SEAWATER DESALINATION TECHNOLOGY

JAPAN INTERNATIONAL COOPERATION AGENCY

March 1979

FOREWORD

It was in 1969 that "Seawater Desalination and By-Products Recovery" was taken up as a subject of the National R & D Program by the Agency of Industrial Science and Technology of the Ministry of International Trade and Industry. The project was brought to successful completion nine years after that at a huge national expense amounting to 7 billion yen and using tremendous manpower. During this period, the Japanese technology in the development of a seawater desalination plant using the multistage flash process, which was the main theme of development under the project, gained a world-wide reputation.

At that time the oil producing countries of the world began to place huge orders for large scale seawater desalination plants to solve their long-standing problem of water shortages and the Japanese industries were beginning to gain a large share of the volume of orders received from the Middle East countries.

Under such circumstances, it was decided to make public, as much as possible, various information and data obtained through studies and researches under this National R & D Program so as to be of some assistance to those who have an interest in this project, and to publish a book to disseminate information, mainly about a new plant of the high flow rate multistage flash process together with the peripheral technical data.

The main part of the book is a compilation of the findings of development research carried out by the National Chemical Laboratory for Industry over a long period of time, which originated from research on the use of seawater, under the instruction of Mr. Shunsei Inoue, Director General of the Agency of Industrial Science and Technology. This was in 1951 and was a special subject of development. The book also refers to the technological assessment of the operation of a plant with respect to the recovery of by-products and problems of raw seawater and plant waste water, as well as basic matters of plant design.

The National Chemical Laboratory for Industry played a central role in the accomplishment of development work under this program by engaging in basic research, collection of data and coordination of research work among the parties concerned. But many private organizations took part in the development of the plant facilities. Development of the main part of the plant was commissioned to Sasakura Engineering Co., Ltd., Kajima Corporation, Ishikawajima-Harima Heavy Industries Co., Ltd., Hitachi, Ltd., Japan Gasoline Co., Ltd. and Mitsubishi Heavy Industries, Ltd., and studies on the recovery of by-products were commissioned to Asahi Glass Co., Ltd. In addition, many other private organizations cooperated in the development of the plant. Among them are Tokyo Kyuei Co., Ltd. which tackled the problem of the intake and discharge of seawater, Ebara Manufacturing Co., Ltd. which conducted a series of pump tests and a number of material suppliers who did research on the material of heating tubes. Also, valuable suggestions were made by persons of learning and experience who took part in the working group.

On the occasion of publishing this book, I would like to express my gratitude to more than 200 of the staff of government agencies, academic institutions and private organizations for their participation in this project.

The book allocates considerable space for the description of a reverse osmosis plant as well as a multistage flash plant. The subject of a reverse osmosis plant was selected for development under the project as technology worthy of special note for the future development of seawater desalination. From technology originated in the United States of America, Japan's own technology is being developed with the special aid of the Ministry of International Trade and Industry. I shall be very happy if the information contained in this book will be of some help to the reader.

Finally, I wish to dedicate this book to the late Mr. Kenji Tanaka, division chief and Mr. Takashi Uchida, section chief, of the National Chemical Laboratory for Industry, who were taken ill in the course of their research work and succumbed without seeing the conclusion of this project after their long services with the laboratory.

March 1978

Seiichi Ishizaka

Director, National Chemical
Laboratory for Industry

- A Note to The Reader -

"Seawater Desalination and By-Products Recovery", the theme of National R & D Program undertaken by the Agency of Industrial Science and Technology, came to successful conclusion at the end of fiscal 1977 after 9 years of intensive research and development work. It is quite natural that the parties concerned have planned to compile all the findings of the research into a book. By the initiative of Mr. Minoru Sayama, senior officer for the development program (desalination) of the Agency of Industrial Science and Technology, and through the courtesy and positive support of Mr. Tsuneo Momota, managing director of the Japan Industrial Technology Promotion Association, the problem of clerical requirements was solved and the plan to publish a book was put on the right track.

The first meeting of the editorial staff was held in March 1977. The book was constructed with the main emphasis placed on the introduction of the findings of studies and research carried out under this National program. Readers of the book were assumed to be plant makers, staff of departments of planning or water resources development in local public bodies, teaching staff and students of universities and colleges, as well as staff presently engaged in research and development on the subject of seawater desalination. It was felt, therefore, that the book should be edited in such a manner that the reader could easily comprehend the present state of technology applied to seawater desalination at the first reading. Of course, the book must contain highly technical information which is useful to specialists in this field and for this reason, attempts were made to include every piece of information available.

Contributors are leading engineers and researchers presently engaged actively in business or research related to seawater desalination. Many of them have been involved directly or indirectly in this project. I am confident, therefore, that the content of the book is the highest level of its kind in the world.

On the occasion of the publication of this book, I would like to extend my thanks to Mr. Toki Honda and Ms. Masako Imoto of the Japan Industrial Technology Promotion Association for their support, especially in putting in order and arranging manuscripts.

Jun Kato
Chairman of Editorial Committee

CONTENTS

I. Water Demand-Supply Situation and Seawater Desalination

1.1. Present State of Development of Water Resources

Water is one of the resources that are indispensable to the existence of mankind. In early days, human beings formed communities along rivers or around lakes and made use of natural water for drinking, washing, bathing, farming and industry. With the increase of population and development of industry, however, there was a shortage of water in the natural state and mankind began to develop water resources by constructing reservoirs, dams in rivers and canals to conduct water from remote areas. This was according to their capital and technology.

During the past century, the rapid progress of civilization has brought about a sharp increase of population, a marked rise in standards of living and a remarkable progress in industry, which, in turn, has contributed to a sharp increase in demand for irrigation, city and industrial water. Even today, the demand for water is increasing steadily.

Under these conditions, it has become increasingly difficult to secure water resources for the future by the conventional means of constructing reservoirs and large dams or regulating the natural flow of rivers, which are now practised to increase the amount of water available for annual consumption. As a result, attention has been directed in recent years to recycling of city waste water or industrial water, or seawater desalination as a means of securing water resources.

The Water Resources Bureau of the National Land Agency made public the result of a survey on the supply limit of water resources in March 1976.

According to the survey, the reserves of water resources in Japan during a period of 18 years from 1956 through 1973 was as shown in Table 1.1-1. A further breakdown of the reserves of water resources into the mountain region, flatland region and coastal region is shown in Table 1.1-2.

The area of mountain, flatland and coastal regions is 237,180 km^2, 86,772 km^2 and 53,532 km^2, respectively, accounting for 62.8%, 23.0% and 14.2%, respectively, of the national total. Of this, the coastal region includes small river basins, each having an area of less than 59 km^2, located close to the coastline where most rainfall drains into the sea. For this reason, much cannot be expected of the coastal region as an object of water resources development.

In Japan, nearly all natural river flow was already utilized for irrigation and other purposes prior to the enactment of the River Law in 1896.

The water rights gained in those days is called the habitual water right under the present River Law and is given the same status as the new water rights accorded under the law.

After the enactment of the River Law in 1896, therefore, the intake of river water for city water, industrial water and irrigation water is authorized only when a new water source has been developed.

On the other hand, the use of ground water has never been controlled because of the lack of a strict law like the River Law and ground water has been used in large quantities as industrial and city water because of its good quality, ease of use and low cost. The ground water now accounts for approximately 34 percent of industrial water supply (fiscal 1975) and 20 percent of city water (fiscal 1974).

The use of large quantities of ground water is causing serious problems, including ground subsidence and the lowering of the ground water level in many parts of the country.

