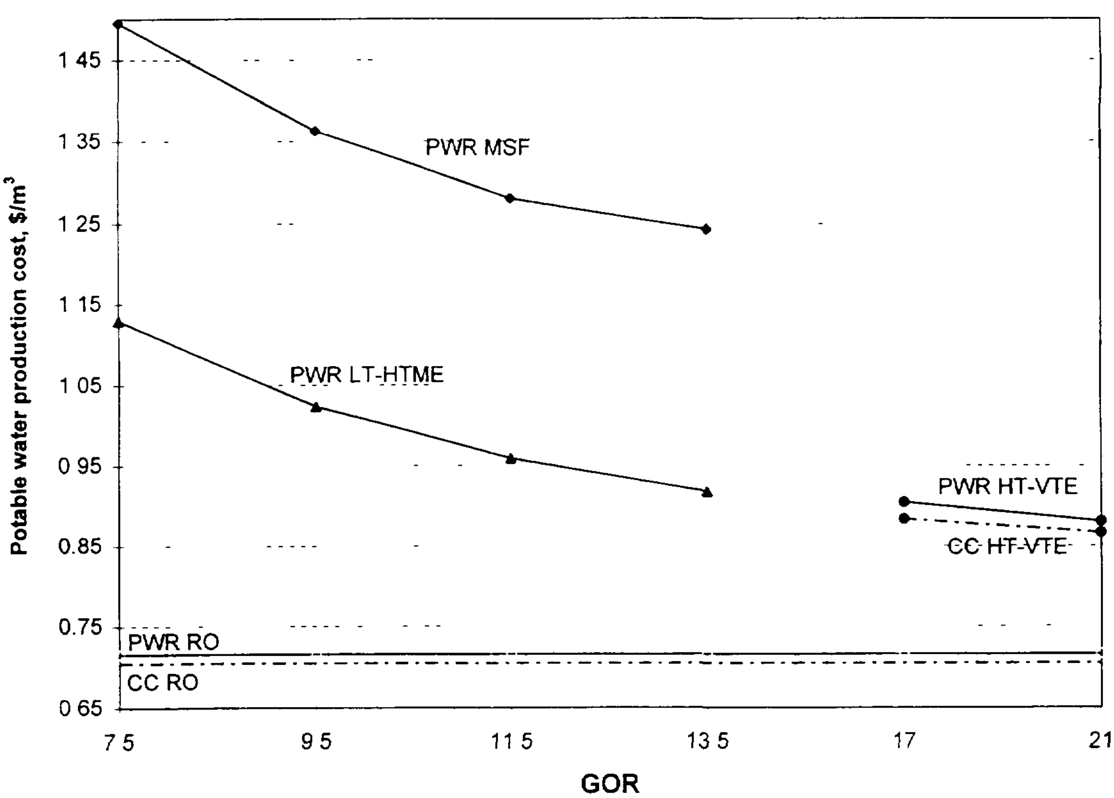


FIG 24 Levelized potable water production cost in $/m³

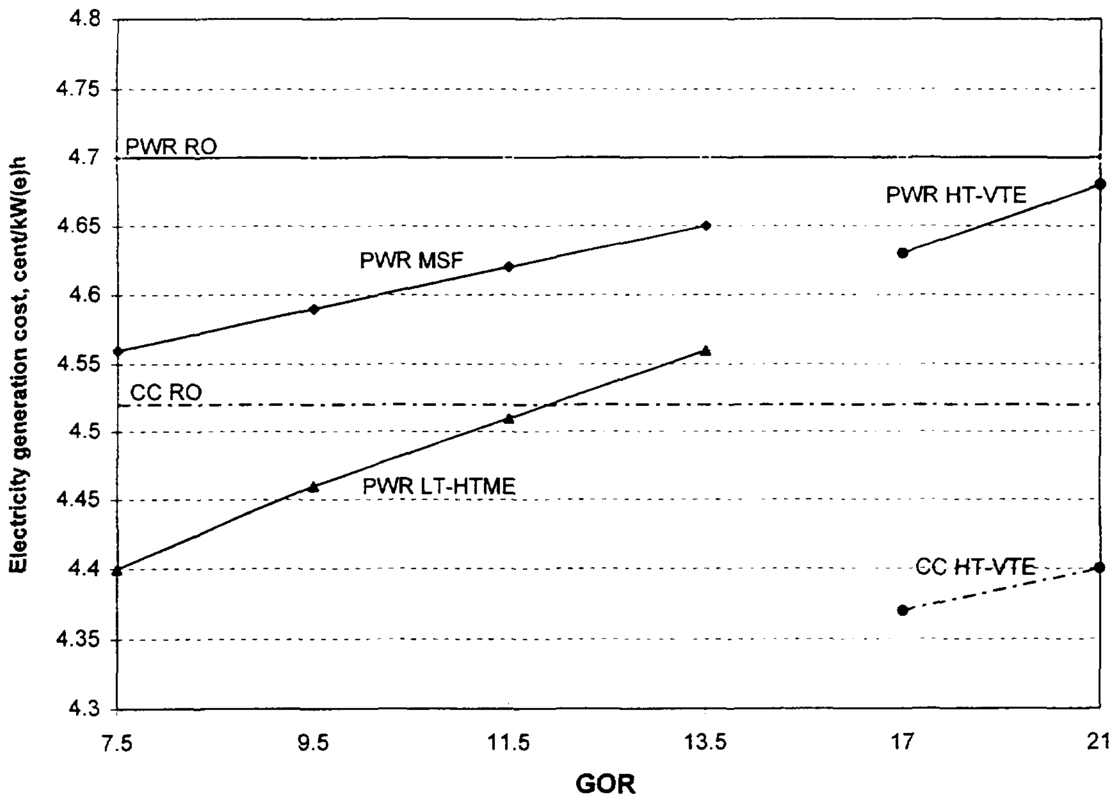


FIG. 25. Levelized electricity generation cost in cent/kW(e)·h.

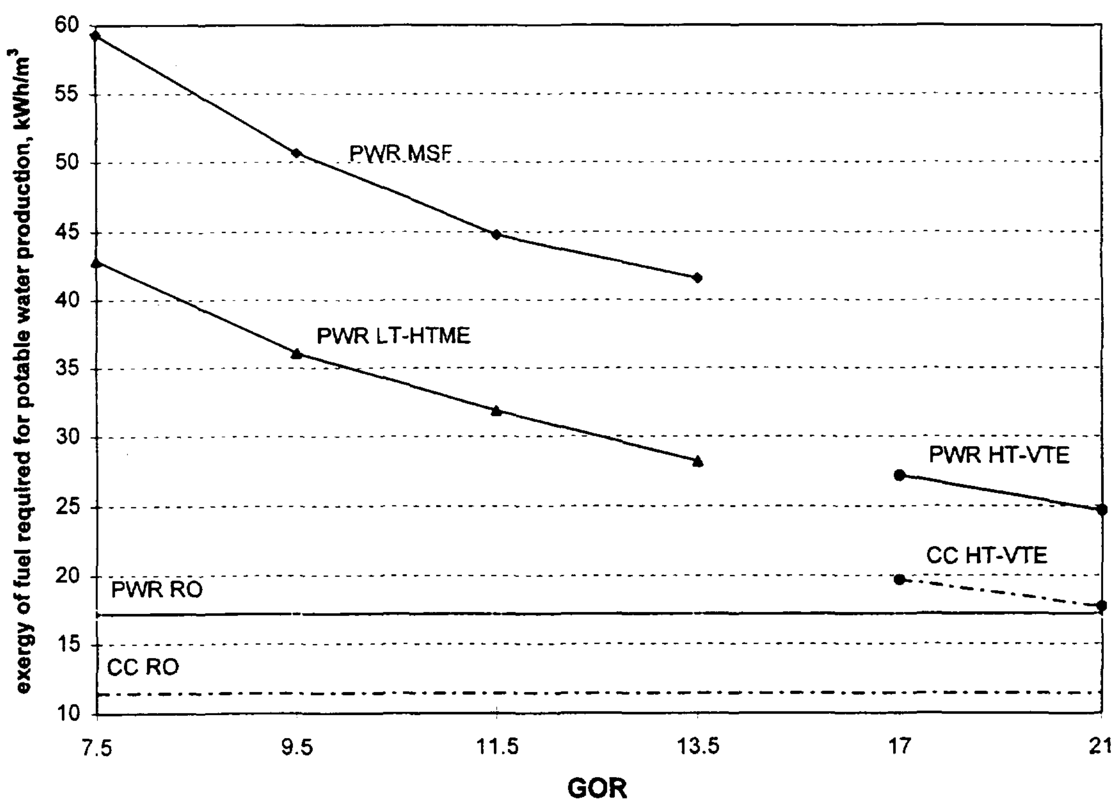


FIG. 26. Exergy of fuel required for potable water production in kW·h/m³.

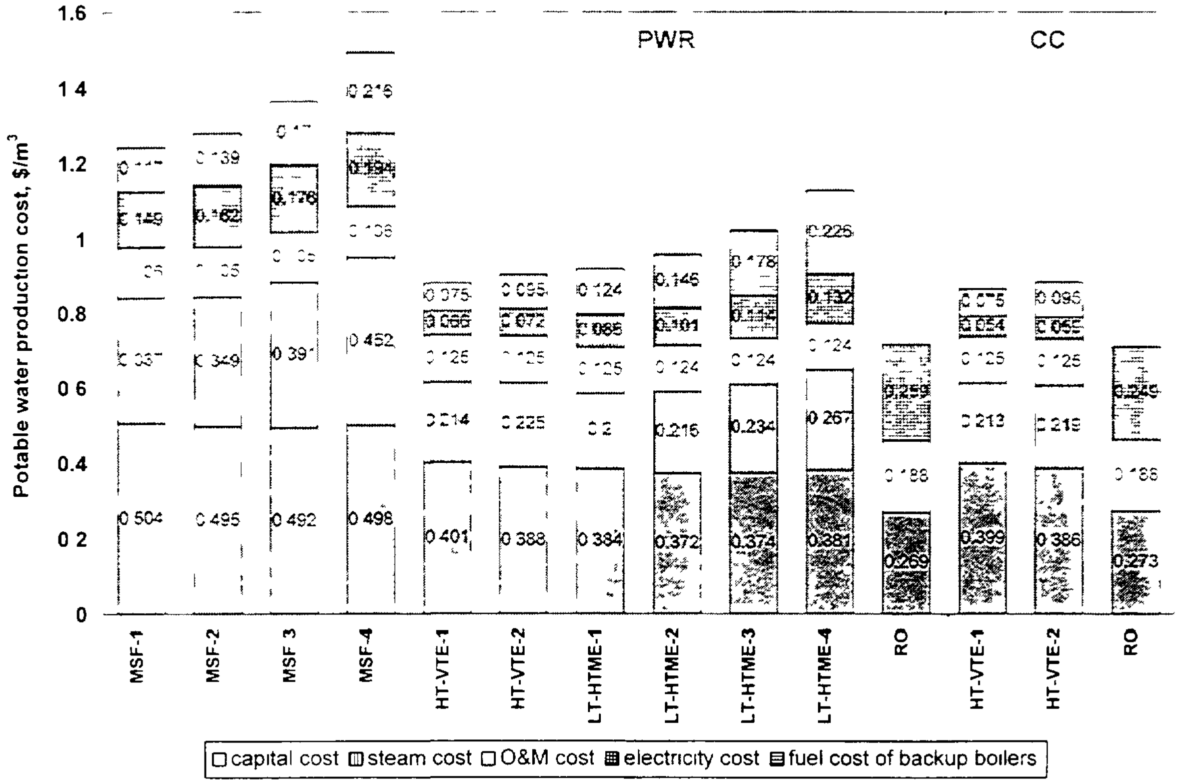


FIG 27 Composition of the levelized potable water production cost

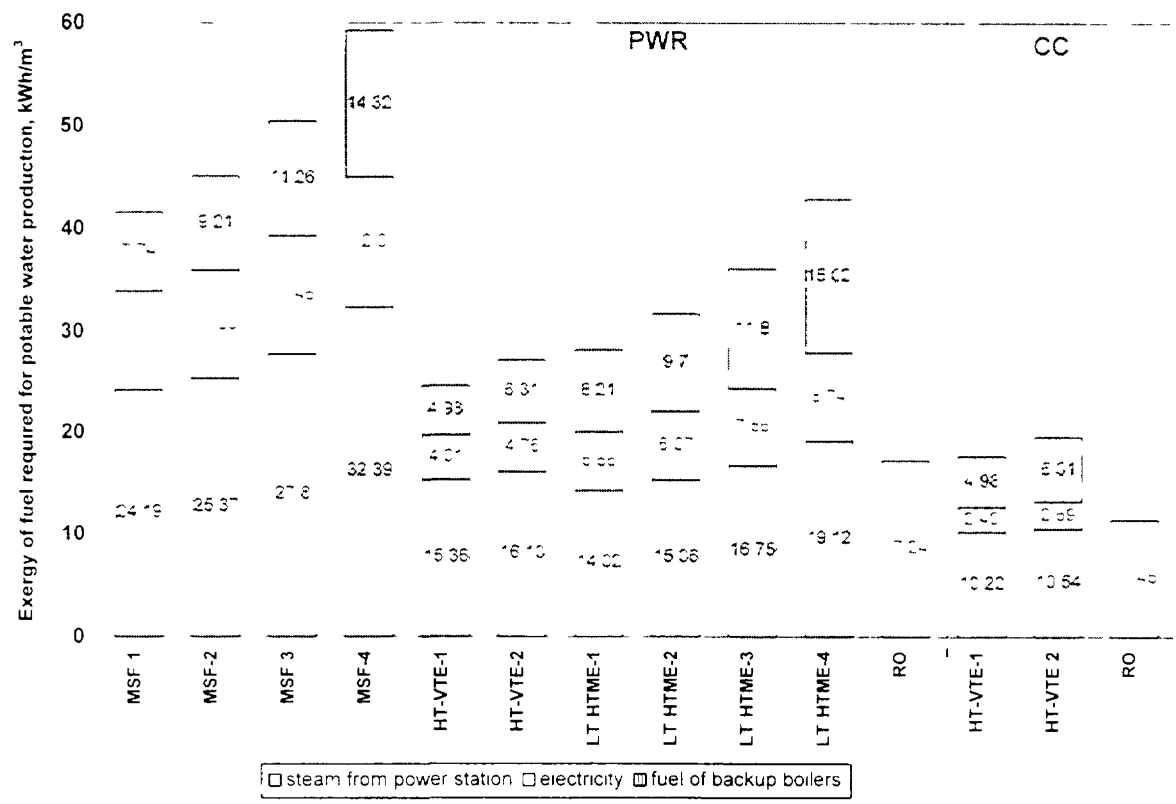


FIG 28 Composition of the exergy of fuel required for potable water production

5) For MSF and LT-HTME plants, the dominant cost factor in potable water production cost is the energy cost component (steam cost, electricity cost, fuel cost of backup heat boilers), followed by the capital cost component. For HT-VTE and RO plants, the capital cost component and the energy cost component are in the same range. The O&M costs component has the smallest impact on the potable water production cost, but it should be remembered that maintenance controls the life, reliability and availability of each plant. The portion of O&M costs is higher for the RO plant than for MED or MSF plants, because of the required membrane replacements.

Electricity generation cost

1) For the distillation plant alternatives, the higher the potable water production cost, the lower is the electricity generation cost.

2) For the MSF plant alternatives, the electricity generation costs are 0.1 to 0.15 cent/kW(e)·h higher than for the MED plant options with the same GOR.

3) The electricity generation cost for the RO plant alternative is higher than for the distillation plant alternatives (i.e. less exergy of fuel of the energy source is needed).

Exergy of fuel required for potable water production

1) The RO process coupled with the combined cycle or the PWR requires the lowest exergy of fuel for potable water production (11.5 kW·h/m^3 for the combined cycle and 17.2 kW·h/m^3 for the PWR).

2) Among the integrated plant alternatives with distillation plants, the lowest exergy of fuel for potable water production is required by the HT-VTE-1 plant (17.7 kW·h/m^3 with the combined cycle and 24.7 kW·h/m^3 with the PWR).

3) For all distillation plant processes considered (MSF, HT-VTE, LT-HTME), the exergy of fuel for potable water production decreases either with increasing GOR or increasing maximum brine temperature.

4) For the MSF plants, the exergy of fuel for potable water production is 13 to 16 kW·h/m^3 higher than that for the MED plant options with the same GOR.

4.5.5. Sensitivity analysis

All values of parameters and costs adopted for the economic assessment are best estimates based on available information, experience and engineering judgement (see Section 4.5.3). They correspond to reference values within reasonable ranges. The parameters which have the largest effect on the economic results are the discount rate, the escalation rate of crude oil price, uncertainties in the costs of new designs of plants and equipment as well as the reliability and availability of both the energy sources and seawater desalination plants.

As an example, a sensitivity analysis was performed for the discount rate and the escalation rate of crude oil price. The effects of using real discount rates of 5% and 10% (reference value 8%) and of real crude oil price escalation rates of 0% and 4% (reference value 2%) respectively on potable water production cost and equivalent electricity generation cost are shown in Figure 29.

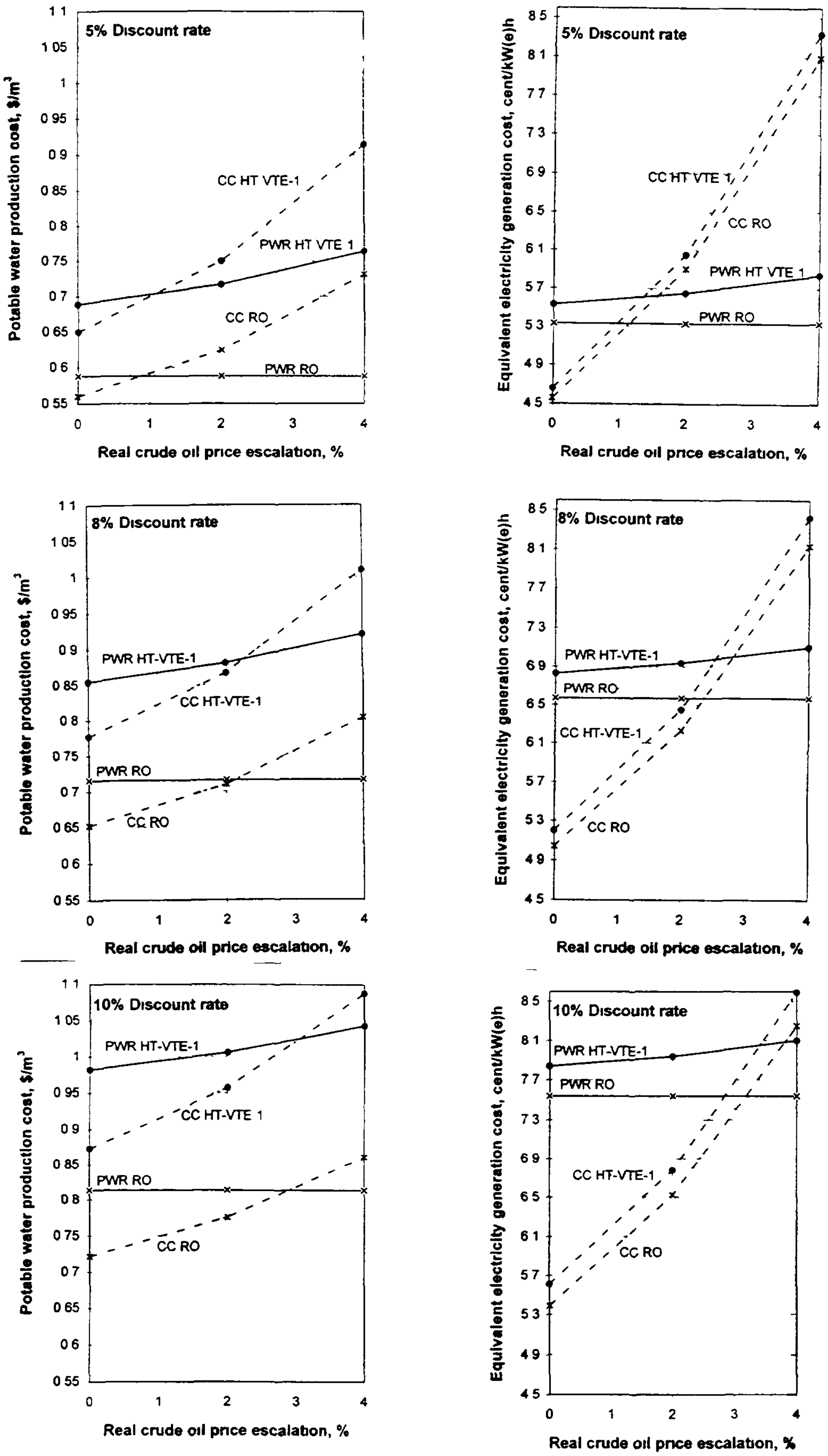


FIG 29 Effect of real discount rate and real crude oil price escalation on potable water production cost and equivalent electricity generation cost

For clarity, only the results of the RO plant alternative and the distillation plant alternative with the lowest potable water production cost/equivalent electricity generation cost for each energy source are shown in the figures.

The results of the sensitivity analysis can be summarized as follows:

1) A real discount rate of 5% will make the nuclear energy source more economic, while for 8 and 10% the fossil energy source is cheaper.

2) Higher real escalation rates of crude oil price tend to favour the nuclear energy source.

3) The RO plant alternative is the most economic solution for each energy source independent of the discount rate and crude oil price escalation rate, within the sensitivity range considered.

5. CONCLUDING REMARKS

5.1. GENERAL

Seawater desalination in integrated co-production plants is one of the most promising options to counteract the freshwater scarcity in arid coastal regions.

For selecting the most economic integrated plant configuration, the methodology of calculating the equivalent electricity generation cost is appropriate to rank different integrated plants.

For allocating the production cost of electricity and potable water, the exergetic cost allocation method is applied. This method leads to somewhat higher potable water production cost and somewhat lower electricity generation cost than the power credit method. While the power credit method allocates all the benefits of co-production to potable water, the exergetic method distributes them to electricity and potable water according the exergy consumption of the processes. From the thermodynamic viewpoint, this is the most equitable method of allocating both investment and operating costs to electricity and potable water production.

The exergetic cost allocation method gives utilities an equitable basis for costing electricity and potable water, both for new and already existing plants. Tariffs established on this basis could also give a clearer message to the consumer on the value of these products, and could lead to more efficient use than subsidized tariffs.

Both the exergetic cost allocation method and the methodology of calculating the equivalent electricity generation cost is valid for any type of nuclear, fossil or renewable energy application, including also plants with district heating or industrial heat supply, as well as seawater desalination.

5.2. CONSIDERATIONS OF THE ECONOMIC ASSESSMENT FOR THE REPRESENTATIVE SITE

A comparative assessment of all possible energy sources for co-production of electricity and potable water requires comparing a wide range of available energy options, including nuclear power, fossil fuels, renewable energies, waste recovery, etc. However, previous studies performed by the IAEA have shown that nuclear power plants and fossil fired combined cycle power plants are the economically most attractive energy sources for large-scale co-production of electricity and potable water [16, 17].

The results of the economic assessment performed in Chapter 4 is only valid for the reference case considered. When varying the demand for electricity and potable water as well as the economic assumptions, other integrated plant alternatives could become the most economical solution.

Uncertainties in the reliability of plant operation as well as unforeseen escalation of costs are further criteria which have to be taken into account. For example, a technically unproven plant design may result in higher outage rates, higher capital and O&M costs as well as in shorter economic life than considered in the economic assessment.

In this connection, MSF plants have, as a result of their long and successful experience, the greatest degree of certainty. For MED plants, which in principle should have a similar degree of certainty as MSF plants, based on the simplicity of the process and utilization of similar materials, there is not yet adequate experience of operation of large-scale units. For both MSF and MED plants, scaling and corrosion problems will increase with increasing maximum brine temperature, and, as a result, the risk of higher outage rates will increase as well. In RO plants, especially in the Gulf, the water pre-treatment is the weakest point during operation, which has led in the past to high outage rates and premature membrane replacement in some cases.

Both energy sources considered — the PWR and the combined cycle power plant — are based on mature technologies, so that a reliable operation can be assumed.

Some countries have cheaper fossil fuel resources than what has been assumed in the economic assessment. The use of nuclear energy for co-production of electricity and potable water, however, would save fossil energy resources that could be sold at world market prices.

In the economic assessment which has been performed, the costs of water transport and distribution were not considered. The cost components are site dependent and can only be analysed on a case by case basis. While water distribution costs, which depend only on the particular characteristics of the consumption centres, would be essentially the same for the supply of potable water from any energy source, water transport costs could be higher for nuclear integrated plants than for the fossil alternative, due to more stringent siting requirements for the nuclear plant.

As shown in Figures 27 and 28, the use of backup heat boilers to increase the load factor of the distillation plants is in the exergetic viewpoint a wasteful measure, and even with only a low load factor, an expensive one. In individual cases, it is to be examined whether an alternate source of potable water can be provided during the time when the power plant is not in operation. For LT-HTME plants, the use of thermal vapour compression in the backup heat system might reduce the energy consumption of the distillation plant.

Annex I

TECHNICAL PERFORMANCE DATA OF RO TRAINS

(Hollow fibre membranes at seawater temperatures of 22°C and 35°C [19])

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DUPONT "PERMASEP" Products Projection: OPUS 4.0

ID: HWP

Project: Nuclear Power Technology

31.10.1995

Design Temp = 22 °C | Max Temp = 22 °C

Feed Pressure = 75 bar | Product Pressure = .7 bar

Overall Conversion = 35 % | Design Period = 43800 hrs

Plant Capacity: Not specified | Bal. Tube = 2.4 bar

Perm. Model 6880TWIN (14000 gpd) | Salt Passage = .65 %

Recycle Fraction = 0 % | SUMMARY

Stage No.	Permeators per Stage	Conversion %	Press. Drop bar	Flow/Perm Feed [lpm]	Flow/Perm Brine [lpm]	Flow/Perm Prod. [lpm]	
1	795,0	35,0	0,4	61,4	39,9	21,5	
Stage No.	**MFRC -**	**Feed Press bar**	**Feed Flow m³/d**	**Brine Press bar**	**Brine Flow m³/d**	**Prod. Press bar**	**Prod. Flow m³/d**
1	0,645	75,0	70262,1	74,6	45670,3	0,7	24591,7
Overall							24591,7

Cations	Raw Feed mg ion/l	Acid Feed mg ion/l	Acid Feed mg CaCO3/l	Brine mg ion/l	Brine mg CaCO3/l	Product mg ion/l	Product mg CaCO3/l
Ca	522,0	522,0	1302,4	802,3	2001,8	1,4	3,4
Mg	1659,0	1659,0	6821,8	2549,9	10485,4	4,4	18,0
Na	13793,8	13793,8	30001,6	21166,1	46036,3	102,5	222,9
K	496,0	496,0	634,4	761,3	973,7	3,3	4,2
Sr	0,0	0,0	0,0	0,0	0,0	0,0	0,0
Ba	0,00	0,00	0,00	0,00	0,00	0,00	0,00
Fe	0,0	0,0	0,0	0,0	0,0	0,0	0,0
Total	16470,8	16470,8	38760,2	25279,7	59497,2	111,5	248,5
Anions							
HCO3	185,0	147,2	120,5	221,1	181,1	9,8	8,0
SO4	3585,0	3594,8	3742,2	5525,3	5751,8	9,5	9,9
Cl	24750,0	24750,0	34897,5	37988,6	53564,3	163,6	230,6
F	0,0	0,0	0,0	0,0	0,0	0,0	0,0
NO3	0,0	0,0	0,0	0,0	0,0	0,0	0,0
PO4	0,0	0,0	0,0	0,0	0,0	0,0	0,0
CO3	0,0	0,0	0,0	0,0	0,0	0,0	0,0
Total	28500,0	28491,9	38760,2	43735,3	59497,2	182,9	248,5
CO2	3,0	30,3		30,3		30,3	
H2S	0,0	0,0		0,0		0,0	
SiO2	43,0	43,0		65,9		0,4	
TDS [ppm]	45013,8	45005,8		69080,9		294,8	
pH	8,00	6,90		6,93		5,72	
Osmotic press. [psi]		483,2		774,1		3,3	
Equiv. NaCl [ppm]		40791,4		63781,5		293,6	
S&DSI	-0,69	H2SO4	30,37	[ppm]	H2SO4	0,72	[lbs/kg prod]
max. allow. Conv. %	w/o	with	Antiscalant				
CaSO4	56,5	65,1					
BaSO4	100,0	100,0					
SrSO4	100,0	100,0					
CaF2	100,0	100,0					
SiO2	73,3						

This Software is supplied by 'Permasep' Products to aid in the design of RO systems. For further assistance to optimize design and assure acceptance of warranty, it is recommended that you contact a 'Permasep' Marketing Representative.

DUPONT "PERMASEP" Products Projection: OPUS 4.0

ID: HWP
31.10.1995

Project: Nuclear Power Technology

Design Temp = 35 °C	Max Temp = 35 °C
Feed Pressure = 67 bar	Product Pressure = .7 bar
Overall Conversion = 35 %	Design Period = 43800 hrs
Plant Capacity: Not specified	Bal. Tube = 2.4 bar
Perm. Model 6880TWIN (14000 gpd)	Salt Passage = .65 %
Recycle Fraction = 0 %	SUMMARY

Stage No.	Permeators per Stage	Conversion %	Press. Drop bar	Flow/Perm Feed [lpm]	Flow/Perm Brine [lpm]	Flow/Perm Prod. [lpm]	
1	795,0	35,0	0,4	60,0	39,0	21,0	
Stage No.	**MFRC -**	**Feed Press bar**	**Feed Flow m³/d**	**Brine Press bar**	**Brine Flow m³/d**	**Prod. Press bar**	**Prod. Flow m³/d**
1	0,645	67,0	68646,2	66,6	44620,0	0,7	24026,2
Overall							24026,2

	Raw Feed	Acid Feed	Acid Feed	Brine	Brine	Product	Product
Cations	mg ion/l	mg ion/l	mg CaCO3/l	mg ion/l	mg CaCO3/l	mg ion/l	mg CaCO3/l
Ca	522,0	522,0	1302,4	802,0	2001,0	2,0	5,0
Mg	1659,0	1659,0	6821,8	2548,9	10480,9	6,4	26,3
Na	13793,8	13793,8	30001,6	21141,7	45983,2	147,8	321,5
K	496,0	496,0	634,4	760,5	972,7	4,8	6,1
Sr	0,0	0,0	0,0	0,0	0,0	0,0	0,0
Ba	0,00	0,00	0,00	0 00	0,00	0,00	0,00
Fe	0,0	0,0	0,0	0,0	0,0	0 0	0,0
Total	16470,8	16470,8	38760,2	25253,1	59437,8	161,0	358,9
Anions							
HCO3	185,0	147,2	120,5	221,1	181,1	9,8	8,0
SO4	3565,0	3594,8	3742,2	5523,0	5749,4	13,9	14,4
Cl	24750,0	24750,0	34897,5	37948,4	53507,3	238,6	336,5
F	0,0	0,0	0,0	0,0	0,0	0,0	0 0
NO3	0,0	0,0	0,0	0,0	0,0	0,0	0,0
PO4	0,0	0,0	0,0	0 0	0,0	0,0	0,0
CO3	0,0	0,0	0,0	0,0	0,0	0,0	0,0
Total	28500,0	28491,9	38760,2	43692,5	59437,8	262 3	358,9
CO2	3,0	30,3		30 3		30 3	
H2S	0,0	0,0		0,0		0,0	
SiO2	43,0	43,0		65,8		0 6	
TDS [ppm]	45013,8	45005,6		69011,4		423,9	
pH	8 00	6,99		6,93		5,72	
Osmotic press. [psi]		505,9		811 4		5 0	
Equiv. NaCl [ppm]		40895,9		64018,4		421,7	
S&DSI	-0,44	H2SO4	30,37	[ppm]	H2SO4	0,72	[lbs/kg prod]
max. allow. Conv. %	w/o	with	Antiscalant				
CaSO4	58,2	65,9					
BaSO4	100,0	100,0					
SrSO4	100,0	100,0					
CaF2	100,0	100,0					
SiO2	78,1						

This Software is supplied by 'Permasep' Products to aid in the design of RO systems For further assistance to optimize design and assure acceptance of warranty, it is recommanded that you contact a 'Permasep' Marketing Representative

Annex II

TECHNICAL PERFORMANCE DATA OF GAS TURBINES

(ABB GT13E2 at annual average air conditions [22])

gas turbine	GT13E2	
fuel	METHANE	
lower heating value	50 000 kJ/kg	
burner	EV BURNER	
ambient pressure	1.013	bar
ambient temperature	28.5	°C
relative humidity	60	%
intake pressure loss	10	mbar
exhaust pressure loss	30	mbar
power factor	0.80	
frequency	50	Hz
load	base	
gross power	146.5	MW
gross efficiency	34.1	%
fuel mass flow	8.578	kg/s
water mass flow	0.0	
exhaust gas flow	494	kg/s
exhaust gas enthalpy	593	kJ/kg
exhaust gas temperature	541	°C
oxygen	13.75	Vol.%
nitrogen	73.94	Vol.%
carbon dioxide	3.07	Vol.%
water	8.31	Vol.%
argon	0.93	Vol.%
sulphur dioxide	0.00	vppm
nitrogen oxides	25	vppm

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Annex III

EXERGY FLOW DIAGRAMS OF THE PWR POWER PLANT COUPLED WITH THE REFERENCE DISTILLATION PLANTS

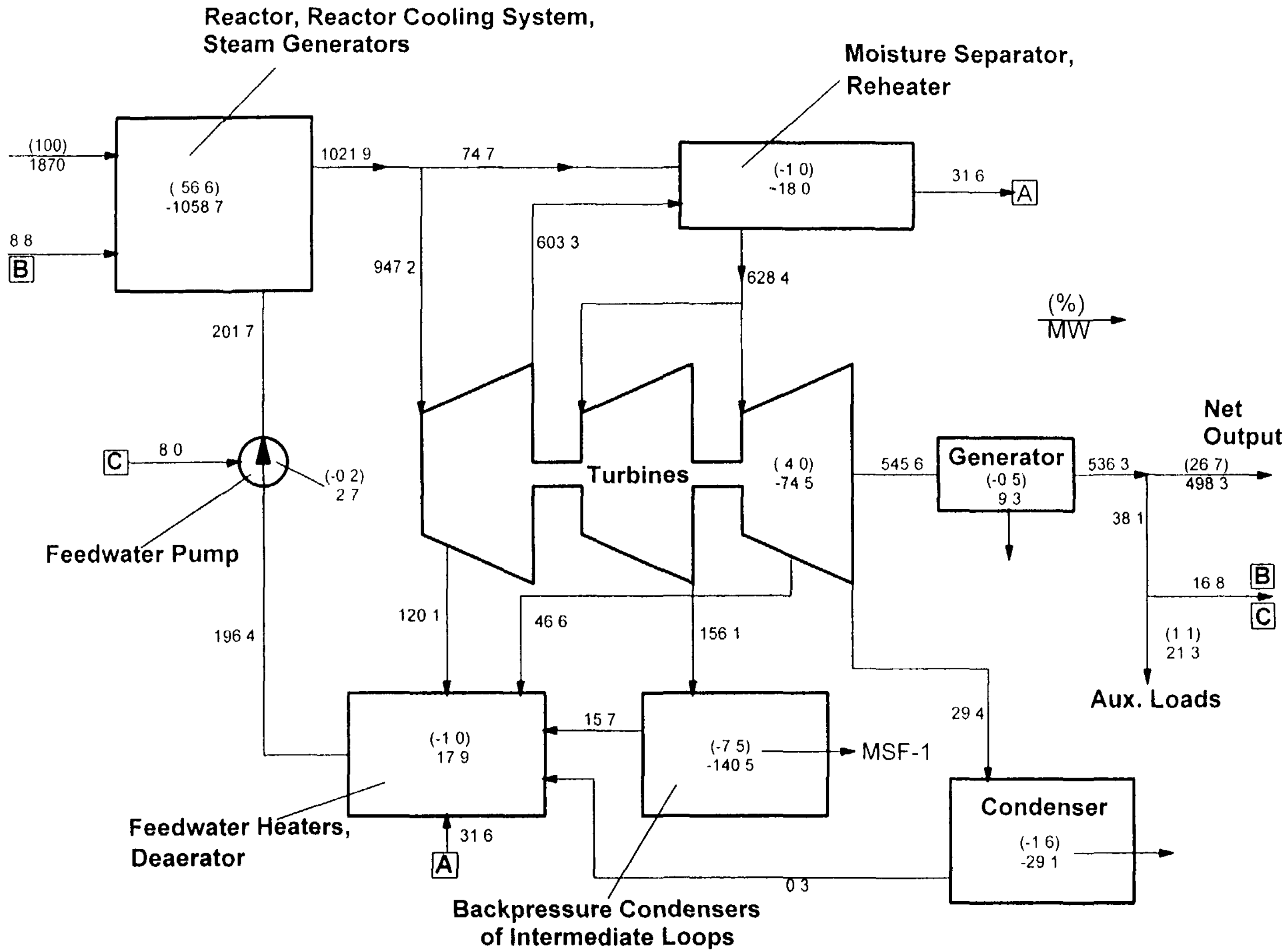

FIG. III.1. Exergy flow diagram of the PWR power plant coupled with the MSF-1 plant.