Under the Water for Industrial Use Act (legislated in 1956), the Koto and Johoku districts of Tokyo, the south district of Saitama, part of Nagoya, the area around Yokkaichi, Osaka City and some major cities in Osaka Prefecture where ground subsidence is rapidly progressing, are specified as areas of compulsory shift of water supply sources of industry from ground water to surface water through the construction of new industrial water supply systems. The shift is now in progress in these areas. Nevertheless, ground subsidence or contamination of ground water caused by excessive pumping of ground water is a nationwide problem and the control on the use of ground water is being tightened through the establishment of new regulations in many prefectures. In the future, it will be extremely difficult to seek a water supply source in ground water and it will be necessary to shift from the existing ground water to surface water in many parts of the country. Under these conditions, the development of new water resources is becoming increasingly important.

Table 1.1-3 shows the number of dams and reservoirs constructed to date, which gives first hand information on the state of development of water resources in Japan.

From the table, the available storage capacity for irrigation, city and industrial water is estimated at 13,000 million cubic meters.

The flow of a river changes almost every hour and every day, and the mean daily discharge plotted for 365 days in terms of the volume of flow is called a current curve.

The usefulness of a dam is that it guarantees water supply throughout the year by storing rain water in spring and autumn for use in the drought seasons of summer and winter.

In other words, an available storage capacity of 13,000 million cubic meters generates a water supply capacity about twice as great as the available storage capacity through the year in the spring → summer, autumn → winter pattern. However, 100 percent use of available water is not possible in practice and an available storage capacity of 13,000 million cubic meters can be considered to equal a supply capacity of approximately 20,000 million cubic meters.

Each river shows a different current curve and reservoirs located closer to the origin of the river have a higher supply efficiency.

Rivers like to Tone, which has many dams constructed on it, show a relatively flat current curve and newly constructed dams have a low supply efficiency. However, the supply capacity denoted by AB-C in Fig. 1.1-1 creates a new water use denoted by BCDE.

In other words, those dams having a total available storage capacity of approximately 13,000 million cubic meters can be considered to have made it possible to use an additional 50,000 million cubic meters of water through the year.

Since the drought water level of natural river flow (Note 1) shown in Fig. 1.1-1 is considered to be 1 m^3/sec. per 100 km^2 (31,540,000 m^3/year) on the average in mountain regions, the total drought water level in an area of 237,180 km^2 of the mountain regions of Japan shown in Table 1.1-2 amounts to approximately 70,000 million cubic meters.

This drought water level is not utilized completely, but is reserved in part for a river maintaining flow. The river maintaining flow is a flow which is essential to the maintenance of inherent roles of a river such as transportation by ship, life support of fish, supply of underflow water and purification of river water.

At any rate, of the 236,900 million cubic meters of water resources in mountain regions shown in Table 1.1-2, which still remain as a possibility for future development, only 100,000 million cubic meters is discharged into the sea without being utilized during high water seasons such as the typhoon season and rainy season. The remaining 120,000 million to 130,000 million cubic meters of water is being utilized in one form or another.

However, the population and industry of Japan are concentrated in specific districts as shown in Fig. 1.1-4 and a district which is abundant in water resources does not always induce population and industry. While 100,000 million cubic meters of natural water is wasted in the nation annually, as mentioned previously, there is a serious water shortage in some parts of the country.

Water resources are highly natural and are different from other resources in nature, as the transportation of water over a long distance is extremely difficult.

In the past, water was readily available in Japan if some investment was made for development. The controversially high rate of city water service accounts for only 0.4 percent of the total living cost on the average and the share of industrial water in the cost of manufactured goods is less than 1 percent for many commodities. It may safely be said, therefore, that the availability of water has never been considered as an important factor for the concentration of population or industrial location in Japan.

In the future, however, there will be an increasing number of districts which have funds but are unable to find room for the development of water resources, or face the problem of lacking the necessary volume of water resources for development.

At present, a number of water resources development projects are being undertaken by the State or local governments, but after 1985 or 1995 it may even be difficult to plan the development of water resources in such regions as the waterfront area of the Kanto region, Hanshin district, the Inland Sea district, and the Kita-Kyushu region.

In August 1973, the Ministry of Construction made public its water demand-supply forecast for 1986 in the "Report of the Second Survey on Wide-area Water Use". The forecast shows that while the new water demand dependent on the supply of river water during a period of 15 years from 1970 to 1987 is 40,160 million cubic meters, the amount of water available through the development of river water resources is estimated at 46,500 million cubic meters and that there will be a water shortage of 60 million cubic meters in Tokiwa-Koriyama, 1,970 million cubic meters in South Kanto, 1,210 million cubic meters in Keihanshin, 220 million cubic meters in Bingo, 160 million cubic meters in Takamatsu, 60 million cubic meters in Toyo, 70 million cubic meters in Matsuyama and 450 million cubic meters in North Kyushu.

The forecast tells indirectly that it will be impossible to meet the water demand after 1985 unless large scale water conveyance systems are provided to transport water from Tohoku or Hokuriku to Kanto or Kinki, from Sanin to Sanyo and from south Kyushu to north Kyushu.

In the meantime, the National Land Agency released an interim report, "Findings of the Second National Water Demand Survey and Its Review", in February 1976.

The water demand-supply forecast contained in the survey is shown in Table 1.1-5.

While the water demand forecast will be discussed in the next section, it is worthwhile to note the amount of available water resources predicted by the National Land Agency.

Against an annual increase of 18,560 million to 20,890 million cubic meters of water demand during a period of 10 years from 1976 to 1985, the annual increase of water supply for the same period is estimated at 14,590 million to 20,530 million cubic meters which is the sum of 20,530 million cubic meters under the existing projects and 5,940 million cubic meters from projects planned. Accordingly, the attainment of the water balance shown in Table 1.1-5 will be extremely difficult when the fact is taken into consideration that the development of water resources generally takes more than 10 years from the begining to the completion.

Since some of the existing projects may not be completed by 1985, the water balance calculated on the basis of the estimated minimum increase of 14,590 million cubic meters in water supply will be as shown in Table 1.1-6.

From this point of view, the National Land Agency gave a warning in newspapers published in February 1976 that there will be an annual shortage of 4,000 million to 6,000 million cubic meters in water supply in 1985 even with the low economic growth rate and decentralization of industries taken into consideration.

It may be appropriate at this point to touch on the present state of dams constructed as a means of development of water resources in Japan.

Most of the dams constructed for the development of water resources in Japan are multi-purpose dams with a combined function of either flood control, power generation, irrigation, industrial use or city water supply, and dams constructed exclusively for the supply of water are very small in size and very few in number.

According to the 1976 data published by the River Bureau of the Ministry of Construction, the multi-purpose dams intended for both city water service and industrial water, under construction or planned, are as shown in Table 1.1-7 and the top 15 dams in order of high construction cost classified by prefecture are as shown in Table 1.1-8.

Figures shown in Table 1.1-8 were calculated by the writer on the basis of data furnished by the Ministry of Construction. The table is indicative of the effort being made for the development of water resources in areas confronted with a problem of water shortages. The raw water cost per cubic meter seems to be higher than 100 yen for many of the dams now planned.

1.2. Present and Future Water Demand

1.2.1. Water Demand in 1985

To know the future trend of water demand, it is necessary to predict the future economic scale, industrial structure, city size and population size and distribution. However, these subjects will not be discussed here and will be left to the Third Comprehensive National Land Development Program formulated by the National Land Agency.

In February 1976, the Water Resources Bureau of the National Land Agency published "An Interim Report of the Second National Survey on Water Demand", in which it forecast the water demand for 1985. Table 1.2-1 is a forecast of city water demand for 1985.

The forecast was made under the following conditions.

(1) The total population will be 124 million in 1985.

(2) The average annual growth rate of industrial shipment during the 1970 - 1985 period will be at a low level of 5.5 percent.

(3) The share of each region in industrial shipment will be determined depending on the measures which call for drastic decentralization of industries to reduce the share of the Pacific belt-zone in industrial shipment by 10 percent during a period of 10 years.