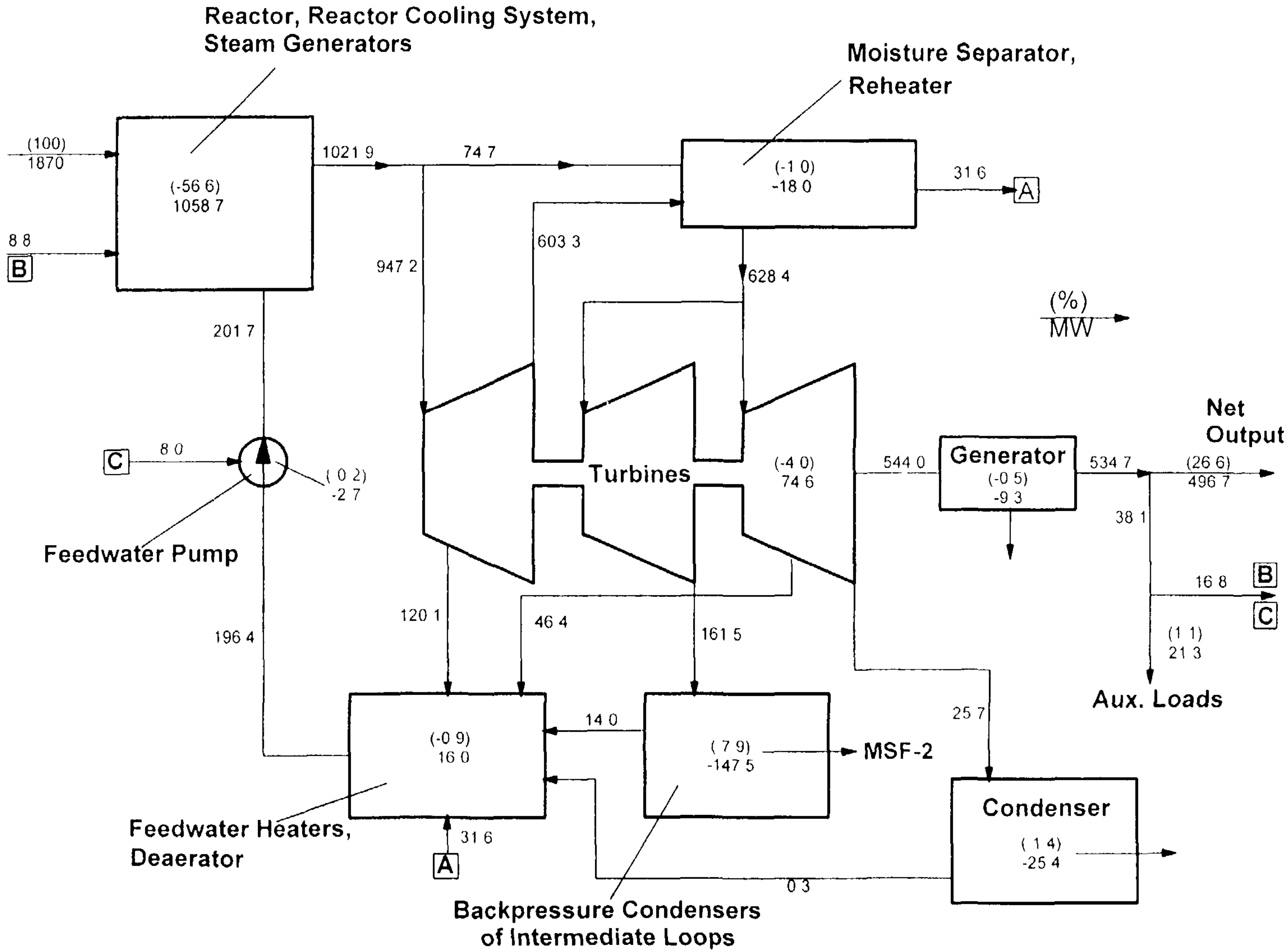

FIG. III.2. Exergy flow diagram of the PWR power plant coupled with the MSF-2 plant.

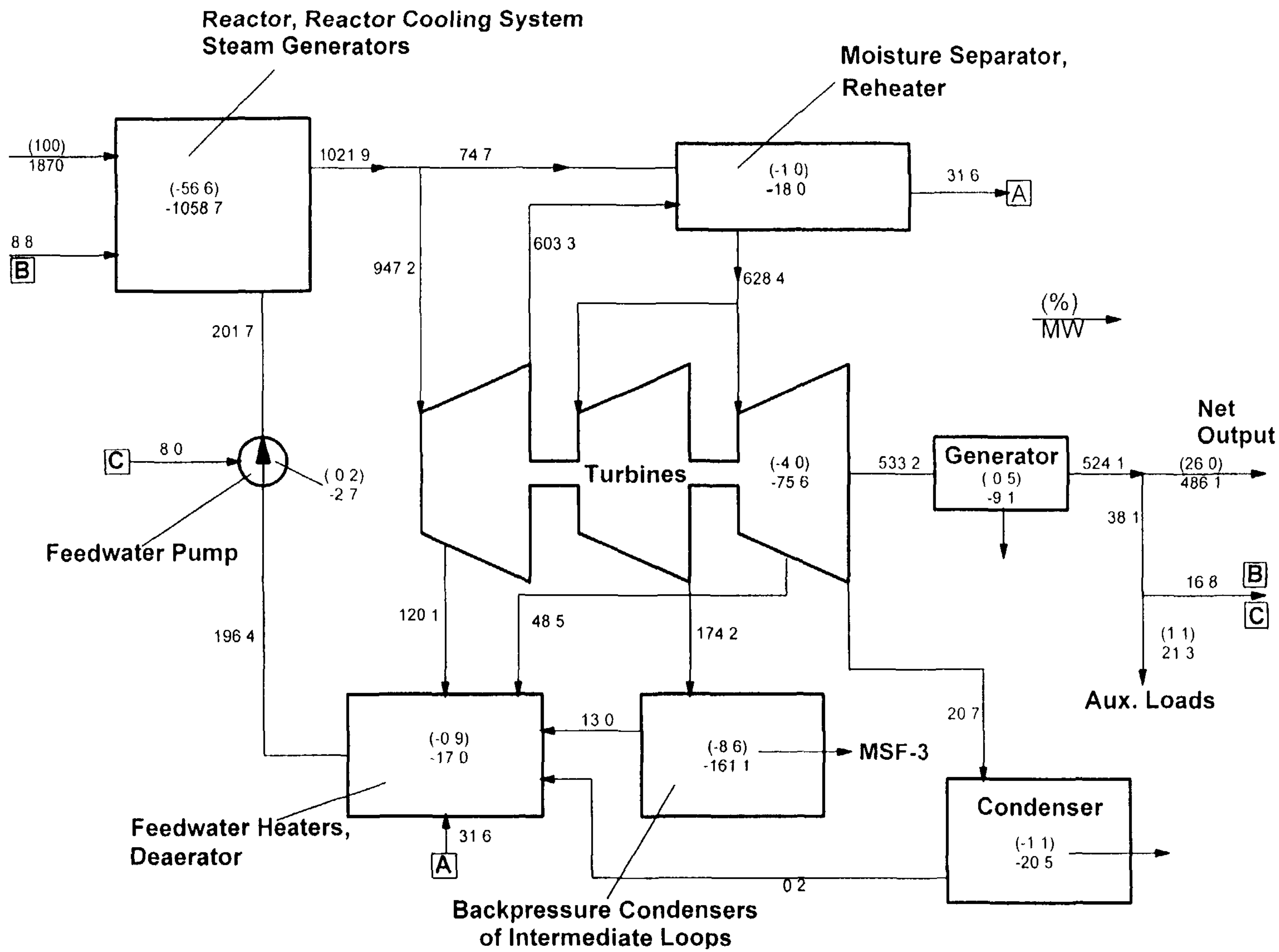

FIG. III.3. Exergy flow diagram of the PWR power plant coupled with the MSF-3 plant.

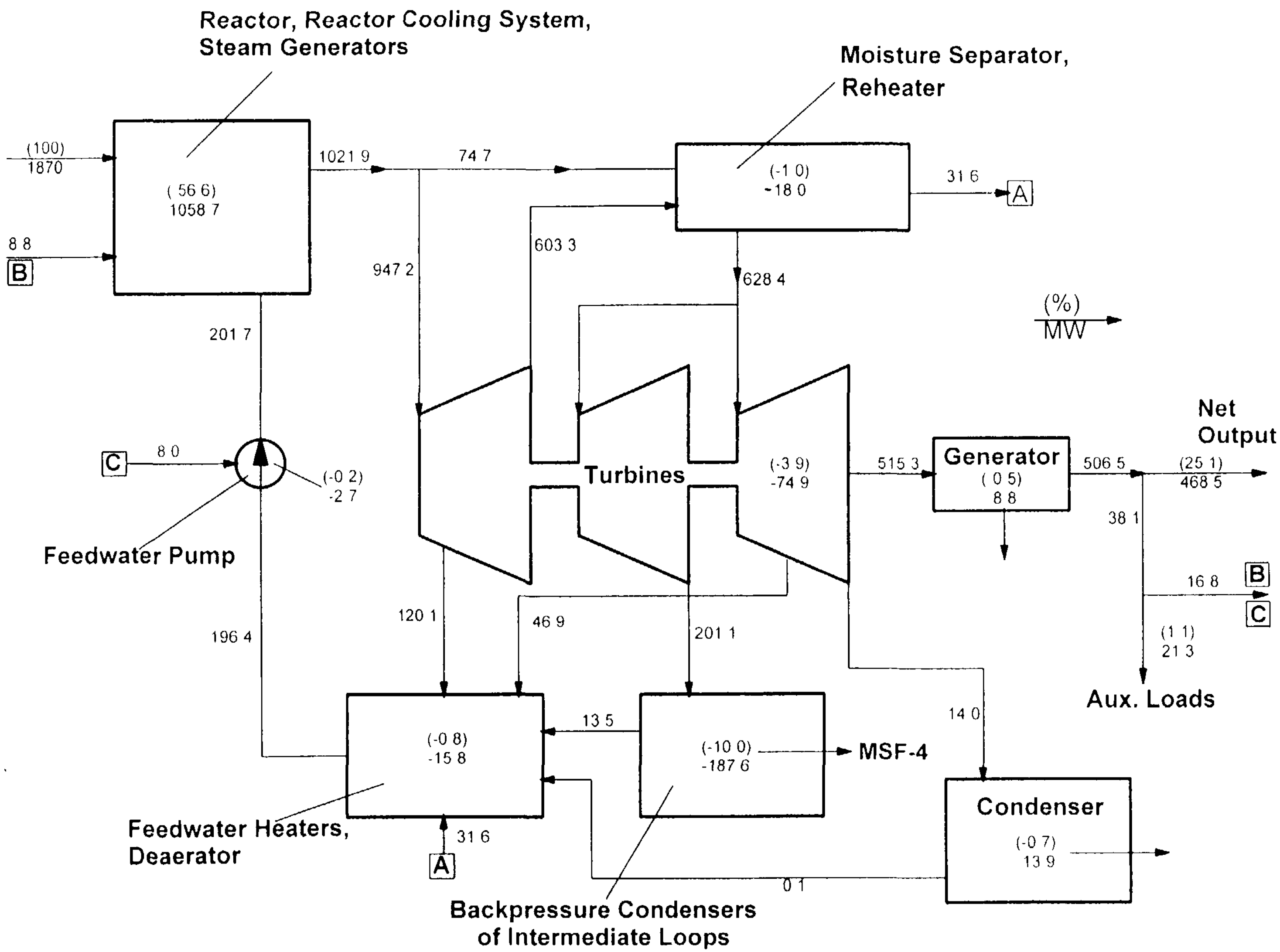

FIG. III.4. Exergy flow diagram of the PWR power plant coupled with the MSF-4 plant.

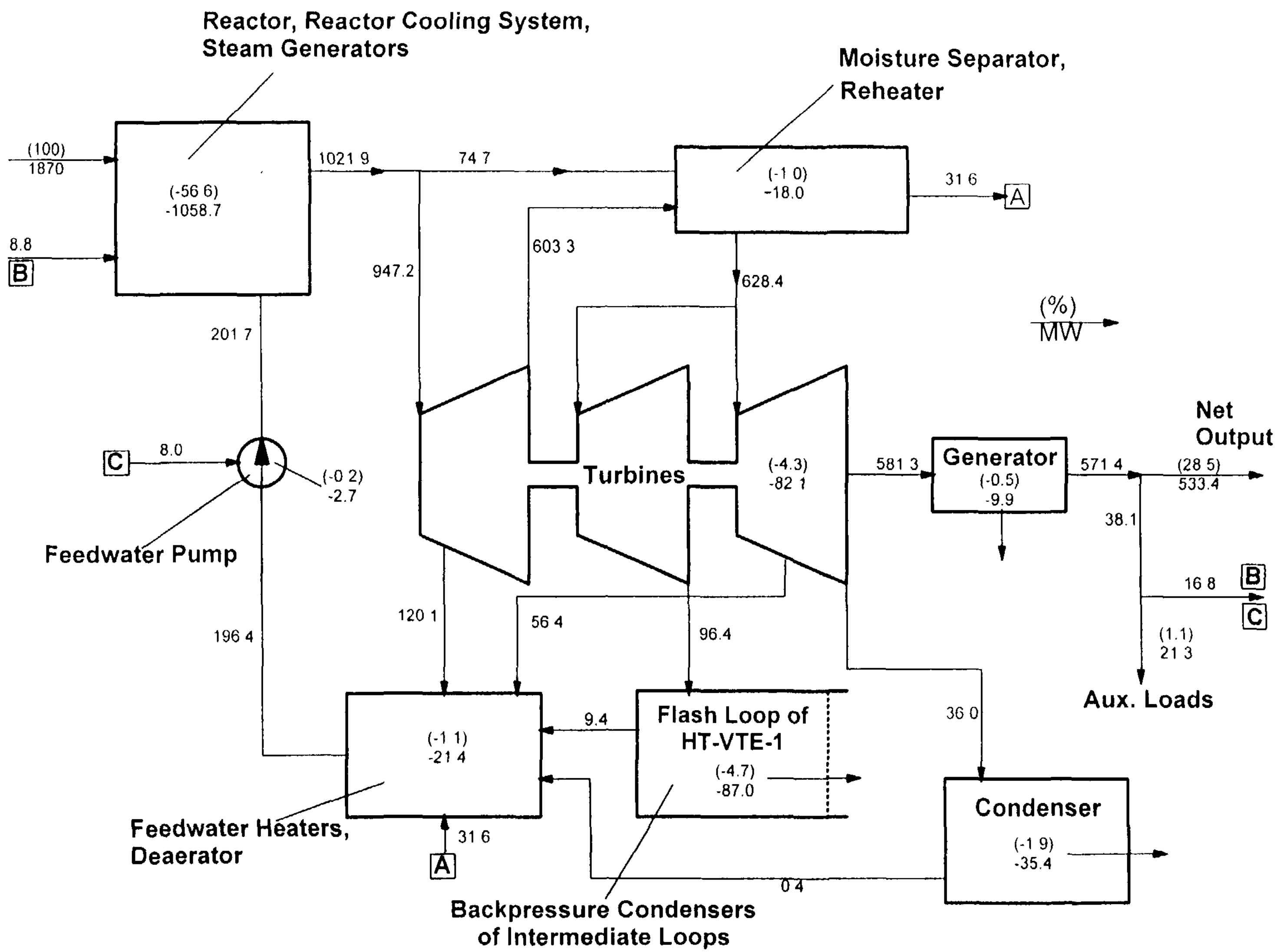

FIG. III.5. Exergy flow diagram of the PWR power plant coupled with the HT-VTE-1 plant.

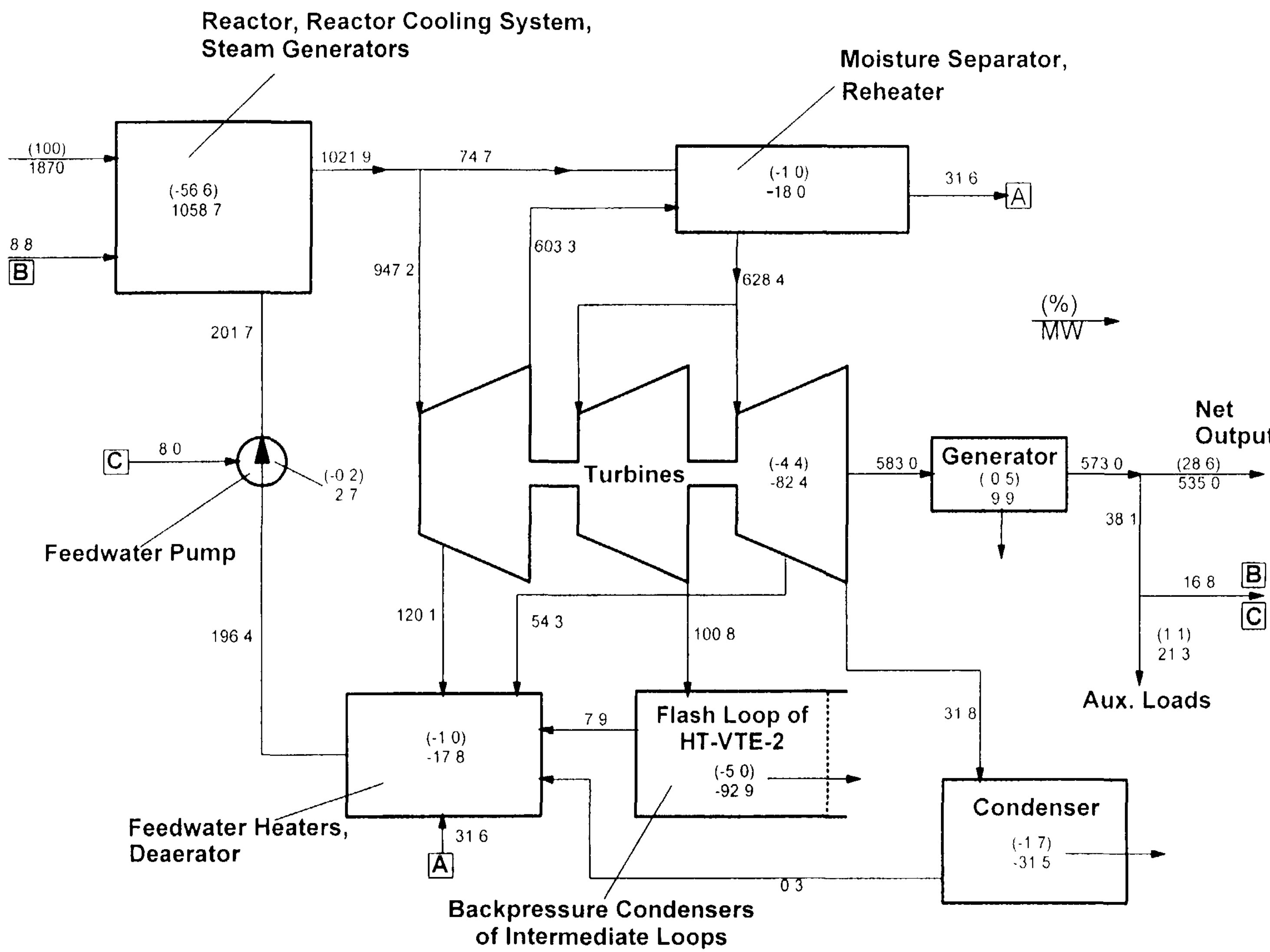

FIG. III.6. Exergy flow diagram of the PWR power plant coupled with the HT-VTE-2 plant.

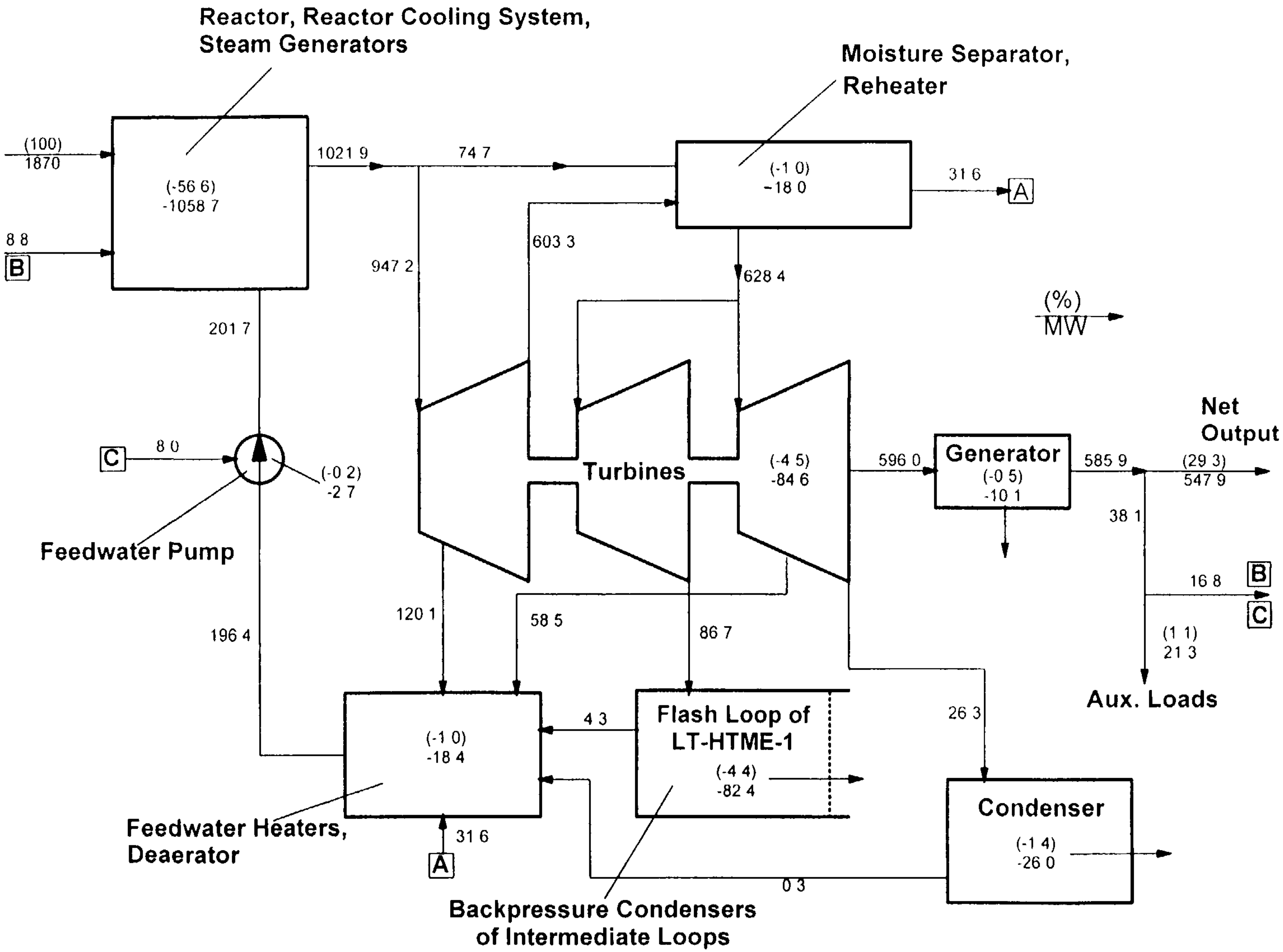

FIG. III.7. Exergy flow diagram of the PWR power plant coupled with the LT-HTME-1 plant.

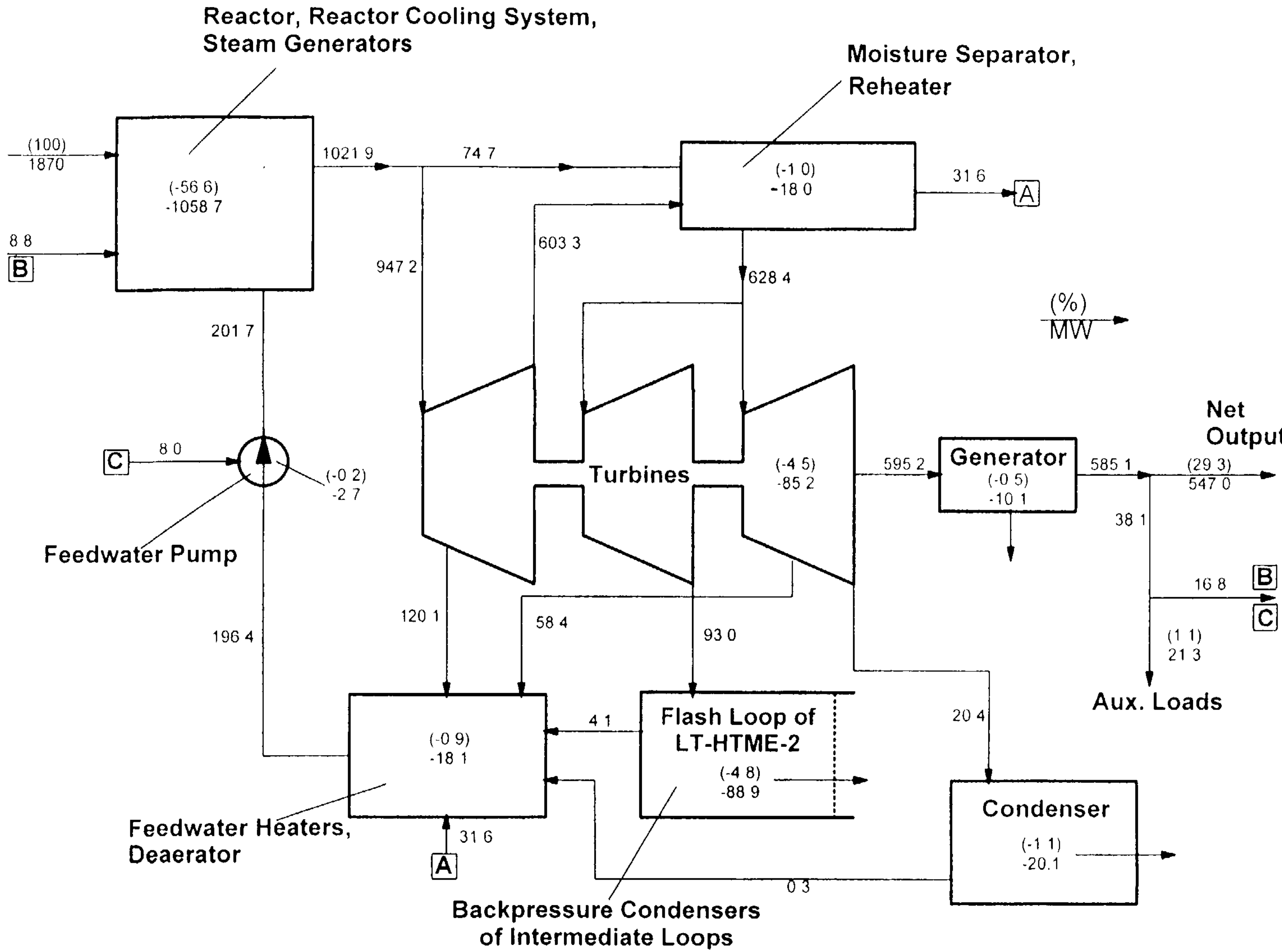

FIG. III.8. Exergy flow diagram of the PWR power plant coupled with the LT-HTME-2 plant.

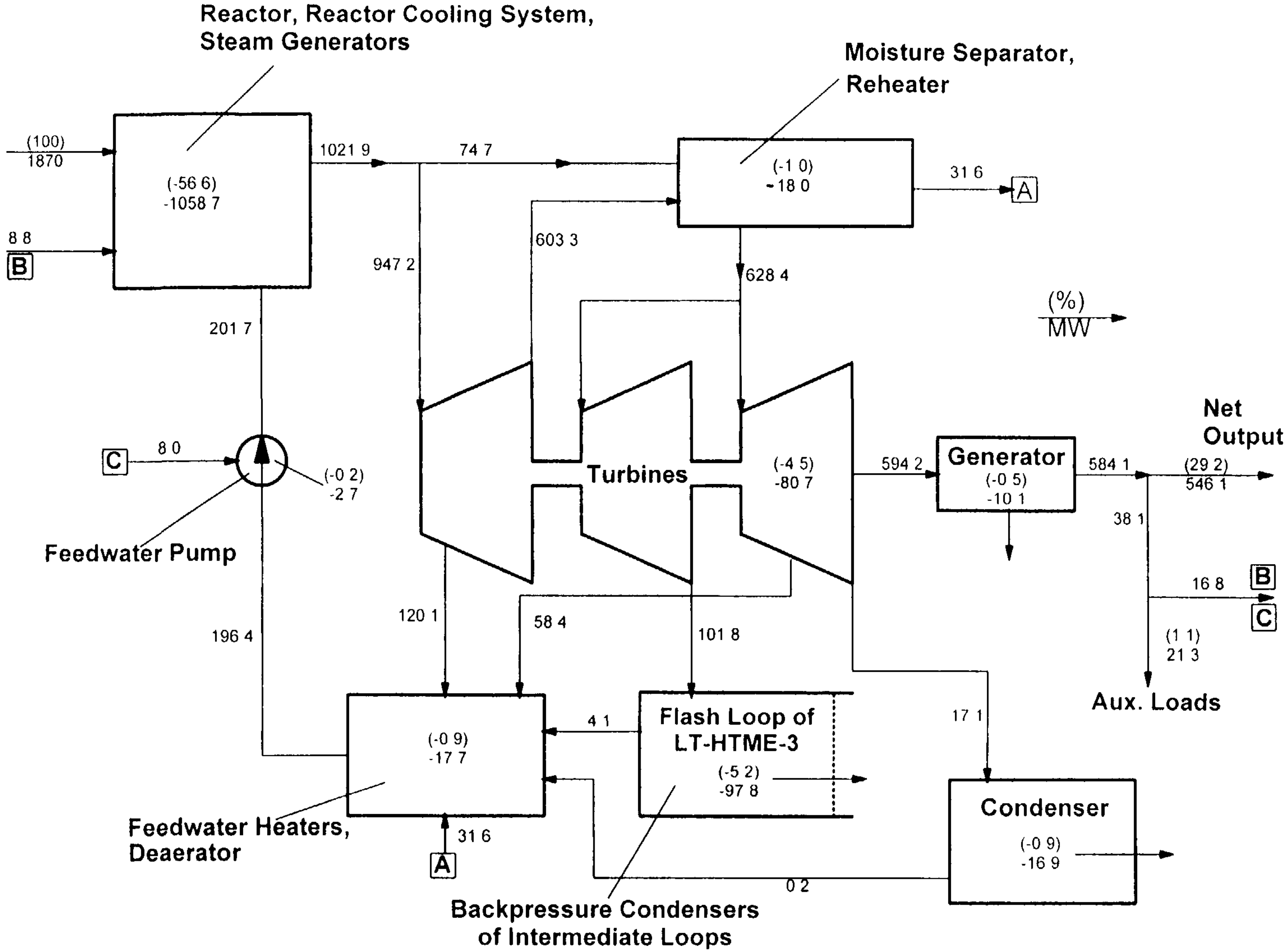

FIG. III.9. Exergy flow diagram of the PWR power plant coupled with the LT-HTME-3 plant.