(4) The unit consumption of domestic water in 1985 will be 399 ℓ/day on the national average, which is considerably lower than the estimated figure up to 1970.

(5) The diffusion rate of water service in 1985 will be 97 percent on the national average.

(6) The recycling rate of industrial water will be increased from 62 percent in 1973 to 70 percent in 1985.

Even in this conservative forecast, a generation of annual water demand for 12,230 million to 14,300 million cubic meters (33.5 million to 39.5 million m^3/day) is estimated during the 1976 - 1985 period as shown in Table 1.2-2 (The discrepancy between these figures and those in Table 1.1-5 is due to the difference between the intake base figure and supply base figure).

Besides the figures in Table 2.1-2, a shift from ground water to surface water in the amount of 1,120 million m^3/year, 1,140 million m^3/year and 220 million m^3/year is being planned for the waterfront area of the Kanto region, Tokai region and the waterfront area of the Kinki region, respectively, to solve various problems resulting from excess pumping of ground water.

A comparison of figures in Table 1.2-2 and the amount of available water supply is shown in Table 1.1-5.

1.2.2. Demand for Service Water

The demand for service water is increasing steadily as the result of the expansion of waterworks, concentration of population in urban areas, a rise in the standard of living, modernization of cities and the growth of the economy.

On the national average, the increase rate of demand for service water dropped sharply as the result of a drastic increase of water rates and the stagnation of the national economy in the wake of the oil crisis in the autumn of 1973, but has turned upward again in recent years.

Table 1.2-3 shows changes in the amount of water supply after 1960 and Fig. 1.2-1 shows changes in the monthly supply of water in recent years.

1.2.3. Demand for Industrial Water

Changes in the consumption of industrial water during the period from 1958 to recent years by source are shown in Table 1.2-4.

According to the table, consumption of industrial water (fresh water) in 1975 amounted to approximately 122 x 10^6 m^3/day, of which make-up water amounted to 40 x 10^6 m^3/day. The remaining 82 x 10^6 m^3/day is recycled water. The rate of recycled water increased sharply from 20 percent in 1958 to 67 percent in 1975 and the Ministry of International Trade and Industry aims at an increase of the rate of recycled water to 70 percent in 1985.

The amount of make-up fresh water used per unit of industrial shipment changed from 103 m^3/day/100 million yen (at prices of 1975) in 1970 to 69 m^3/day/100 million yen in 1975 owing to the aforementioned improvement of the water recycling rate and the increase of the ratio of non water-oriented industries such as machinery and electronics industries of high added values to the water-oriented industries such as chemical, iron and steel, pulp and paper and textile industries in the industrial shipment. This figure is expected to decrease further to 41 - 43 m^3/day/ 100 million yen in 1985 as shown in Table 1.2-1.

However, a new water demand amounting to 2,910 million - 4,980 million m^3/year is expected to be generated during a 10 year period from 1976 to 1985. Since the use of ground water will be restricted and most of the requirement will have to depend on the development of new water resources, supply of industrial water may become a serious problem in the future. Incidentally, the amount of water supplied by industrial water works provided during a period of 17 years from 1956, the year of the first use of industrial water, to 1975 is 11.95 million m^3/day (4,360 million m^3/year) as shown in Table 1.2-4.

1.2.4. Demand for Irrigation Water

It is extremely difficult to determine the actual state of the use of irrigation water due to the fact that the amount of intake fluctuates greatly from hour to hour, that countless intake facilities are in use and that water meters are seldom used in these facilities.

According to a wide-area survey on water use conducted by the Ministry of Construction, the irrigation water led to a tract of paddy field is consumed through evaporation from the free water surface or transpiration and percolation to the ground. The portion percolated into the ground and the remainder on the free water surface is used again by paddy fields downstream. The amount of irrigation water which is reused in this fashion in 1970 is estimated at 5,240 m^3/year.

The "Interim Report of the Second National Survey of Water Demand" published by the National Land Agency estimates the amount of irrigation water, which was not reused, in all prefectures in 1970 at 109,000 million cubic meters. Of this, 84 percent is for irrigation of paddy field, 15 percent for miscellaneous use and less than 1 percent for irrigation of uplands. By season, the irrigation season accounts for 84 percent and the non-irrigation season accounts for 16 percent of total consumption.

As for future demand for irrigation water, the area of paddy fields is expected to decrease by 350,000 ha and the area of upland requiring irrigation is expected to increase by 460,000 ha during the period from 1970 to 1985. The annual consumption of irrigation water is expected to increase from 109,000 million cubic meters in 1970 to 113,800 million cubic meters in 1985.

1.3. Water Demand-Supply Situation and Seawater Desalination

As discussed so far, demand for water is expected to increase indefinitely in proportion to the growth of the economy. In Japan, which is relatively abundant in water resources, the potential water resources for development have become scarce year by year and some regions find it difficult to secure the necessary water resources.

During the "Tokyo Olympic Drought" of 1964, the restriction on the use of water in Tokyo was tightened gradually from the initial 25 percent to 35.5 percent and finally to 47.8 percent during the period from July to August and urban functions and industrial activities in the region were reduced by half for about one month. Besides Tokyo, a major drought occurred in Nagasaki in 1967 and in the Keihan, Setouchi and Sanin regions in 1973, and the impact of drought has become more serious with the sophistication of the mechanism of communities.
The 1976 drought in Europe was protracted and especially in England, industrial activities came to a complete stop, giving a hard blow to the English economy.

A major drought, such as becomes the object of public concern, has occurred every 10 years or so in some regions of Japan and restrictions on the supply of water by 10 to 25 percent have frequently been enforced to date.

In addition, the gap between the demand and supply of water is widening, as the development of water resources makes very slow progress owing to various problems such as the requirement for compensation to residents of dam sites and environmental protection. The wide-area water utilization, which has been advocated by the Ministry of Construction for the past 10 years, and decentralization of industries encouraged by the Cabinet have made little progress to date.

Some experts on the development of water resources have even suggested that the object drought year for the development of water resources should be changed to the third worst drought year (the third worst year in terms of least river discharge during a 10 year period - The second worst year was adopted in the past).

Adoption of the third worst year of drought as the object of development of water resources means that the nation must endure the hardest year and second hardest year of drought in 10 years.

So far, major droughts have become a social problem but have never led to the loss of life. During the major droughts of the past, people managed to overcome the difficult with patience, though economic activities came to a standstill to some extent. This may be the reason why the adoption of the third worst year of drought has been suggested.

If the object of technical advances is to relieve mankind from the burden of hardship and to build up an affluent society, a new field of technical advancement may be found in seawater desalination.

In arid zones like the Middle Near East, seawater desalination is indispensable for the modernization of urban areas and development of industries and there is no question about it. In Japan, however, it may be difficult to justify seawater desalination as a general source of water supply.

In Japan, seawater desalination has been practised as a means of securing water supply in isolated islands or for the operation of thermal or nuclear power plants which had to be located in such areas as the tips of peninsulas or headlands, where little water resources are available, because of the difficulty in securing land space in inland areas. This tendency is expected to continue in the future.

As a new trend of recent years, seawater desalination is being considered in areas like Okinawa and Kita-Kyushu where there is no room for the development of water resources and the matter has been given serious attention in waterfront areas of the Kanto and Kinki regions where the water demand-supply gap is widening owing to the difficulty in developing new water resources.

At any rate, seawater desalination has come closer to the development of river water resources with respect to the development cost, but it should be regarded as the last resort in the development of water resources at this stage because it still consumes valuable energy resources, though very small in quantity, even when its multi-purpose of the utilization of waste heat, incineration of waste materials, generation of thermal and nuclear power and saving of some energy is taken into consideration.

It is important, however, to give seawater desalination a position as an integral part of the total system for development of water resources by linking it systematically with the development of river water resources. One example of such total systems may be the aforementioned third worst drought year theory. Supposing a seawater desalination plant is constructed in a certain area, the object year of development of river water resources for that area may be the third or fourth hardest year of drought and the total investment for the development of water resources may be held to a minimum.

Supposing the object year of development of water resources has to be the third hardest year of drought, it will be advantageous from an economic point of view to provide a seawater desalination plant corresponding to the estimated value of hardship to the community and economic loss caused by water shortages in the first and second hardest years of drought. The Third Comprehensive National Land Development Program estimates the amount of industrial shipment in the Tokyo region in 1985 at approximately 60,000 billion yen (at prices of 1975). If the loss caused by a major drought, which may occur once in every 10 years, is assumed to be 2 percent, the total loss in industrial shipment in the Tokyo region will be 1,200 billion yen. Water Science, Japan estimates that the construction of a seawater desalination plant having a capacity of 100 x 10^6 m^3/day in the Tokyo Metropolitan Area will adequately cope with the situation when the progress of development of water resources in the Kanto region in 1985 is taken into consideration. The construction cost of a seawater desalination plant having a capacity of

100×10^6 m^3/day is estimated at around 120 billion yen at 1975 prices. If a plant having a capacity of 100×10^6 m^3/day can be constructed at a cost of 120 billion yen, the annual operating expense or fixed cost, including depreciation charges, interest, maintemance cost and personnel expense, will be less than 20 billion yen. If calculation of social and economic losses in monetary terms is possible in this manner, construction of a seawater desalination plant is justified as an insurance for the maintenance of society.

In any case, the above-mentioned two questions are still premature for general acceptance at present but should be given due consideration as a future possibility.

The water situation in 1985, only seven or eight years hence, is hotly debated with a sense of crisis at present, but the situation in 1995 or 2005 has been discussed very little. When the time comes to take up the water situation after 1985, however, the question of seawater desalination will be tackled squarely. By then, the technology of seawater desalination will have been improved considerably and plant design will have changed to a more energy-saving and lower-cost type.

At present, the cost of seawater desalination with a "high flow rate long tube multistage flash evaporation" plant having a capacity of 100,000 m^3/day, which has been developed under this national project of the Agency of Industrial Science and Technology of the Ministry of International Trade and Industry, is approximately 170 yen/m^3.

This cost, 170 yen/m^3, is equivalent to 60 to 70 percent of the cost of seawater desalination with existing technology. This is the cost at the plant gate at an annual operating rate of 80 percent. A breakdown of this cost is shown in Table 1.3-2.

It must be noted here that the cost of raw water in the case of development of river water resources shown in Table 1.1-8 is not the cost of water supplied from a dam but is the value obtained from division of annual operating expense of the dam by the volume of water which is made available from the dam through the year. In other words, the cost of raw water in the case of development of river water resources includes the cost of river water during the high water season.

Judging from the number of storage days of dams shown in Table 1.1-8, the period of discharge from the dam during the low water season is about 100 days.

If fresh water from the sea is combined with river water obtained during the high water season as in the case of a dam, the result will be as follows. Assuming that a seawater desalination plant having a capacity of 100,000 m^3/day is operated for 100 days during the low water season and river water is used during the remaining season, the annual operating expense - (only the fixed cost is considered during the high water season) - will be;

Low water season 100 days x 100,000 m^3/day x 170.7 yen/m^3
= 1,707.4 yen

High water season 265 days x 100,000 m^3/day x 53.74 yen/m^3
= 1,424 yen

The total of the above is 3,131.5 million yen. If this total is divided by the volume of annual water supply, 100,000 m^3/365 days = 36,500,000 m^3, the result is 85.8 yen/m^3.

While the development of river water resources requires installation of intake facilities down the river, water mains, a water treatment plant, a service reservoir and water pipes, a seawater desalination requires no intake facilities or water treatment facilities.

It can be said, therefore, that seawater desalination has come to almost the same level as the development of river water resources as far as the cost is concerned.

Table 1.1-1 Reserve of water resources in Japan

Constituent / Basic year	rainfall mm/year	Precipitation $10^8m^3/y$	Reserve mm/year	Annual reserve $10^8m^3/y$.	Remark
Drought year	1,480	5,587	883	3,333	§1
Normal year	1,788	6,749	1,191	4,494	§2
Bountiful year	2,131	8,044	1,534	5,791	§3

Note: Annual reserve under column 3 is the balance from annual precipitation less annual loss, giving an effective precipitation in a sense.

Remark: §1 Represents the history's second drought,
§2 Average of consecutive years,
§3 Represents the history's second bountiful.

Table 1.1-2 Availability of water resources in Japan
(Unit: hundred million m^3/year
for endowment, % for share)

Basic year / Region	Drought year		Normal year		Bountiful year	
	Availability	Share	Availability	Share	Abailability	Share
Mountain	2,369	71.1	3,108	69.2	3,944	68.1
Flatland	581	17.4	819	18.2	1,077	18.6
Coastal	383	11.5	567	12.6	770	13.3
Total	3,333	100.0	4,494	100.0	5,791	100.0

Table 1.1-3 Dams and water storage in Japan

	No. of dams	Effective water stg.	Remarks
Multi-purpose dam	339	12, 991 $10^6 m^3$	
Agricultural dam	2, 384	12, 284	Inclusive of duplication with multiple-purpose dam.
Urban water dam	94	376	
Industrial water dam	7	86	
Generating dam	532	12, 018	Inclusive of duplications with the above four kinds of dams.
Flood control dam	118	491	

Source: Creating a Water Resources Model, March 1977, by by Water Resources Bureau of National Land Agency.

Note: Projecting dams since 1975 include in number of dams.

Table 1.1-4 Population, industrial outputs and water resources in the Pacific coast of Japan

District	Area	Population 1973 year	Industrial outputs 1975 year	Availability of water resources	
				Total amt.	amt, per capita.
	(Km²)	(10^4)	(10^8yen)	($10^8 m^3$/yr)	(m^3/year)
Kanto	32,279	3,097	366,310	215,5	651
Tokai	22,819	1,071	176,377	322,0	3,006
Kinki	33,026	1,958	207,258	342,5	1,749
Sanyo	21,621	588	88,621	169,0	2,874
Northern Kyushu	24,430	942	49,220	196,4	2,084
Sub-total	134,175	7,657	887,786	1,245,4	1,624
Nationwide total	377,400	10,820	1,040,105	3,333,1	3,080
National share	35.5 %	70.8 %	85.4 %	37.4 %	–

Table 1.1-5 Water demand projection for 1976 - 1985

(unit: hundred million m^3/year) In-take amount basis

		Demand				Supply
		Demand increment for '76 - '85	Tentative in-take in '75	Underground water availability for	Total New addition of river water for '76 - '85	Supply available for '76-'85
Hokkaido		13.7~15.2	0	△0.9	12.8~14.3	3.7~15.1
Tohoku		21.3~22.7	0.3	△0.6	21.0~22.4	11.4~23.8
Kanto	Inlan	18.1~19.0	—	△0.4	—	—
	Seaside	16.2~20.6	—	11.2	—	—
	Total	34.3~39.6	14.6	10.8	59.7~65.0	48.4~57.6
Tokai		11.8~17.7	3.5	11.4	26.7~32.6	29.9~38.2
Hokuriku		3.4~4.8	1.0	1.7	6.1~7.5	5.8~8.5
Kinki	Inland	6.1~6.5	—	△1.4	—	—
	Seaside	6.7~9.0	—	2.2	—	—
	Total	12.8~15.5	11.5	0.8	25.1~27.8	18.5~25.4
hugoku	San-in	2.6~2.1	—	△0.3	—	—
	San-yo	4.5~7.6	—	0.7	—	—
	Total	7.1~9.7	1.5	0.4	9.0~11.6	10.9~11.6
Shikoku		6.1~6.8	0	0.4	6.5~7.2	7.0~7.3
Kyushu	North	11.8~13.7	0.3	0.1	12.2~14.1	—
	South	7.8~7.1	0	△1.3	6.5~5.8	—
	Total	19.6~20.8	0.3	△1.2	18.7~19.9	10.3~17.8
Okinawa						
Nationwide total		130.1~152.8	32.7	22.8	185.6~208.3	145.9~205.3

Note: 1. The lower limits under column of supplies include on-going projects, while the upper include future undertakings.