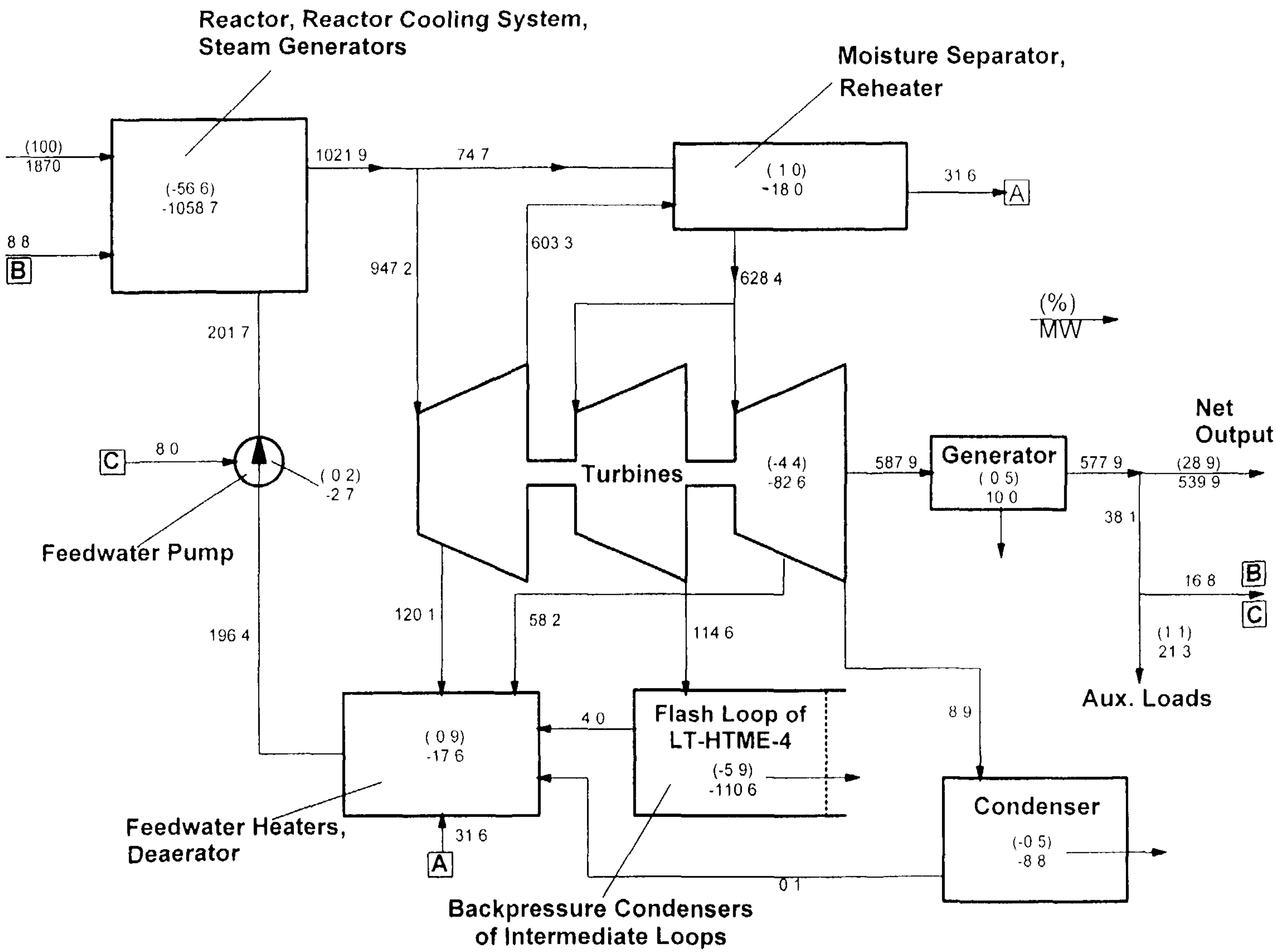

FIG. III.10. Exergy flow diagram of the PWR power plant coupled with the LT-HTME-4 plant.

Annex IV

EXERGY FLOW DIAGRAMS OF THE COMBINED CYCLE POWER PLANT COUPLED WITH THE REFERENCE HT-VTE PLANTS

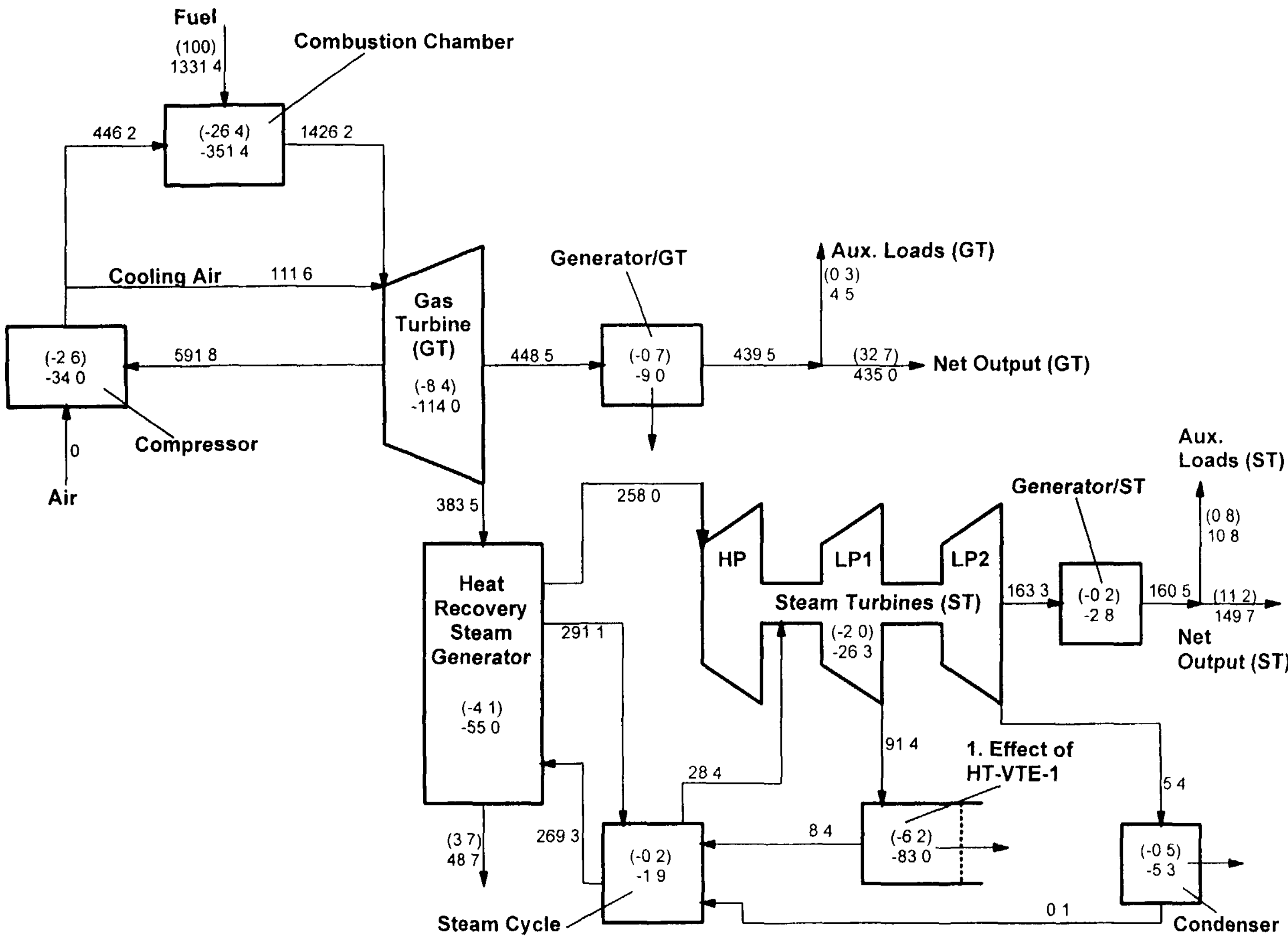

FIG. IV.1. Exergy flow diagram of the combined cycle power plant coupled with the HT-VTE-1 plant.

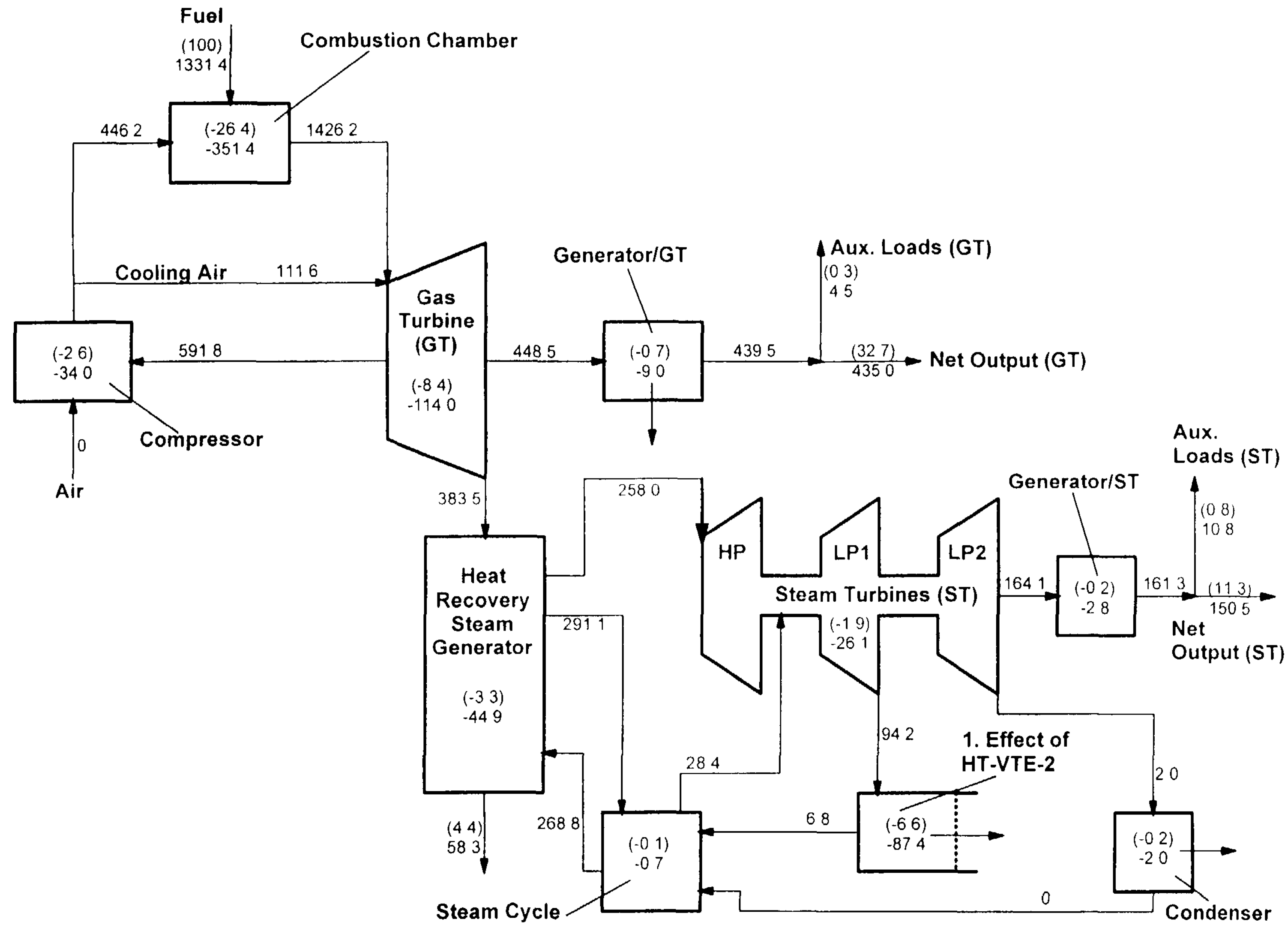

FIG. IV.2. Exergy flow diagram of the combined cycle power plant coupled with the HT-VTE-2 plant.

Annex V

ECONOMIC COST ANALYSIS OF THE PWR POWER PLANT AS ENERGY SOURCE

Coupled with:

1. MSF Plants
2. MED Plants
3. RO Plant

	A	B	C	D	E	F	G	H	I
1									
2	**Exergetic Cost Analysis of Seawater Desalination Plants**								
3	**combined with a Base Load Power Plant for Arabic Countries**								
4									
5	**Power Plant:**	**Nuclear Pressurized Water Reactor (PWR) Power Plant, 600 MW(e) (net)**							
6	**Desalination Plant:**	**MSF Once Through System, 288 000 m^3/day, 4 Units of 72 000 m^3/d**							
7	**Economic Data:**	**All values in US$ (1995), Interest Rate 8 %, Service Year 2005**							
8									
9									
10	Case	MSF-1	MSF-2	MSF-3	MSF-4				
11	Water Plant Capacity, m^3/d	288000	288000	288000	288000				
12	GOR, kg Water/kg Steam	13.5	11.5	9.5	7.5				
13	Maximum Brine Temperature, °C	125	110	98	90				
14									
15	**Base Power Plant Performance Data:**								
16	Total Power Output (gross), MW(e)	634.8	634.8	634.8	634.8				
17	Power Plant Auxiliary Loads, MW(e)	38.1	38.1	38.1	38.1				
18	Total Net Output, MW(e)	596.7	596.7	596.7	596.7				
19	Thermal Power, MW(th)	1870	1870	1870	1870				
20	Net Efficiency, %	31.9	31.9	31.9	31.9				
21	Average Annual Cooling Water Temperature,°C	28.5	28.5	28.5	28.5				
22	Condensing Temperature, °C	40	40	40	40				
23	Condenser Cooling Water Range, °C	8	8	8	8				
24	Condenser Cooling Water Pump Head, bar	1.7	1.7	1.7	1.7				
25	Condenser Cooling Water Pump Efficiency	0.85	0.85	0.85	0.85				
26	Heat Rejection in Condenser, MW(th)	1233	1233	1233	1233				
27	Cooling Water Mass Flow Rate, kg/s	36819	36819	36819	36819				
28	Planned Outage Rate of Power Plant	0.1	0.1	0.1	0.1				
29	Unplanned Outage Rate of Power Plant	0.11	0.11	0.11	0.11				
30	Load Factor of Power Plant	0.801	0.801	0.801	0.801				
31									
32									

	A	B	C	D	E	F	G	H
33	**Dual-Purpose Power Plant Performance Data:**							
34	Condensing Temp of Back-Pressure Turbine, °C	133	118	106	98			
35	Condensing Pressure of Back-Pressure Turbine, bar	3.07	1.94	1.30	0.98			
36	Low-pressure Steam Mass Flow Rate to Water Plant, kg/s	250.8	298.5	363.2	461.4			
37	Heat to Water Plant, MW(th)	541	644	787	1001			
38	Heat Rejection in Low-pressure Turbine Condenser, MW(th)	791	690	557	361			
39	Condenser Cooling Water Mass Flow Rate, kg/s	23628	20609	16645	10784			
40	Power Gross Output, MW(e)	536.3	534.7	524.1	506.5			
41	Less Pump Power in Condenser Compared with Base Power Plant, MW(e)	3.4	4.1	5.0	6.3			
42	Power Net Output, MW(e)	501.6	500.7	491.0	474.8			
43	Exergy of Fuel Allocated to Electricity Generation, MW	1522	1510	1467	1404			
44	Exergy of Fuel Allocated to Steam Production MW	348	360	403	466			
45								
46	**MSF Plant Performance Data:**							
47	Seawater Temperature, °C	30	30	30	30			
48	Seawater Salinity, ppm	45000	45000	45000	45000			
49	Maximum Brine Temperature, °C	125	110	98	90			
50	GOR, kg Water/kg Steam	13.5	11.5	9.5	7.5			
51	Product Water TDS before Post-treatment, ppm	25	25	25	25			
52	Number of Stages per Unit	44	35	27	20			
53	Unit Size, m^3/d	72000	72000	72000	72000			
54	Number of Units	4	4	4	4			
55	Specific Thermal Heat Consumption, kW(th)h/m^3	45.12	53.67	65.61	83.45			
56	Specific Power Use, kW(e)h/m^3	2.60	2.80	2.97	3.23			
57	Seawater Flow, m^3/h	84000	100000	120000	140000			
58	Seawater Mass Flow Rate, kg/s	24033	28611	34333	40056			
59	Seawater Head + Pressure Loss, bar	1.7	1.7	1.7	1.7			
60	Seawater Pump Efficiency	0.85	0.85	0.85	0.85			
61	Seawater Pumping Power, MW(e)	4.86	5.79	6.94	8.10			
62	Intermediate Loop Hot Temperature, °C	129	114	102	94			
63	Intermediate Loop Cold Temperature, °C	122	107	95	87			
64	Intermediate Loop Pressure Loss, bar	1	1	1	1			
65	Intermediate Loop Pump Efficiency	0.85	0.85	0.85	0.85			
66	Intermediate Loop Flow Rate, kg/s	18477	21979	26871	34175			

	A	B	C	D	E	F	G	H
67	Intermediate Loop Pumping Power, MW(e)	2.30	2.73	3.34	4.24			
68	Total MSF Plant Power Use, MW(e)	38.36	42.12	45.92	51.11			
69	MSF Plant Planned Outage Rate	0.03	0.03	0.03	0.03			
70	MSF Plant Unplanned Outage Rate	0.065	0.065	0.065	0.065			
71	MSF Plant Load Factor	0.907	0.907	0.907	0.907			
72	Backup Heat Source Size, MW(th)	541	644	787	1001			
73	Backup Heat Source Planned Outage Rate	0.05	0.05	0.05	0.05			
74	Backup Heat Source Unplanned Outage Rate	0.05	0.05	0.05	0.05			
75	Combined Heat Source Load Factor	0.990	0.990	0.990	0.990			
76	Total Water Plant Load Factor	0.898	0.898	0.898	0.898			
77	Annual Water Production, m^3/y	94389965	94389965	94389965	94389965			
78								
79	**Load Factors:**							
80	Load Factor of Power Plant as Heat Source for Water Plant	0.749	0.749	0.749	0.749			
81	Load Factor of Backup Heat Source	0.149	0.149	0.149	0.149			
82	Load Factor of Power Plant without Heat Coupling	0.052	0.052	0.052	0.052			
83								
84	**Economic Parameters:**							
85	Service Year	2005	2005	2005	2005			
86	Currency Year	1995	1995	1995	1995			
87	Discount Rate, %/y	8	8	8	8			
88	Interest Rate During Construction, %/y	8	8	8	8			
89	Economic Life, Years	30	30	30	30			
90	Fixed Charge Rate, %/y	8.883	8.883	8.883	8.883			
91	Crude Oil Price (Backup Heat Source), $/bbl	17.0	17.0	17.0	17.0			
92	Real Crude Oil Price Escalation, %/y	2.0	2.0	2.0	2.0			
93	Fuel Levelized Factor	1.509	1.509	1.509	1.509			
94								
95	**Cost of Base Power Plant:**							
96	Specific Overnight Cost, $/kW(e)	1874	1874	1874	1874			
97	Overnight Cost, M$	1118.22	1118.22	1118.22	1118.22			
98	Construction Lead Time, months	60	60	60	60			
99	Factor IDC	0.2122	0.2122	0.2122	0.2122			
100	IDC, M$	237.24	237.24	237.24	237.24			