2. Figures under column of demand increment is on in-take basis.

3. Left-side figures under column of demand are computed by the National land Development Ministry based on estimates by regional offices, while right-side figures computed on the basis of report by Industry Rationalization Commission in July 1975.

Table 1.1-6 Demand increment/supply balance for '76 - '85 on regional basis

(unit: hundred million ton, Δ shortage)

	Demand	Supply (A)	Shortage
Hokkaido	12.8～14.3	3.7	△9.1～△10.6
Tohoku	21.0～22.4	11.4	△9.6～△11.0
Kanto	59.7～65.0	48.4	△11.3～△16.6
Tokai	26.7～32.6	29.9	3.2～△2.7
Hokuriku	6.1～7.5	5.8	△0.3～△1.7
Kinki	25.1～27.8	18.5	△6.6～△9.3
Chugoku	9.0～11.6	10.9	1.9～△0.7
Shikoku	6.5～7.2	7.0	0.5～△0.2
Kyushu	18.7～19.9	10.3	△8.7～△9.6
National total	185.6～208.3	145.9	△39.7～△62.4

Table 1.1-7 Hydraulic dam under construction and planning

	Dams by Ministry of Construction and by Water Resource Dev. Corporation	Auxiliary dams by regional offices
No. of dams	39	63
Total effective water storage	5,666 $10^6 m^3$	616,4 $10^6 m^3$
Water availability	2,543 $10^6 m^3$	273,6 $10^6 m^3$
Water development used	28,645,815 m^3/d	5,517,580 m^3/d

Source: Report on integrated development project on rivers (September 1976) by Ministry of Construction, River Bureau.

Note: The available water includes service water and sewage and excludes agricultural water.

Table 1.1-8 Water cost at site of the most expensive 15 dams in terms of water development cost

Ranking	Location	Capacity ($10^6 m^3$)	Investment (10^6 yen)	Availability (m^3/day)	Storage (days)	Water cost (yen/m^3)
1	Kagawa	0.14	1,318	1,000	140	87.12
2	Shimane	0.095	315	1,100	86.4	78.75
3	Okinawa	13.70	27,084	120,000	114.0	61.80
4	Nagasaki	0.72	1,620	8,000	90.0	55.48
5	Hyogo	7.2	17,833	92,000	78.3	53.11
6	Kumamoto	0.48	1,035	8,000	60.0	35.45
7	Osaka	7.47	11,177	84,000	86.4	35.44
8	Tochigi	21.40	25,056	216,000	99.9	31.78
9	Saga	0.04	167	1,500	26.7	30.36
10	Fukuoka	0.73	2,649	25,000	29.2	29.01
11	Saga	0.226	290	2,800	80.7	28.43
12	Saga	2.40	4,352	42,163	57.3	28.20
13	Ishikawa	0.48	1,347	14,000	34.3	26.36
14	Nagasaki	3.96	5,663	60,000	66.0	25.86
15	Ibaragi	1.20	3,233	34,560	34.7	25.64

Note: 1. The ranking is in the order of water cost at site.
2. Storage days are computed by water storage capacity/availability.
3. Water cost at site is computed by investment cost × 10%/availability × 365 days.

Table 1.2-1 Demand increment projection of urban water in 1985

District		Service water				Industrial water			Uraban water		
		P	D	Basic rate	De-mand	Output	F.W.S.	Demand	Demand	U.W.D	R.W.D.
		10^6	(%)	ℓ/m/d	10^8			$10^8 m^3/yr$.	$10^8 m^3/yr$.	$10^8 m^3/yr$	$10^8 m^3/yr$
Hokkaido		632	96.4	366	8.1	5.7~ 6.1	78~78	16.1~17.5	24.3~25.7	2.6	23.3~24.8
Tohoku		1,228	93.4	345	14.4	11.3~12.3	59~58	24.8~26.0	39.2~40.5	10.3	31.1~32.5
Kanto	Inland	1,019	95.7	366	13.0	14.6~13.8	35~39	18.7~19.5	31.7~32.5	11.4	21.8~22.7
	Seaside	3,065	99.1	447	49.5	36.7~32.8	17~23	23.9~27.8	73.4~77.4	6.6	71.8~76.2
	Total	4,084	98.3	427	62.5	51.3~46.6	23~28	42.6~47.3	105.1~109.9	18.0	93.6~98.9
Tokai		1,437	97.0	385	19.6	25.5~25.4	50~56	46.9~52.1	66.5~71.7	23.4	46.3~51.9
Hokuriku		313	97.5	400	4.5	4.4~ 4.6	70~78	11.3~12.7	15.8~17.1	7.7	8.7~10.1
Kinki	Inland	565	97.5	390	7.9	5.8~ 7.1	39~34	8.4~ 8.8	16.2~16.6	6.9	10.0~10.4
	Seaside	1,532	99.3	445	24.7	21.4~22.2	25~28	20.2~22.4	44.9~47.1	5.3	42.6~44.9
	Total	2,097	98.8	431	32.6	27.2~29.3	28~29	28.6~31.2	61.1~63.7	12.2	52.6~55.3
Chugoku	San-in	140	96.1	348	1.7	1.8~ 1.2	54~67	3.5~ 3.0	5.2~ 4.7	1.5	3.9~ 3.4
	San-yo	674	97.9	387	9.3	10.5~10.9	43~50	16.9~19.8	26.2~29.1	2.7	25.3~28.4
	Total	814	97.6	379	11.0	12.3~12.1	45~52	20.4~22.8	31.4~33.8	4.2	29.2~31.8
Shikoku		412	95.9	401	5.8	4.7~ 4.8	68~71	11.7~12.4	17.5~18.2	4.9	13.6~14.3
Kyushum	North	853	95.1	363	10.7	8.1~ 9.4	36~36	10.8~12.5	21.5~23.2	3.6	19.2~21.1
	South	455	87.1	335	4.9	3.5~ 3.4	78~78	10.2~ 9.5	15.0~14.4	7.0	8.7~ 8.0
	Total	1,308	92.3	354	15.6	11.6~12.8	47~47	21.0~22.0	36.5~37.6	10.6	27.9~29.1
Okinawa											
National total		12,325	96.9	399	174.1	154.0	41~43	223.4~244.1	397.5~418.2	93.9	326.3~348.7

Note: 1. River water dependency includes intake loss, etc. of 7%.
2. Left-side figures under industrial water and urban water columns are computed on the basis of estimates by regional offices, while right-side figures computed based on report by Industry Rationalization Commission.

Source: Ministry of Land Development, Water Resources Bureau.

Note: P; Population D; Diffusion F.W.S.; Fresh water supplement
U.W.D.; Underground water dependency R.W.D.; River water dependency

Table 1.2-2 Water demand increment for 1976 - 1985

(Unit: hundred million m^3/year)

		Service	Industrial	Agricultural	Total
Hokkaido		3.7	5.9~ 7.3	3.4	13.0~14.4
Tohoku		5.9	7.0~ 8.2	7.4	20.3~21.5
Kanto	Inland	5.5	5.0~ 5.8	6.9	17.4~18.2
	Seaside	12.5	0.6~ 4.5	1.3	14.4~18.3
	Total	18.0	5.6~10.3	8.2	31.8~36.5
Tokai		5.6	1.0~ 6.2	3.9	10.5~15.7
Hokuriku		1.6	△0.2~ 1.2	1.8	3.2~ 4.6
Kinki	Inland	2.9	1.2~ 1.6	1.9	6.0~ 6.4
	Seaside	5.6	△0.3~ 1.9	0.8	6.1~ 8.3
	Total	8.5	0.9~ 3.5	2.7	12.1~14.7
Chugoku	San-in	0.6	1.2~ 0.7	0.8	2.6~ 2.1
	San-yo	3.3	△0.1~ 2.8	1.0	4.2~ 7.1
	Total	3.9	1.1~ 3.5	1.8	6.8~ 9.2
Shikoku		2.0	1.0~ 1.7	2.8	5.8~ 6.5
Kyushu	North	3.8	4.1~ 5.8	3.3	11.2~12.9
	South	1.7	2.7~ 2.0	3.2	7.6~ 6.9
	Total	5.5	6.8~ 7.8	6.5	18.8~19.8
Okinawa					
Nation-wide total		54.7	29.1~49.8	38.5	122.3~143.0

Note: 1. Service water and Industrial water are on supply basis.
2. Left-side figures under demand increment column are computed by Ministry of Land Development based on estimates by regional offices, while right-side figures computed on the basis of report by Industry Rationalization Commission in July 1975.

Source: Survey and review of water demand increment, in February 1976, by Ministry of Land Development, Water Resources Bureau.

Table 1.2-3 Transition of water supply on national scale

	1960	1965	1970	1974
Total population (thousand)	93,419	98,275	103,720	110,752
Population served (thousand)	39,858	56,422	72,361	85,135
Daily average supply (10^3m^3)	10,821	16,618	25,391	31,418
Daily average supply per capita (ℓ)	272	295	351	356
Daily maximum supply (10^3m^3)	13,494	21,483	32,644	40,102
Daily maximum supply per capita (ℓ)	339	381	451	467
Diffusion rate (%)	42.7	57.4	69.8	76.9
Load factor (%)	80.2	77.4	77.8	78.3

Source: "Statictis of Water works" by Japan Water works Association.

Note: Load factor is daily average supply/daily maximum supply.

Table 1.2-4 Transition of Industrial water requirement in terms of supply source

(unit: thousand m^3/day)

S / Year	Total	P.W.		Surface water	Underground water	Well	Others	Recovered water
		I.W.	S.W.					
1958	23,930 (100.0)	1,395 (5.8)	1,517 (6.3)	5,867 (24.5)	1,956 (8.2)	7,836 (32.8)	547 (2.3)	4,812 (20.1)
1962	37,320 (100.0)	2,216 (6.0)	3,069 (8.2)	7,321 (19.6)	2,587 (6.9)	11,092 (29.7)	672 (1.8)	10,363 (27.8)
1965	49,162 (100.0)	4,444 (9.0)	2,780 (5.7)	7,281 (14.8)	3,554 (7.2)	12,679 (25.8)	598 (1.2)	17,826 (36.3)
1970	85,042 (100.0)	9,801 (11.5)	3,491 (4.1)	8,286 (9.7)	3,247 (3.8)	15,360 (18.1)	871 (1.0)	43,986 (51.7)
1971	95,247 (100.0)	10,395 (10.9)	3,876 (4.1)	8,292 (8.7)	3,188 (3.3)	14,915 (15.7)	1,271 (1.3)	53,310 (56.2)
1972	101,457 (100.0)	11,491 (11.3)	3,530 (3.5)	8,257 (8.1)	3,163 (3.1)	15,243 (15.0)	884 (0.9)	58,889 (58.1)
1973	113,915 (100.0)	11,437 (10.0)	3,880 (3.4)	8,397 (77.4)	3,131 (2.8)	15,326 (13.4)	1,086 (1.0)	70,038 (62.0)
1974	120.040 (100)	11,995 (10.0)	3,351 (2.8)	8,192 (6.8)	3,065 (2.6)	14,646 (12.2)	1,000 (0.8)	77,790 (64.8)
1975	121,625 (100)	11,945 (9.8)	3,152 (2.6)	7,925 (2.4)	2,925 (2.4)	13,622 (11.2)	628 (0.5)	81,432 (67.0)

Note: Figures parenthesised represent share.

S; Source P.W.; Public waterworks
I.W.; Industrial water S.W.; Service water

Table 1.3-1 Desalting plants by process in Japan

($>100m^3$/day)

	Distillat		Reverse osmosis		Electro-dialysis		Total	
	N.P	P.C.	N.P	P.C.	N.P.	P.C.	N.P	P.C.
Island	3	3,850	–	–	5	2,520	8	6,370
Electricity	8	10,700	1	9,600	3	15,440	11	35,740
Industry	1	500	12	25,545	–	–	13	26,045
Others	–	–	1	400	2	500	3	900
Total	12	15,050	14	35,545	10	18,460	36	69,055

(as of December 1976)

Note: N.P.; Number of plants P.C.; Plant capacity (m^3/day)

Table 1.3-2 Fresh water cost by 100,000 m^3/day seawater desalting plant

(Dual-purpose plant with thermal power under big project, sponsored by Agency of Industrial Science and Technology)

		Fresh water cost	yen/m^3
Fixed	Depreciation	41.93	53.74
	Maintenance	2.04	
	Personnel	9.77	
Variable	Steam	101.29	117.00
	Electric power	12.10	
	Chemicals	3.61	
Total		170.74	