	A	B	C	D	E	F	G	H
101	Total Investment Cost, M$	1355 45	1355 45	1355 45	1355 45			
102	Levelized Annual Capital Cost, M$/y	120 40	120 40	120 40	120 40			
103	Fixed Annual O&M Cost, M$/y	39 42	39 42	39 42	39 42			
104	Variable Annual O&M Cost, $/kW(e)h	0 0005	0 0005	0 0005	0 0005			
105	Total Levelized Annual O&M Cost, M$/y	41 51	41 51	41 51	41 51			
106	Fuel Cost, $/kW(e)h	0 00749	0 00749	0 00749	0 00749			
107	Levelized Annual Fuel Cost, M$/y	31 36	31 36	31 36	31 36			
108	Decommissioning Cost, $/kW(e)h	0 001	0 001	0 001	0 001			
109	Levelized Annual Decommissioning Cost, M$/y	4 19	4 19	4 19	4 19			
110	Total Levelized Annual Cost, M$/y	197 46	197 46	197 46	197 46			
111	Annual Electricity Production, kW(e)h/y	4 187E+09	4 187E+09	4 187E+09	4 19E+09			
112	Levelized Electricity Generation Cost, $/kW(e)h	0 0472	0 0472	0 0472	0 0472			
113								
114	**Overnight Cost of Dual-Purpose Power Plant:**							
115	Overnight Cost (Only Power Production), M$	224 90	224 90	224 90	224 90			
116	Savings Though Common Intake/Outfall, M$	17 37	20 40	24 16	29 13			
117	Total Overnight Cost (Only Power Production), M$	207 53	204 50	200 74	195 77			
118	Overnight Cost (Only Heat Production), M$	0	0	0	0			
119	Common Overnight Cost, M$	893 32	893 32	893 32	893 32			
120								
121	**Exergetic Prorated Electricity Cost:**							
122	Exergetic Prorated Overnight Cost, M$	934 61	925 84	901 54	866 47			
123	IDC, M$	198 28	196 42	191 27	183 83			
124	Total Exergetic Prorated Investment Cost,M$	1132 89	1122 26	1092 81	1050 30			
125	Levelized Annual Capital Cost, M$/y	100 63	99 69	97 07	93 30			
126	Exergetic Prorated Levelized Annual O&M Cost, M$/y	33 81	33 55	32 60	31 20			
127	Exergetic Prorated Levelized Annual Fuel Cost, M$/y	25 90	25 72	25 04	24 05			
128	Exergetic Prorated Levelized Annual Decom Cost, M$/y	3 41	3 38	3 28	3 14			
129	Total Exergetic Prorated Levelized Annual Cost, M$/y	163 76	162 33	157 99	151 69			
130	Annual Electricity Production, kW(e)h/y	3 518E+09	3 511E+09	3 443E+09	3 33E+09			
131	Exergetic Prorated Levelized Electricity Generation Cost, $/kW(e)h	0 0465	0 0462	0 0459	0 0456			
132								
133	**Exergetic Prorated Heat Cost:**							
134	Exergetic Prorated Overnight Cost, M$	166 24	171 98	192 52	222 61			
135	IDC M$	35 27	36 49	40 84	47 23			

	A	B	C	D	E	F	G	H
136	Total Exergetic Prorated Investment Cost, M$	201.51	208 46	233 36	269.84			
137	Levelized Annual Capital Cost, M$/y	17 90	18 52	20 73	23.97			
138	Exergetic Prorated Levelized Annual O&M Cost, M$/y	7 70	7.97	8 92	10.31			
139	Exergetic Prorated Levelized Annual Fuel Cost, M$/y	5.46	5.64	6.32	7 31			
140	Exergetic Prorated Levelized Annual Decom Cost, M$/y	0 78	0.81	0.90	1 04			
141	Exergetic Prorated Total Levelized Annual Heat Cost, M$/y	31 84	32 93	36.87	42 63			
142	Annual Heat to Water Plant, kW(th)h/y	3.552E+09	4.225E+09	5 166E+09	6 57E+09			
143	Exergetic Prorated Levelized Heat Production Cost, $/kW(th)h	0 0090	0.0078	0 0071	0.0065			
144								
145	**Cost MSF Plant:**							
146	Unit Base Cost, M$	105.0	100.0	95.5	92 0			
147	Correction Factor for Number of Units	0.871	0 871	0.871	0 871			
148	MSF Plant Owners Cost Factor	0.05	0.05	0 05	0 05			
149	MSF Plant Contingency Factor	0.10	0 10	0 10	0 10			
150	Base Overnight Cost of MSF Plant, M$	422.30	402 19	384.10	370 02			
151	In/Outfall Base Cost, M$	21.31	24.87	29.20	34.13			
152	Backup Heat Source Unit Cost, $/MW(th)	50000	50000	50000	50000			
153	Backup Heat Source Cost, M$	27.07	32 20	39 37	50 07			
154	Specific Intermediate Loop Unit Cost, $/(m^3/d)	88.44	97.37	109.19	125.83			
155	Intermediate Loop Unit Cost, M$	25.47	28 04	31.45	36.24			
156	MSF Plant Total Overnight Cost, M$	496.16	487.31	484.12	490.46			
157	MSF Plant Lead Time, Months	24	24	24	24			
158	IDC of MSF Plant, M$	39.69	38.98	38.73	39.24			
159	Total MSF Plant Investment, M$	535.85	526.29	522.85	529.70			
160	Levelized Annual MSF Plant Capital Cost, M$/y	47 60	46 75	46.44	47 05			
161	Levelized Annual Steam Cost (Power Plant), M$/y	31.84	32.93	36.87	42 63			
162	Levelized Annual Fuel Cost of Backup Heat Source, M$/y	11.01	13.09	16.01	20 36			
163	Total Levelized Annual Steam Cost, M$/y	42.84	46 03	52.88	62.99			
164	Levelized Annual Electricity Cost, M$/y	14 04	15.32	16 58	18.32			
165	Number of Management Personnel	5	5	5	5			
166	Average Management Salary, $/y	66000	66000	66000	66000			
167	Number of Labour Personnel	35	35	35	35			
168	Average Labour Salary, $/y	29700	29700	29700	29700			
169	Total Annual Personnel Cost, M$/y	1.37	1.37	1.37	1.37			

	A	B	C	D	E	F	G	H
170	Specific O&M Spare Parts Cost, $/m^3	0.035	0.035	0.035	0.035			
171	Specific O&M Chemicals Cost, $/m^3	0.06	0.06	0.06	0.06			
172	Annual MSF Plant O&M Insurance Cost, % of Total InvC/y	0.5	0.5	0.5	0.5			
173	Total Levelized Annual Water Plant O&M Cost, M$/y	12.82	12.77	12.76	12.79			
174	Total Levelized Annual Water Cost, M$/y	117.30	120.87	128.65	141.15			
175	Levelized Potable Water Production Cost, $/m^3	1.243	1.281	1.363	1.495			
176								
177	**Summary:**							
178	**Levelized Potable Water Production Cost, $/m^3**	**1.243**	**1.281**	**1.363**	**1.495**			
179	Total Specific Exergy Consumption, kWh/m^3	41.63	44.82	50.71	59.31			
180	Net Saleable Electricity, MW(e)	458.4	453.2	439.2	417.1			
181	Levelized Electricity Generation Cost, $/kW(e)h	0.0465	0.0462	0.0459	0.0456			
182	Total Levelized Annual Cost, M$/y	267.0	267.9	270.1	274.5			
183	**Levelized Equivalent Electricity Generation Cost, $/kW(e)h**	**0.0830**	**0.0842**	**0.0876**	**0.0938**			

	A	B	C	D	E	F	G	H	I
1									
2	**Exergetic Cost Analysis of Seawater Desalination Plants**								
3	**combined with a Base Load Power Plant for Arabic Countries**								
4									
5	**Power Plant:**	**Nuclear Pressurized Water Reactor (PWR) Power Plant, 600 MW(e) (net)**							
6	**Desalination Plant:**	**MED System, 288 000 m^3/day, 8 Units of 36 000 m^3/day**							
7	**Economic Data:**	**All values in US$ (1995), Interest Rate 8 %, Service Year 2005**							
8									
9									
10	Case	HT-VTE-1	HT-VTE-2	LT-HTME-	LT-HTME-2	LT-HTME-3	LT-HTME-4		
11	Water Plant Capacity, m^3/d	288000	288000	288000	288000	288000	288000		
12	GOR, kg Water/kg Steam	21	17	13 5	11 5	9 5	7 5		
13	Maximum Brine Temperature, °C	120	100	70	65	60	55		
14									
15	**Base Power Plant Performance Data:**								
16	Total Power Output (gross), MW(e)	634 8	634 8	634 8	634 8	634 8	634 8		
17	Power Plant Auxiliary Loads, MW(e)	38 1	38 1	38 1	38 1	38 1	38 1		
18	Total Net Output, MW(e)	596 7	596 7	596 7	596 7	596 7	596 7		
19	Thermal Power, MW(th)	1870	1870	1870	1870	1870	1870		
20	Net Efficiency, %	31 9	31 9	31 9	31 9	31 9	31 9		
21	Average Annual Cooling Water Temperature,°C	28 5	28 5	28 5	28 5	28 5	28 5		
22	Condensing Temperature, °C	40	40	40	40	40	40		
23	Condenser Cooling Water Range, °C	8	8	8	8	8	8		
24	Condenser Cooling Water Pump Head, bar	1 7	1 7	1 7	1 7	1 7	1 7		
25	Condenser Cooling Water Pump Efficiency	0 85	0 85	0 85	0 85	0 85	0 85		
26	Heat Rejection in Condenser, MW(th)	1233	1233	1233	1233	1233	1233		
27	Cooling Water Mass Flow Rate, kg/s	36819	36819	36819	36819	36819	36819		
28	Planned Outage Rate of Power Plant	0 1	0 1	0 1	0 1	0 1	0 1		
29	Unplanned Outage Rate of Power Plant	0 11	0 11	0 11	0 11	0 11	0 11		
30	Load Factor of Power Plant	0 801	0 801	0 801	0 801	0 801	0 801		
31									
32									

	A	B	C	D	E	F	G	H	I
33	**Dual-Purpose Power Plant Performance Data.**								
34	Condensing Temp of Back-Pressure Turbine, °C	129	109	79	74	69	64		
35	Condensing Pressure of Back-Pressure Turbine, bar	2 726	1 440	0 473	0 384	0 310	0 249		
36	Low-pressure Steam Mass Flow Rate to Water Plant, kg/s	161 1	204 7	268 4	317 1	388 0	493 2		
37	Heat to Water Plant, MW(th)	348	441	574	678	825	1050		
38	Heat Rejection in Low-pressure Turbine Condenser, MW(th)	942	855	708	603	459	240		
39	Condenser Cooling Water Mass Flow Rate kg/s	28132	25519	21136	17992	13704	7152		
40	Power Gross Output, MW(e)	571 4	573 0	585 9	585 1	584 1	577 9		
41	Less Pump Power in Cond Comp with Base Power Plant, MW(e)	2 2	2 8	3 6	4 3	5 2	6 6		
42	Power Net Output, MW(e)	535 5	537 7	551 4	551 3	551 2	546 5		
43	Exergy of Fuel Allocated to Electricity Generation, MW	1649	1638	1664	1647	1629	1595		
44	Exergy of Fuel Allocated to Steam Production MW	221	232	206	223	241	275		
45									
46	**MED Plant Performance Data:**								
47	Seawater Temperature, °C	30	30	30	30	30	30		
48	Seawater Salinity, ppm	45000	45000	45000	45000	45000	45000		
49	Maximum Brine Temperature, °C	120	100	70	65	60	55		
50	GOR, kg Water/kg Steam	21 0	17 0	13 5	11 5	9 5	7 5		
51	Product Water TDS before Post-treatment, ppm	25	25	25	25	25	25		
52	Number of Effects per Unit	27+1	22+1	17+1	14+1	11+1	8+1		
53	Unit Size, m^3/d	36000	36000	36000	36000	36000	36000		
54	Number of Units	8	8	8	8	8	8		
55	Specific Thermal Heat Consumption kW(th)h/m^3	29 00	36 72	47 85	56 48	68 73	87 52		
56	Specific Power Use, kW(e)h/m^3	0 90	0 96	1 09	1 25	1 38	1 55		
57	Seawater Flow, m^3/h	68000	80000	104000	136000	160000	192000		
58	Seawater Mass Flow Rate, kg/s	19456	22889	29756	38911	45778	54933		
59	Seawater Head + Pressure Loss bar	1 7	1 7	1 7	1 7	1 7	1 7		
60	Seawater Pump Efficiency	0 85	0 85	0 85	0 85	0 85	0 85		
61	Seawater Pumping Power, MW(e)	3 94	4 63	6 02	7 87	9 26	11 11		
62	Temp Condenser Outlet of Intermediate Loop, °C	127	107	77	72	67	62		
63	Temp Condenser Inlet of Intermediate Loop °C	122	102	72	67	62	57		
64	Intermediate Loop Pressure Loss, bar	1	1	1	1	1	1		
65	Intermediate Loop Pump Efficiency	0 85	0 85	0 85	0 85	0 85	0 85		
66	Intermediate Loop Flow Rate, kg/s	16625	21055	27437	32381	39404	50179		

	A	B	C	D	E	F	G	H	I
67	Intermediate Loop Pumping Power, MW(e)	2.07	2.62	3.41	4.02	4.89	6.23		
68	Total MED Plant Power Use, MW(e)	16.80	18.76	22.51	26.89	30.71	35.94		
69	MED Plant Planned Outage Rate	0.03	0.03	0.03	0.03	0.03	0.03		
70	MED Plant Unplanned Outage Rate	0.065	0.065	0.065	0.065	0.065	0.065		
71	MED Plant Load Factor	0.907	0.907	0.907	0.907	0.907	0.907		
72	Backup Heat Source Size, MW(th)	348	441	574	678	825	1050		
73	Backup Heat Source Planned Outage Rate	0.05	0.05	0.05	0.05	0.05	0.05		
74	Backup Heat Source Unplanned Outage Rate	0.05	0.05	0.05	0.05	0.05	0.05		
75	Combined Heat Source Load Factor	0.990	0.990	0.990	0.990	0.990	0.990		
76	Total Water Plant Load Factor	0.898	0.898	0.898	0.898	0.898	0.898		
77	Annual Water Production, m^3/y	94389965	94389965	94389965	94389965	94389965	94389965		
78									
79	**Load Factors:**								
80	Load Factor of Power Plant as Heat Source for Water Plant	0.749	0.749	0.749	0.749	0.749	0.749		
81	Load Factor of Backup Heat Source	0.149	0.149	0.149	0.149	0.149	0.149		
82	Load Factor of Power Plant without Heat Coupling	0.052	0.052	0.052	0.052	0.052	0.052		
83									
84	**Economic Parameters:**								
85	Service Year	2005	2005	2005	2005	2005	2005		
86	Currency Year	1995	1995	1995	1995	1995	1995		
87	Discount Rate, %/y	8	8	8	8	8	8		
88	Interest Rate During Construction, %/y	8	8	8	8	8	8		
89	Economic Life, Years	30	30	30	30	30	30		
90	Fixed Charge Rate, %/y	8.883	8.883	8.883	8.883	8.883	8.883		
91	Crude Oil Price (Backup Heat Source), $/bbl	17.0	17.0	17.0	17.0	17.0	17.0		
92	Real Crude Oil Price Escalation, %/y	2.0	2.0	2.0	2.0	2.0	2.0		
93	Fuel Levelized Factor	1.509	1.509	1.509	1.509	1.509	1.509		
94									
95	**Cost of Base Power Plant:**								
96	Specific Overnight Cost, $/kW(e)	1874	1874	1874	1874	1874	1874		
97	Overnight Cost, M$	1118.22	1118.22	1118.22	1118.22	1118.22	1118.22		
98	Construction Lead Time, months	60	60	60	60	60	60		
99	Factor IDC	0.2122	0.2122	0.2122	0.2122	0.2122	0.2122		
100	IDC, M$	237.24	237.24	237.24	237.24	237.24	237.24		