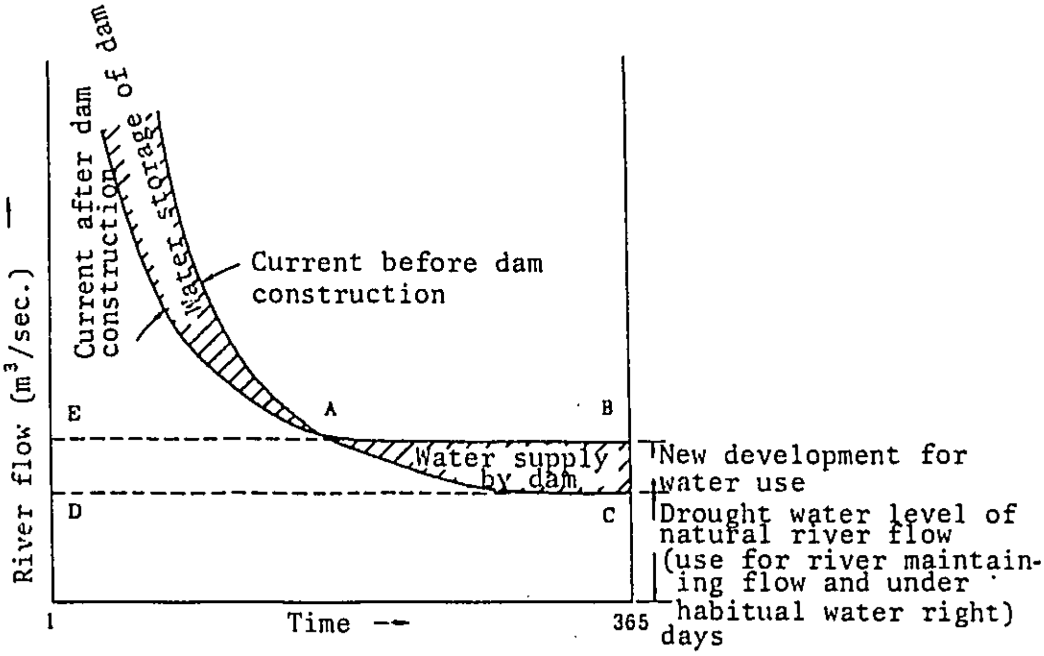


Fig. 1.1-1 Changes of current curve by dam construction

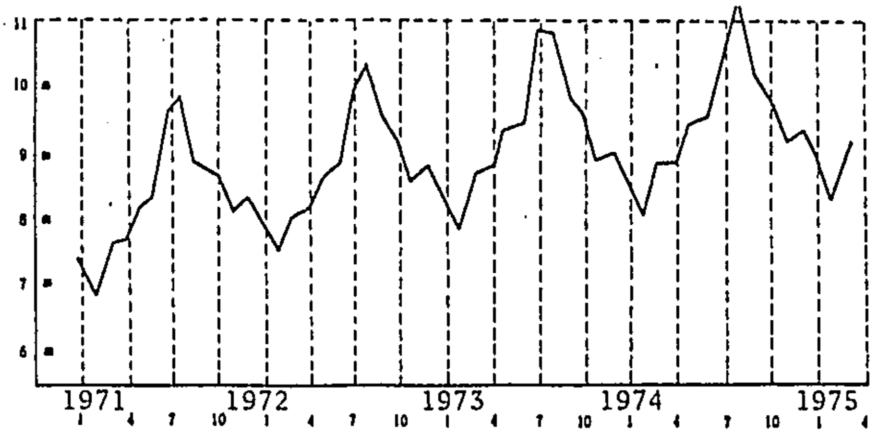


Fig. 1.2-1 Monthly supply of service water

Source: "Statictis of Water works" by Japan Water works Association

II. Nature of Seawater

2.1. Principal Nature of Seawater

2.1.1. Physical Properties of Seawater

Seawater is an aqueous solution containing about 3.5 percent electrolyte, which is higher in density and is much higher in electric conductivity and osmotic pressure than fresh water. On the other hand, it has a low freezing point and low vapor pressure and accordingly has a high boiling point. Its other properties include a rather greater refractive index and surface tension. Changes of these properties with a change of temperature or concentration are shown in the following table.

(1) Density (Concentration)

The density of seawater is more than 1, but is expressed as (d-1,000) x 1,000 in the table for convenience sake. Salinity and chlorinity are expressed as S $^0/_{00}$ and Cℓ $^0/_{00}$, respectively. The same method is applied hereafter.

Table 2.1 - 1 Density

S‰	Temperature (℃)						
	0	5	10	15	20	25	30
30	24.10	23.73	23.08	22.15	20.98	19.60	18.01
32	25.71	25.32	24.64	23.64	22.50	21.10	19.50
34	27.32	26.89	26.19	25.22	24.02	22.61	21.00
36	28.93	28.48	27.75	26.76	25.54	24.12	22.50
38	30.55	30.07	29.31	28.30	27.07	25.63	24.00
40	32.16	31.90	32.43	31.38	30.11	28.64	27.00

(2) Specific heat

The specific heat of seawater does not change greatly with a change of concentration or temperature and merely decreases slightly with an increase of concentration and increases slightly with a rise of temperature.

Table 2.1-2 Specific heat (Isopiestic specific heat)

S‰	Temperature (℃)						
	0	5	10	15	20	25	30
30	0.9585	0.9589	0.9592	0.9594	0.9599	0.9606	0.9611
32	0.9561	0.9563	0.9566	0.9570	0.9575	0.9580	0.9587
34	0.9532	0.9537	0.9539	0.9544	0.9551	0.9556	0.9563
36	0.9506	0.9511	0.9515	0.9520	0.9527	0.9534	0.9542
38	0.9480	0.9848	0.9489	0.9496	0.9503	0.9511	0.9520
40	0.9453	0.9458	0.9465	0.9472	0.9480	0.9489	0.9499

(3) Electric conductivity

Since seawater is a strong electrolyte, it has high electric conductivity. Accordingly, conductivity increases considerably with an increase of concentration and with a rise of temperature.

Table 2.1-3 Electric conductivity (Ω^{-1}, cm^{-1})

S‰	Temperature (℃)						
	0	5	10	15	20	25	30
30	0.02524	0.02910	0.03314	0.03735	0.04171	0.04621	0.05085
32	2677	3084	3512	3958	4419	4895	5386
34	2830	3259	3710	4179	4666	5167	5684
36	2982	3433	3906	4400	4911	5437	5980
38	3134	3606	4102	4619	5154	5706	6274
40	3285	3778	4296	4837	5396	5973	6567

(4) Refractive index

The refractive index of seawater is an important physical property for oceanography but is not especially important for a desalination process. The refractive index increases with an increase of concentration and decreases with a rise of temperature. Here, only the refractive index of standard seawater is referred to.

Table 2.1-4 Refractive index

S‰	Temperature (℃)					
	0	5	10	15	20	25
35	1.34088	1.34061	1.34030	1.33990	1.33940	1.33886

As seen from the values in the table, the refractive index of seawater is higher than that of fresh water, as the refraction of fresh water and that of dissolved ions is added to the former.

(5) Surface tension

The surface tension of seawater is an important physical property in relation to wave motion in oceanography and has a correlation with evaporation resistance. The surface tension of seawater is slightly greater than that of fresh water. It increases with an increase of concentration and decreases with a rise of temperature.

The empirical formula of surface tension (dyne/cm) = 75.64 − 0.144t + 0.099 Cℓ ‰, has been reported.

Table 2.1-5 Surface tension (dyne/cm)

S‰	Temperature(℃)			
	0	10	20	30
0	75.64	74.20	72.76	71.32
10	75.86	74.42	72.98	71.54
20	76.08	74.64	73.20	71.76
30	76.30	74.86	73.42	71.98
35	76.41	74.97	73.53	72.09
40	76.52	75.08	73.64	72.20

(6) Viscosity

The viscosity of seawater, though a strong electrolytic aqueous solution, is higher than that of fresh water. Like the surface tension, it increases with an increase of concentration and decreases with a rise of temperature. It is an important physical property, as it changes the thickness of laminar films on the heating surface of an evaporation vessel and on the surface of various membranes.

Table 2.1-6 Viscosity (Relative viscosity ηr)

S‰	Temperature (℃)					
	20	30	40	60	80	100
15	1.031	1.033	1.035	1.038	1.041	1.043
20	1.041	1.044	1.047	1.051	1.055	1.058
25	1.052	1.056	1.059	1.060	1.068	1.072
30	1.064	1.068	1.071	1.077	1.082	1.086
35	1.075	1.080	1.084	1.090	1.096	1.101
40	1.087	1.092	1.096	1.104	1.110	1.115

(7) Vapor perssure

The vapor pressure of seawater is lower than that of fresh water and decreases slightly with an increase of concentration but

Table 2.1-7 Vapor pressure mm Hg

S ‰	Temperature (℃)				
	20	40	60	80	100
0	17.55 (0.00)	55.35 (0.00)	149.4 (0.00)	355.2 (0.00)	760.0 (0.00)
20	17.33 (0.17)	54.75 (0.20)	147.8 (0.23)	351.5 (0.26)	751.9 (0.31)
34.32	17.18 (0.31)	54.30 (0.36)	146.6 (0.41)	348.6 (0.47)	745.6 (0.54)
40	17.18 (0.36)	54.08 (0.42)	146.1 (0.49)	347.4 (0.56)	743.0 (0.64)

increases sharply with a rise of temperature. Naturally, it is closely related to the raising of the boiling point.

Figures in parentheses in the above table represent a boiling point rise (°C). Seawater of S $^0/_{00}$ 35 at 25°C, for example, is assumed to have a vapor pressure of 21 mmHg and a boiling point rise of 0.52°C.

(8) Freezing point drop

The freezing point of seawater lowers of the lowering of the freezing point becomes greater with an increase of concentration. This property is an important index for saline water conversion using a refrigeration process.

Table 2.1-8 Freezing point drop (℃)

S‰	Concentration							
	5	10	15	20	25	30	35	40
E.P D. (℃)	- 0.268	- 0.535	- 0.801	- 1.068	- 1.341	- 1.621	- 1.906	- 2.196

(9) Osmotic pressure

The osmotic pressure of seawater is extremely high and is almost in proportional relation up to a certain level of concentration. It is needless to say that the osmotic pressure is very important for saline water conversion using a reverse osmotic pressure process. The osmotic pressure is influenced by the temperature.