	A	B	C	D	E	F	G	H	I
101	Total Investment Cost, M$	1355 45	1355 45	1355 45	1355 45	1355 45	1355 45		
102	Levelized Annual Capital Cost, M$/y	120 40	120 40	120 40	120 40	120 40	120 40		
103	Fixed Annual O&M Cost, M$/y	39 42	39 42	39 42	39 42	39 42	39 42		
104	Variable Annual O&M Cost, $/kW(e)h	0 0005	0 0005	0 0005	0 0005	0 0005	0 0005		
105	Total Levelized Annual O&M Cost, M$/y	41 51	41 51	41 51	41 51	41 51	41 51		
106	Fuel Cost, $/kW(e)h	0 00749	0 00749	0 00749	0 00749	0 00749	0 00749		
107	Levelized Annual Fuel Cost, M$/y	31 36	31 36	31 36	31 36	31 36	31 36		
108	Decommissioning Cost, $/kW(e)h	0 001	0 001	0 001	0 001	0 001	0 001		
109	Levelized Annual Decommissioning Cost, M$/y	4 19	4 19	4 19	4 19	4 19	4 19		
110	Total Levelized Annual Cost, M$/y	197 46	197 46	197 46	197 46	197 46	197 46		
111	Annual Electricity Production, kW(e)h/y	4 187E+09	4 187E+09	4 187E+09	4 187E+09	4 187E+09	4 187E+09		
112	Levelized Electricity Generation Cost, $/kW(e)h	0 0472	0 0472	0 0472	0 0472	0 0472	0 0472		
113									
114	**Overnight Cost of Dual-Purpose Power Plant:**								
115	Overnight Cost (Only Power Production), M$	224 90	224 90	224 90	224 90	224 90	224 90		
116	Savings Through Common Intake/Outfall, M$	13 35	15 91	20 32	24 02	27 73	32 94		
117	Total Overnight Cost (Only Power Production), M$	211 55	208 99	204 58	200 88	197 17	191 96		
118	Overnight Cost (Only Heat Production), M$	0	0	0	0	0	0		
119	Common Overnight Cost, M$	893 32	893 32	893 32	893 32	893 32	893 32		
120									
121	**Exergetic Prorated Electricity Cost:**								
122	Exergetic Prorated Overnight Cost, M$	999 30	991 48	999 49	987 66	975 36	953 91		
123	IDC, M$	212 01	210 35	212 05	209 54	206 93	202 38		
124	Total Exergetic Prorated Investment Cost,M$	1211 30	1201 83	1211 54	1197 20	1182 29	1156 29		
125	Levelized Annual Capital Cost, M$/y	107 60	106 76	107 62	106 34	105 02	102 71		
126	Exergetic Prorated Levelized Annual O&M Cost, M$/y	36 62	36 38	36 96	36 58	36 18	35 43		
127	Exergetic Prorated Levelized Annual Fuel Cost, M$/y	27 89	27 72	28 13	27 86	27 58	27 05		
128	Exergetic Prorated Levelized Annual Decom Cost, M$/y	3 69	3 67	3 73	3 69	3 65	3 57		
129	Total Exergetic Prorated Levelized Annual Cost, M$/y	175 81	174 53	176 43	174 47	172 43	168 76		
130	Annual Electricity Production, kW(e)h/y	3 756E+09	3 772E+09	3 867E+09	3866275337	3865549974	3831305012		
131	Exergetic Prorated Levelized Electricity Generation Cost $/kW(e)h	0 0468	0 0463	0 0456	0 0451	0 0446	0 0440		
132									
133	**Exergetic Prorated Heat Cost:**								
134	Exergetic Prorated Overnight Cost, M$	105 57	110 83	98 41	106 53	115 13	131 37		
135	IDC M$	22 40	23 51	20 88	22 60	24 43	27 87		

	A	B	C	D	E	F	G	H	I
136	Total Exergetic Prorated Investment Cost, M$	127 97	134 34	119 29	129 13	139 55	159 24		
137	Levelized Annual Capital Cost, M$/y	11 37	11 93	10 60	11 47	12 40	14 14		
138	Exergetic Prorated Levelized Annual O&M Cost, M$/y	4 89	5 13	4 56	4 93	5 33	6 08		
139	Exergetic Prorated Levelized Annual Fuel Cost, M$/y	3 47	3 64	3 23	3 50	3 78	4 31		
140	Exergetic Prorated Levelized Annual Decom Cost, M$/y	0 49	0 52	0 46	0 50	0 54	0 62		
141	Exergetic Prorated Total Levelized Annual Heat Cost, M$/y	20 22	21 22	18 85	20 40	22 05	25 16		
142	Annual Heat to Water Plant, kW(th)h/y	2 283E+09	2 891E+09	3 767E+09	4446415771	5410775303	6890265570		
143	Exergetic Prorated Levelized Heat Production Cost, $/kW(th)h	0 0089	0 0073	0 0050	0 0046	0 0041	0 0037		
144									
145	**Cost MED Plant:**								
146	Unit Base Cost, M$	45 3	42 3	39 8	36 4	34 6	32 6		
147	Correction Factor for Number of Units	0 812	0 812	0 812	0 812	0 812	0 812		
148	MED Plant Owners Cost Factor	0 05	0 05	0 05	0 05	0 05	0 05		
149	MED Plant Contingency Factor	0 10	0 10	0 10	0 10	0 10	0 10		
150	Base Overnight Cost of MED Plant, M$	339 99	317 47	298 71	273 19	259 68	244 67		
151	In/Outfall Base Cost, M$	17 27	20 10	25 34	30 92	35 39	41 35		
152	Backup Heat Source Unit Cost, $/MW(th)	50000	50000	50000	50000	50000	50000		
153	Backup Heat Source Cost, M$	17 40	22 03	28 71	33 89	41 24	52 51		
154	Specific Intermediate Loop Unit Cost, $/(m^3/d)	67 84	77 01	88 44	97 37	109 19	125 83		
155	Intermediate Loop Unit Cost, M$	19 54	22 18	25 47	28 04	31 45	36 24		
156	MED Plant Total Overnight Cost, M$	394 19	381 79	378 23	366 04	367 75	374 77		
157	MED Plant Lead Time, Months	24	24	24	24	24	24		
158	IDC of MED Plant, M$	31 54	30 54	30 26	29 28	29 42	29 98		
159	Total MED Plant Investment, M$	425 73	412 33	408 49	395 32	397 17	404 76		
160	Levelized Annual MED Plant Capital Cost, M$/y	37 82	36 63	36 28	35 12	35 28	35 95		
161	Levelized Annual Steam Cost (Power Plant), M$/y	20 22	21 22	18 85	20 40	22 05	25 16		
162	Levelized Annual Fuel Cost of Backup Heat Source, M$/y	7 08	8 96	11 68	13 78	16 77	21 35		
163	Total Levelized Annual Steam Cost, M$/y	27 29	30 18	30 52	34 18	38 82	46 51		
164	Levelized Annual Electricity Cost, M$/y	6 19	6 83	8 08	9 55	10 78	12 45		
165	Number of Management Personnel	5	5	5	5	5	5		
166	Average Management Salary, $/y	66000	66000	66000	66000	66000	66000		
167	Number of Labour Personnel	35	35	35	35	35	35		
168	Average Labour Salary, $/y	29700	29700	29700	29700	29700	29700		
169	Total Annual Personnel Cost, M$/y	1 37	1 37	1 37	1 37	1 37	1 37		

	A	B	C	D	E	F	G	H	I
170	Specific O&M Spare Parts Cost, \$/m^3	0.03	0.03	0.03	0.03	0.03	0.03		
171	Specific O&M Chemicals Cost, \$/m^3	0.06	0.06	0.06	0.06	0.06	0.06		
172	Annual MLD Plant O&M Insurance Cost % of Total Invest Cost/y	0.5	0.5	0.5	0.5	0.5	0.5		
173	Total Levelized Annual Water Plant O&M Cost, M\$/y	11.84	11.77	11.76	11.69	11.70	11.74		
174	Total Levelized Annual Water Cost, M\$/y	83.13	85.41	86.64	90.54	96.58	106.66		
175	Levelized Potable Water Production Cost, \$/m^3	0.881	0.905	0.918	0.959	1.023	1.130		
176									
177	**Summary:**								
178	**Levelized Potable Water Production Cost, \$/m^3**	**0.881**	**0.905**	**0.918**	**0.959**	**1.023**	**1.130**		
179	Total Specific Exergy Consumption, kWh/m^3	24.65	27.19	28.19	31.89	36.11	42.88		
180	Net Saleable Electricity, MW(e)	516.5	516.5	525.9	520.9	516.5	505.7		
181	Levelized Electricity Generation Cost, \$/kW(e)h	0.0468	0.0463	0.0456	0.0451	0.0446	0.0440		
182	Total Levelized Annual Cost, M\$/y	252.8	253.1	255.0	255.5	258.2	263.0		
183	**Levelized Equivalent Electricity Generation Cost, \$/kW(e)h**	**0.0697**	**0.0698**	**0.0691**	**0.0699**	**0.0713**	**0.0741**		

	A	B	C	D	E	F	G	H	I
1									
2	**Exergetic Cost Analysis of Seawater Desalination Plants**								
3	**combined with a Base Load Power Plant for Arabic Countries**								
4									
5	**Power Plant:**	**Nuclear Pressurized Water Reactor (PWR) Power Plant, 600 MW(e) (net)**							
6	**Desalination Plant:**	**RO System, 288 000 m^3/d, 12 Units of 24 000 m^3/d**							
7	**Economic Data:**	**All values in US$ (1995), Interest Rate 8 %, Service Year 2005**							
8									
9									
10	Case	RO							
11	Water Plant Capacity, m^3/d	288000							
12	RO Membrane Type	Hollow Fiber							
13	Seawater Total Dissolved Solids, ppm	45000							
14									
15	**Base Power Plant Performance Data:**								
16	Total Power Output (gross), MW(e)	634 8							
17	Power Plant Auxiliary Loads, MW(e)	38 1							
18	Total Net Output, MW(e)	596 7							
19	Thermal Power, MW(th)	1870							
20	Net Efficiency, %	31 9							
21	Average Annual Cooling Water Temperature,°C	28 5							
22	Condensing Temperature, °C	40							
23	Condenser Cooling Water Range, °C	8							
24	Condenser Cooling Water Pump Head, bar	1 7							
25	Condenser Cooling Water Pump Efficiency	0 85							
26	Heat Rejection in Condenser, MW(th)	1233							
27	Cooling Water Mass Flow Rate, kg/s	36819							
28	Planned Outage Rate of Power Plant	0 1							
29	Unplanned Outage Rate of Power Plant	0 11							
30	Load Factor of Power Plant	0 801							
31									

	A	B	C	D	E	F	G	H	I
32	**RO Plant Performance Data:**								
33	Seawater Temperature, °C	27							
34	Seawater Salinity, ppm	45000							
35	Recovery Ratio	0.35							
36	Feed Pressure, bar	72							
37	Unit Size, m^3/d	24000							
38	Number of Units	12							
39	Number of Permeators per Unit	795							
40	Seawater Flow, m^3/h	34286							
41	Seawater Mass Flow Rate, kg/s	9810							
42	Seawater Head + Pressure Loss, bar	1.7							
43	Seawater Pump Efficiency	0.85							
44	Booster Pump Head, bar	3.3							
45	Booster Pump Efficiency	0.85							
46	High Head Pump Pressure Rise, bar	71.0							
47	High Head Pump Efficiency	0.85							
48	Hydraulic Coupling Efficiency	0.966							
49	Energy Recovery Efficiency	0.85							
50	Other Specific Power Use, $kW(e)/(m^3/d)$	0.0408							
51	Seawater Pumping Power, MW(e)	1.98							
52	Booster Pump Power, MW(e)	3.85							
53	High Head Pump Power, MW(e)	85.78							
54	Energy Recovery, MW(e)	-37.36							
55	Other Power, MW(e)	11.75							
56	Total RO Plant Power Use, MW(e)	66.01							
57	RO Plant Planned Outage Rate	0.032							
58	RO Plant Unplanned Outage Rate	0.06							
59	RO Plant Load Factor	0.910							
60	Annual Water Production, m^3/y	95650790							
61	Specific Power Consumption, $kW(e)h/m^3$	5.50							
62									
63	**Economic Parameters:**								
64	Service Year	2005							
65	Currency Year	1995							

	A	B	C	D	E	F	G	H	I
66	Discount Rate, %/y	8							
67	Interest Rate During Construction, %/y	8							
68	Economic Life, Years	30							
69	Fixed Charge Rate, %/y	8 883							
70									
71	**Cost of Base Power Plant:**								
72	Specific Overnight Cost, $/kW(e)	1874							
73	Overnight Cost, M$	1118 22							
74	Construction Lead Time, months	60							
75	Factor IDC	0 2122							
76	IDC, M$	237 24							
77	Total Investment Cost, M$	1355 45							
78	Levelized Annual Capital Cost, M$/y	120 40							
79	Fixed Annual O&M Cost, M$/y	39 42							
80	Variable Annual O&M Cost, $/kW(e)h	0 0005							
81	Total Levelized Annual O&M Cost, M$/y	41 51							
82	Fuel Cost, $/kW(e)h	0 00749							
83	Levelized Annual Fuel Cost, M$/y	31 36							
84	Decommissioning Cost, $/kW(e)h	0 001							
85	Levelized Annual Decommissioning Cost, M$/y	4 19							
86	Total Levelized Annual Cost, M$/y	197 46							
87	Annual Electricity Production, kW(e)h/y	4 187E+09							
88	Levelized Electricity Generation Cost, $/kW(e)h	0 0472							
89									
90	**Cost of Contiguous Power Plant:**								
91	Savings Through Common Intake/Outfall, M$	5 22							
92	Total Overnight Cost of Contig Power Plant, M$	1113 00							
93	Levelized Annual Capital Cost, M$/y	119 84							
94	Total Levelized Annual Cost, M$/y	196 90							
95	Levelized Electricity Generation Cost, $/kW(e)h	0 0470							
96									
97	**Cost RO Plant:**								
98	Specific Unit Base Cost, $/(m^3/d)	1000							
99	Membrane Price per Permeator, $	4000							

	A	B	C	D	E	F	G	H	I
100	Membrane Cost, M$	38.16							
101	Correction Factor for Number of Units	0.780							
102	RO Plant Owners Cost Factor	0.05							
103	RO Plant Contingency Factor	0.1							
104	Base Overnight Cost of RO Plant, M$	259.45							
105	In/Outfall Base Cost, M$	8.82							
106	RO Plant Total Overnight Cost, M$	268.27							
107	RO Plant Lead Time, Months	24							
108	IDC of RO Plant, M$	21.46							
109	Total RO Plant Investment, M$	289.73							
110	Levelized Annual RO Plant Capital Cost, M$/y	25.74							
111	Levelized Annual Electricity Cost, M$/y	24.74							
112	Number of Management Personnel	5							
113	Average Management Salary, $/y	66000							
114	Number of Labour Personnel	35							
115	Average Labour Salary, $/y	29700							
116	Total Annual Personnel Cost, M$/y	1.37							
117	Specific O&M Spare Parts Cost, $/m^3	0.04							
118	Specific O&M Chemicals Cost, $/m^3	0.06							
119	Annual Membrane Replacement Rate, %/y	15							
120	Annual Membrane Replacement Cost, M$/y	5.72							
121	Annual RO Plant O&M Insurance Cost, % of Total Invest. Cost/y	0.5							
122	Levelized Annual RO Plant Total O&M Cost, M$/y	18.00							
123	Total Levelized Annual Water Cost, M$/y	68.48							
124	Levelized Potable Water Production Cost, $/m^3	0.716							
125									
126	**Summary:**								
127	**Levelized Potable Water Production Cost, $/m^3**	**0.716**							
128	Total Specific Exergy Consumption, kWh/m^3	17.24							
129	Net Saleable Electricity, MW(e)	521.7							
130	Levelized Electricity Generation Cost, $/kW(e)h	0.0470							
131	Total Levelized Annual Cost, M$/y	240.6							
132	**Levelized Equivalent Electricity Generation Cost, $/kW(e)**	**0.0657**							

Annex VI

ECONOMIC COST ANALYSIS OF THE COMBINED CYCLE POWER PLANT AS ENERGY SOURCE

Coupled with:

1. MED Plants
2. RO Plant

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Exergetic Cost Analysis of Seawater Desalination Plants combined with a Base Load Power Plant for Arabic Countries

	A	B	C
5	**Power Plant:**	**Combined Cycle Power Plant Fueled with Natural Gas, 640 MW(e) (net)**	
6	**Desalination Plant:**	**MED System, 288 000 m^3/day, 8 Units of 36 000 m^3/day**	
7	**Economic Data:**	**All values in US$ (1995), Interest Rate 8 %, Service Year 2005**	

	A	B	C	D	E	F	G	H	I
10	Case	HT-VTE-1	HT-VTE-2						
11	Water Plant Capacity, m^3/d	288000	288000						
12	GOR, kg Water/kg Steam	21	17						
13	Maximum Brine Temperature, °C	120	100						
14									
15	**Base Power Plant Performance Data:**								
16	Gas Turbine Output (gross), MW(e)	3 x 146.5	3 x 146.5						
17	Steam Turbine Output (gross), MW(e)	215.4	215.4						
18	Total Power Output (gross), MW(e)	654.9	654.9						
19	Power Plant Auxiliary Loads, MW(e)	15.2	15.2						
20	Total Net Output, MW(e)	639.7	639.7						
21	Thermal Power, MW(th)	1286 7	1286 7						
22	Net Efficiency, %	49.7	49.7						
23	Average Annual Ambient Temperature, °C	28.5	28.5						
24	Exhaust Gas Temperature (behind Turbine), °C	541	541						
25	Exhaust Gas Temperature (behind HRSG), °C	106.5	106.5						
26	Gross Efficiency of Gas Turbines, %	34 2	34.2						
27	Average Annual Cooling Water Temperature,°C	28.5	28 5						
28	Condensing Temperature, °C	40	40						
29	Condenser Cooling Water Range, °C	8	8						
30	Condenser Cooling Water Pump Head, bar	1.7	1.7						
31	Condenser Cooling Water Pump Efficiency	0 85	0.85						

	A	B	C	D	E	F	G	H	I
32	Heat Rejection in Condenser, MW(th)	493	493						
33	Cooling Water Mass Flow Rate, kg/s	14722	14722						
34	Planned Outage Rate of Power Plant	0.1	0.1						
35	Unplanned Outage Rate of Power Plant	0.11	0.11						
36	Load Factor of Power Plant	0.801	0.801						
37									
38	**Dual-Purpose Power Plant Performance Data:**								
39	Condensing Temp. of Back-Pressure Turbine, °C	123	103						
40	Condensing Pressure of Back-Pressure Turbine, bar	2.268	1.172						
41	Low-pressure Steam Mass Flow Rate, kg/s	161.4	203.8						
42	Heat to Water Plant, MW(th)	348	441						
43	Heat Rejection in Low-pressure Turbine Condenser, MW(th)	144	53						
44	Condenser Cooling Water Mass Flow Rate, kg/s	4305	1590						
45	Power Gross Output, MW(e)	600.0	600.8						
46	Less Pump Power in Cond. Compa. with Base Power Plant, MW(e)	2.2	2.8						
47	Power Net Output, MW(e)	587.0	588.4						
48	Exergy of Fuel Allocated to Electricity Generation, MW	1184	1178						
49	Exergy of Fuel Allocated to Production, MW	147	153						
50									
51	**MED Plant Performance Data:**								
52	Seawater Temperature, °C	30	30						
53	Seawater Salinity, ppm	45000	45000						
54	Maximum Brine Temperature, °C	120	100						
55	GOR, kg Water/kg Steam	21.0	17.0						
56	Product Water TDS before Post-treatment, ppm	25	25						
57	Number of Effects per Unit	28	23						
58	Unit Size, m^3/d	36000	36000						
59	Number of Units	8	8						
60	Specific Thermal Heat Consumption, $kW(th)h/m^3$	29.00	36.72						
61	Specific Power Use, $kW(e)h/m^3$	0.90	0.96						
62	Seawater Flow, m^3/h	68000	80000						
63	Seawater Mass Flow Rate, kg/s	19456	22889						
64	Seawater Head + Pressure Loss, bar	1.7	1.7						
65	Seawater Pump Efficiency	0.85	0.85						

	A	B	C	D	E	F	G	H	I
66	Seawater Pumping Power, MW(e)	3 94	4 63						
67	Total MED Plant Power Use, MW(e)	14 74	16.15						
68	MED Plant Planned Outage Rate	0 03	0 03						
69	MED Plant Unplanned Outage Rate	0 065	0 065						
70	MED Plant Load Factor	0.907	0 907						
71	Backup Heat Source Size, MW(th)	348	441						
72	Backup Heat Source Planned Outage Rate	0 05	0 05						
73	Backup Heat Source Unplanned Outage Rate	0 05	0 05						
74	Combined Heat Source Load Factor	0.990	0.990						
75	Total Water Plant Load Factor	0 898	0.898						
76	Annual Water Production, m^3/y	94389965	94389965						
77									
78	**Load Factors:**								
79	Load Factor of Power Plant as Heat Source for Water Plant	0 749	0 749						
80	Load Factor of Backup Heat Source	0 149	0 149						
81	Load Factor of Power Plant without Heat Coupling	0.052	0.052						
82									
83	**Economic Parameters:**								
84	Service Year	2005	2005						
85	Currency Year	1995	1995						
86	Discount Rate, %/y	8	8						
87	Interest Rate During Construction, %/y	8	8						
88	Economic Life, Years	30	30						
89	Fixed Charge Rate, %/y	8.883	8 883						
90	Crude Oil Price, $/bbl	17 0	17.0						
91	Real Crude Oil Price Escalation, %/y	2.0	2 0						
92	Fuel Levelized Factor	1.509	1.509						
93									
94	**Cost of Base Power Plant:**								
95	Specific Overnight Cost, $/kW(e)	600	600						
96	Overnight Cost, M$	383.84	383.84						
97	Construction Lead Time, months	36	36						
98	Factor IDC	0.1224	0.1224						
99	IDC, M$	46.97	46 97						

	A	B	C	D	E	F	G	H	I
100	Total Investment Cost, M$	430 81	430 81						
101	Levelized Annual Capital Cost, M$/y	38 27	38 27						
102	Total Annual O&M Cost mills/kW(e)h	5 50	5 50						
103	Fixed Annual O&M Cost, M$/y	7 41	7 41						
104	Variable Annual O&M Cost $/kW(e)h	0 00385	0 00385						
105	Total Levelized Annual O&M Cost, M$/y	24 69	24 69						
106	Fuel Cost, $/kW(e)h	0 0208	0 0208						
107	Levelized Annual Fuel Cost M$/y	140 65	140 65						
108	Total Levelized Annual Cost, M$/y	203 61	203 61						
109	Annual Electricity Production, kW(e)h/y	4 489E+09	4 489E+09						
110	Levelized Electricity Generation Cost, $/kW(e)h	0 0454	0 0454						
111									
112	**Overnight Cost of Dual Purpose Power Plant:**								
113	Overnight Cost (Only Power Production), M$	115 15	115 15						
114	Savings Through Common Intake/Outfall, M$	21 17	24 92						
115	Total Overnight Cost (Only Power Production), M$	93 98	90 23						
116	Overnight Cost (Only Heat Production), M$	0	0						
117	Common Overnight Cost, M$	268 69	268 69						
118									
119	**Exergetic Prorated Electricity Cost:**								
120	Exergetic Prorated Overnight Cost, M$	332 94	328 00						
121	IDC, M$	40 74	40 14						
122	Total Exergetic Prorated Investment Cost M$	373 68	368 14						
123	Levelized Annual Capital Cost, M$/y	33 19	32 70						
124	Exergetic Prorated Levelized Annual O&M Cost M$/y	22 08	21 98						
125	Exergetic Prorated Levelized Annual Fuel Cost, M$/y	126 10	125 52						
126	Total Exergetic Prorated Levelized Annual Cost M$/y	181 38	180 20						
127	Annual Electricity Production, kW(e)h/y	4 118E+09	4 127E+09						
128	Exergetic Prorated Levelized Electricity Cost $/kW(e)h	0 0440	0 0437						
129									
130	**Exergetic Prorated Heat Cost:**								
131	Exergetic Prorated Overnight Cost, M$	29 73	30 92						
132	IDC M$	3 64	3 78						
133	Total Exergetic Prorated Investment Cost, M$	33 36	34 70						

	A	B	C	D	E	F	G	H	I
134	Levelized Annual Capital Cost, M$/y	2.96	3.08						
135	Exergetic Prorated Levelized Annual O&M Cost, M$/y	2.61	2.49						
136	Exergetic Prorated Levelized Annual Fuel Cost, M$/y	14.55	15.13						
137	Exergetic Prorated Levelized Total Annual Heat Cost, M$/y	20.12	20.71						
138	Annual Heat to Water Plant, kW(th)h/y	2.283E+09	2.891E+09						
139	Exergetic Prorated Levelized Heat Cost, $/kW(th)h	0.0088	0.0072						
140									
141	**Cost MED Plant:**								
142	Unit Base Cost, M$	46.5	43.6						
143	Correction Factor for Number of Units	0.812	0.812						
144	MED Plant Owner's Cost Factor	0.05	0.05						
145	MED Plant Contingency Factor	0.10	0.10						
146	Base Overnight Cost of MED Plant, M$	349.0	327.2						
147	In/Outfall Base Cost, M$	26.6	30.7						
148	Backup Heat Source Unit Cost, $/MW(th)	50000	50000						
149	Backup Heat Source Cost, M$	17.4	22.0						
150	MED Plant Total Overnight Cost, M$	393.0	379.9						
151	MED Plant Lead Time, Months	24	24						
152	IDC of MED Plant, M$	31.44	30.40						
153	Total MED Plant Investment, M$	424.39	410.34						
154	Levelized Annual MED Plant Capital Cost, M$/y	37.70	36.45						
155	Levelized Annual Steam Cost (Power Plant), M$/y	20.12	20.71						
156	Levelized Annual Fuel Cost of Backup Heat Source, M$/y	7.08	8.96						
157	Total Levelized Annual Steam Cost, M$/y	27.20	29.67						
158	Levelized Annual Electricity Cost, M$/y	5.10	5.55						
159	Number of Management Personnel	5	5						
160	Average Management Salary, $/y	66000	66000						
161	Number of Labour Personnel	35	35						
162	Average Labour Salary, $/y	29700	29700						
163	Total Annual Personnel Cost, M$/y	1.37	1.37						
164	Specific O&M Spare Parts Cost, $/m^3	0.03	0.03						
165	Specific O&M Chemicals Cost, $/m^3	0.06	0.06						
166	Annual MED Plant O&M Insurance Cost, % of Total Invest. Cost/y	0.5	0.5						
167	Total Levelized Annual Water Plant O&M Cost, M$/y	11.83	11.76						

	A	B	C	D	E	F	G	H	I
168	Total Levelized Annual Water Cost, M$/y	81 83	83 43						
169	Levelized Potable Water Production Cost, $/m^3	0 867	0 884						
170									
171	**Summary:**								
172	**Levelized Potable Water Production Cost, $/m^3**	**0.867**	**0.884**						
173	Total Specific Exergy Consumption, kWh/m^3	17 69	19 65						
174	Net Saleable Electricity, MW(e)	570 4	570 1						
175	Levelized Electricity Generation Cost, $/kW(e)h	0 0440	0 0437						
176	Total Levelized Annual Cost, M$/y	258 1	258 1						
177	**Levelized Equivalent Electricity Generation, $/kW(e)h**	**0.0645**	**0.0645**						

	A	B	C	D	E	F	G	H	I	J
1										
2	**Exergetic Cost Analysis of Seawater Desalination Plants**									
3	**combined with a Base Load Power Plant for Arabic Countries**									
4										
5	**Power Plant:**	**Combined Cycle Power Plant Fueled with Natural Gas, 640 MW(e) (net)**								
6	**Desalination Plant:**	**RO System, 288 000 m^3/d, 12 Units of 24 000 m^3/d**								
7	**Economic Data:**	**All values in US$ (1995), Interest Rate 8 %, Service Year 2005**								
8										
9										
10	Case	RO								
11	Water Plant Capacity, m^3/d	288000								
12	RO Membrane Type	Hollow Fiber								
13	Seawater Total Dissolved Solids, ppm	45000								
14										
15	**Base Power Plant Performance Data:**									
16	Gas Turbine Output (gross), MW(e)	3 x 146.5								
17	Steam Turbine Output (gross), MW(e)	215.4								
18	Total Power Output (gross), MW(e)	654.9								
19	Power Plant Auxiliary Loads, MW(e)	15 2								
20	Total Net Output, MW(e)	639 7								
21	Thermal Power, MW(th)	1286 7								
22	Net Efficiency, %	49 7								
23	Average Annual Ambient Temperature, °C	28.5								
24	Exhaust Gas Temperature (behind Turbine), °C	541								
25	Exhaust Gas Temperature (behind HRSG), °C	106.5								
26	Gross Efficiency of Gasturbines, %	34.2								
27	Average Annual Cooling Water Temperature,°C	28 5								
28	Condensing Temperature, °C	40								
29	Condenser Cooling Water Range, °C	8								
30	Condenser Cooling Water Pump Head, bar	1.7								
31	Condenser Cooling Water Pump Efficiency	0 85								
32	Heat Rejection in Condenser, MW(th)	493								

	A	B	C	D	E	F	G	H	I	J
33	Cooling Water Mass Flow Rate, kg/s	14722								
34	Planned Outage Rate of Power Plant	0.1								
35	Unplanned Outage Rate of Power Plant	0.11								
36	Load Factor of Power Plant	0.801								
37										
38	**RO Plant Performance Data:**									
39	Seawater Temperature, °C	27								
40	Seawater Salinity, ppm	45000								
41	Recovery Ratio	0.35								
42	Feed Pressure, bar	72								
43	Unit Size, m^3/d	24000								
44	Number of Units	12								
45	Number of Permeators per Unit	795								
46	Seawater Flow, m^3/h	34286								
47	Seawater Mass Flow Rate, kg/s	9810								
48	Seawater Head + Pressure Loss, bar	1.7								
49	Seawater Pump Efficiency	0.85								
50	Booster Pump Head, bar	3.3								
51	Booster Pump Efficiency	0.85								
52	High Head Pump Pressure Rise, bar	71.0								
53	High Head Pump Efficiency	0.85								
54	Hydraulic Coupling Efficiency	0.966								
55	Energy Recovery Efficiency	0.85								
56	Other Specific Power Use, kW(e)/(m^3/d)	0.0408								
57	Seawater Pumping Power, MW(e)	1.98								
58	Booster Pump Power, MW(e)	3.85								
59	High Head Pump Power, MW(e)	85.78								
60	Energy Recovery, MW(e)	37.36								
61	Other Power, MW(e)	11.75								
62	Total RO Plant Power Use, MW(e)	66.01								
63	RO Plant Planned Outage Rate	0.032								
64	RO Plant Unplanned Outage Rate	0.06								
65	RO Plant Load Factor	0.910								
66	Annual Water Production, m^3/y	95650790								

	A	B	C	D	E	F	G	H	I	J
67	Specific Power Consumption, kW(e)h/m^3	5 50								
68										
69	**Economic Parameters:**									
70	Service Year	2005								
71	Currency Year	1995								
72	Discount Rate, %/y	8								
73	Interest Rate During Construction, %/y	8								
74	Economic Life, Years	30								
75	Fixed Charge Rate, %/y	8 883								
76	Crude Oil Price, $/bbl	17 0								
77	Real Crude Oil Price Escalation, %/y	2 0								
78	Fuel Levelized Factor	1 509								
79										
80	**Cost of Base Power Plant:**									
81	Specific Overnight Cost, $/kW(e)	600								
82	Overnight Cost, M$	383 82								
83	Construction Lead Time, months	36								
84	Factor IDC	0 1224								
85	IDC, M$	46 97								
86	Total Investment Cost, M$	430 79								
87	Levelized Annual Capital Cost, M$/y	38 27								
88	Total Annual O&M Cost, mills/kW(e)h	5 50								
89	Fixed Annual O&M Cost, M$/y	7 41								
90	Variable Annual O&M Cost, M$/y	17 28								
91	Total Levelized Annual O&M Cost, M$/y	24 69								
92	Fuel Cost, $/kW(e)h	0 0208								
93	Levelized Annual Fuel Cost, M$/y	140 65								
94	Total Levelized Annual Cost, M$/y	203 61								
95	Annual Electricity Production, kW(e)h/y	4 489E+09								
96	Levelized Electricity Generation Cost, $/kW(e)h	0 0454								
97										
98	**Cost of Contiguous Power Plant:**									
99	Savings Through Common Intake/Outfall, M$	7 34								
100	Total Overnight Cost of Contig Power Plant, M$	376 48								

	A	B	C	D	E	F	G	H	I	J
101	Levelized Annual Capital Cost, M$/y	37 53								
102	Total Levelized Annual Cost, M$/y	202 87								
103	Levelized Electricity Generation Cost, $/kW(e)h	0 0452								
104										
105	**Cost RO Plant:**									
106	Specific Unit Base Cost, $/(m^3/d)	1000								
107	Membrane Price per Permeator, $	4000								
108	Membrane Cost, M$	38 16								
109	Correction Factor for Number of Units	0 780								
110	RO Plant Owners Cost Factor	0 05								
111	RO Plant Contingency Factor	0 1								
112	Base Overnight of RO Plant, M$	259 45								
113	In/Outfall Base Cost, M$	13 13								
114	RO Plant Total Overnight Cost, M$	272 58								
115	RO Plant Lead Time, Months	24								
116	IDC of RO Plant, M$	21 81								
117	Total RO Plant Investment, M$	294 39								
118	Levelized Annual RO Plant Capital Cost, M$/y	26 15								
119	Levelized Annual Electricity Cost, M$/y	23 78								
120	Number of Management Personnel	5								
121	Average Management Salary, $/y	66000								
122	Number of Labour Personnel	35								
123	Average Labour Salary, $/y	29700								
124	Annual Total Personnel Cost, M$/y	1 37								
125	Specific O&M Spare Parts Cost, $/m^3	0 04								
126	Specific O&M Chemicals Cost, $/m^3	0 06								
127	Annual Membrane Replacement Rate, %/y	15								
128	Annual Membrane Replacement Cost, M$/y	5 72								
129	Annual RO Plant O&M Insurance Cost % of Total Invest Cost/y	0 5								
130	Levelized Annual Water Plant Total O&M Cost, M$/y	18 02								
131	Total Levelized Annual Water Cost, M$/y	67 95								
132	Levelized Potable Water Production Cost, $/m^3	0 710								
133										
134										

	A	B	C	D	E	F	G	H	I	J
135	Summary:									
136	**Levelized Potable Water Production Cost, $/m³**	**0.710**								
137	Total Specific Exergy Consumption, kWh/m³	11 45								
138	Net Saleable Electricity, MW(e)	564 7								
139	Levelized Electricity Generation Cost, $/kW(e)h	0 0452								
140	Total Levelized Annual Cost, M$/y	247 0								
141	**Levelized Equivalent Electricity Generation Cost, $/kW(e)h**	**0.0623**								

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ABBREVIATIONS

BPE	boiling point elevation
BR	brine recycle
CC	combined cycle
GCC	Gulf Cooperation Council
GOR	gain-output ratio
HRSG	heat recovery steam generator
HT-VTE	high temperature vertical tube evaporator
IDC	interest during construction
LT-HTME	low-temperature horizontal-tube multi-effect distillation
MED	multiple effect distillation
MSF	multi-stage flash distillation
NEA	Nuclear Energy Agency (of the OECD)
NEL	non-equilibrium loss
O&M	operation & maintenance
OECD	Organisation for Economic Co-operation and Development
OT	Once through
PWR	pressurized water reactor
RO	reverse osmosis
TDS	total dissolved solids
VTE	vertical tube evaporator
WHO	World Health Organization

DEFINITIONS

The following definitions are used throughout the report:

Desalination plant	Installations comprizing all buildings, structures, systems and components necessary to produce potable water from saline water, with an input of energy, in the form of heat and/or electricity.
Dual purpose plant	Nuclear or fossil fuelled power plant with a product output of both heat (steam or hot water) and electricity.
Power plant	Installation comprizing all buildings, structures, systems and components necesarry to produce electricity. Note: In this report, the term "dual purpose plant" is also covered by the term "power plant".
Single purpose plant	Plant with a single output (product), e.g. potable water only, or electricity only.
Integrated plant	Power plant jointly located with desalination plant, sharing seawater intake/outlet structures.

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Consultants Meeting

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