Table 2.1-9 Osmotic pressure (atm.)

S‰	Temperature (℃)				
	25	40	60	80	100
10	7.1	7.4	7.8	8.1	8.4
20	14.3	14.9	15.7	16.4	16.9
34.5	25.1	26.3	27.7	28.9	29.9
50	37.5	39.3	41.5	43.3	44.7
75	59.3	62.4	65.9	68.8	71.1

(10) Heat conductivity

The heat conductivity of seawater is influenced very little by concentration but increases under the influence of the temperature.

Table 2.1-10 Heat conductivity(Kcal/m·hr.℃)

S‰	Temperature (℃)					
	0	20	40	60	80	100
20	0.488	0.517	0.540	0.559	0.573	0.582
35	0.486	0.515	0.539	0.558	0.573	0.582
40	0.486	0.515	0.539	0.558	0.573	0.583

The physical properties of seawater of a standard concentration assumed from the above tables may be summarized as shown in Table 2.1-11. In this case, however, the seawater is assumed to have a concentration of S $^0/_{00}$ 35 and a temperature of 25°C.

Table 2.1-11 Properties of normal concentrate seawater

D W=1	S.H. (W=1)	E.C. (Ω^{-1}, cm^{-1})	R.I. (W=1)	S.T. (Dyne/cm)	V. (ηr)	V.P. (m.m)	B.P.R (℃)	F.P.D (℃)	O.P (atm)	H.C. (Kcal/m hr,℃)
1.0234	0.9544	0.05303	1.33886	72.81	1.078	(21)	0.52	- 1.906	25.5	0.521

Note: D; density, S.H; specific heat, E.C.; electric conductivity,
R.I; refractive index, S.T.; surface tension, V; viscosity,
V.P.; vapor pressure, E.P.D.; freezing point drop, O.P.; osmotic pressure
H.C.; heat conductivity, B.P.R.; boiling point raising

In addition to these physical properties, the physical factors of seawater include diffusion constant, eco transmission, extinction of light and compressibility. These factors are important for analysis of marine phenomena but have no relation whatsoever to the utilization of seawater, especially to seawater desalination. These factors, therefore, are excluded from discussion. The secondary phenomena associated with the aforementioned physical properties, the latent heats of evaporation and freezing for example, are also excluded from discussion.

2.2.2. Chemical Properties of Seawater

(1) Relationship between chlorinity and salinity

The electrolyte concentration of seawater is 0.61N and if the activity is discounted, the ionic strength (I) of seawater is around 0.72. The greater portion of this electrolyte is chloride, in which sodium chloride accounts for nearly 78 percent as a salt. For this reason, the concentration index of seawater is generally expressed by chlorinity. In the past, the chlorinity of $C\ell^-$, 19 $^0/_{00}$ was often used as the standard value of ordinary seawater. This value, however, is a little too small to be applied to the entire sea area and the recent tendency is to use a salinity of S, 35 $^0/_{00}$ as the standard value. This value is equivalent to a chlorinity of $C\ell$, 19.374 $^0/_{00}$ according to the formula shown below. There is a close relation between chlorinity and salinity, and the following formula developed by Knudsen had been used for many years:

$$S\ ^0/_{00} = 1.8050\ C\ell + 0.030$$

After a study, however, the following formula, which has no correct numeral but is exactly proportional, has been adopted by UNESCO recently and is generally accepted.

$$S\ ^0/_{00} = 1.80655\ C\ell$$

The cholorinity referred to here is the total value with bromide ions and a trace of iodide ions contained in seawater also converted to equivalent chloride ions. Also, the salinity referred to here is not the total value of dissolved salts but is the value which considers bromide and iodide compounds as chlorides, carbonate as oxides and organic matters as completely decomposed by oxidization in minerals dissolved in 1 kg of seawater, based on the research made by Sørensen et al. The boric acid [$B(OH)_3$], which is not specified here, should also be considered as an oxide. The value by this definition is said to be smaller than the total value of salts contained in standard seawater by approximately 0.45 percent. This value is also considered to contain organic matter and others.

(2) Chemical composition of seawater

Seawater is said to contain as many as 78 elements dissolved in it. Of these elements, cases are nitrogen, oxygen, carbon dioxide and six rare gases including argon. Besides, various types of stable and many radioactive isotopes are known to exist in seawater, of which light isotopes such as ${}^{2}_{1}H(D)$, 3H(T), ${}^{14}_{6}C$, ${}^{16}_{8}O$, ${}^{18}_{8}O$ and heavy isotopes such as ${}^{230}_{90}Th$, (Io), ${}^{232}_{90}Th$, ${}^{235}_{92}U$, ${}^{238}_{92}U$ are being studied by researchers of chemical oceanography. Even natural seawater, which has not been contaminated by artificial radioactivity, shows very weak radioactivity and 98 percent of such radioactivity is due to the disintegration of ${}^{40}_{19}K$.

In general, the components which are contained in seawater in the amount of 1 mg/ℓ or more as elements, are called the main dissolved components of seawater with the exception of gases. Of these, silicon, which fluctuates greatly in the content, is listed in Table (a) but is often excluded from the category of main dissolved elements. Of these elements, several show almost constant component ratios regardless of the location of sea areas and concentration. This was confirmed by Dittmar and was published in 1884. This finding is based on the result of detailed analysis of 77 samples obtained from different sea areas of the world and the value obtained then is not so much different from the value of today. It was found thereafter, however, that the component ratio on the basis of chloride ions changes slightly depending on the location of sea areas, as in the case of calcium, for example.

(a) Main dissolved components of seawater

Values of main dissolved components of seawater, which are considered most appropriate at present, are shown below.

With the increase of the accuracy of analysis, these values may change more or less in the future, but the change will be very minor.

Table 2.1-12 Composition of seawater

Element, symbol		Contents ppm		Main chemical species
		S =35‰	C l =19‰	
Chlorine	C l	19, 354	18, 970	Cl^-
Sodium	Na	10, 770	10, 543	Na^+
Magnesium	Mg	1, 290	1, 266	Mg^{2-}
Sulphur	S	905	889	SO_4^{2-}
Calcium	Ca	412	404	Ca^{2+}
Potassium	K	399	393	K^+
Bromine	Br	67. 3	66	Br^-
Carbon	C	28	28	HCO_3^-
Strontium	Sr	7. 9	7. 8	Sr^{2+}
Boron	B	4. 5	4. 4	$B(OH)_3$, $B(OH)_4^-$
Silicon	Si	3	3	$Si(OH)_4$
Fluorine	F	1. 3	1. 3	F^-, MgF^+

Note: Figures are calculated proportionately to those obtained at S=35% . Generally, major elements show lesser coefficient of variation.

(b) State of main ions dissolved in seawater

Chemical species dissolved in seawater include ion states, molecular states, ion pairs, complex ions and others.
The numerical values obtained by Garrel et al are shown here. Calculation was troublesome because it involved the estimation of the activity coefficient (γ) of various chemical species, selection of a formation constant of ion pairs from the existing data or the measurement of it specially for this purpose and also the solution of simultaneous equations.

In recent years, however, calculation has become easier with the use of computers and numerous findings extending to other chemical species, which are contained in a very small amount, have been reported. However, values fluctuate greatly depending on the method of data collection, especially in the case of components contained in a very small amount, and for this reason, these reported findings may not be completely reliable in many cases. It should be added that the value of pH, Eh (oxidation-reduction potential) has an important bearing on this calculation.

Table 2.1-13(a) Chemical speciation of major components

Ion	Free ion (%)	Ion pair (%)			-2 Activity coefficient (γ)
		SO_4^{2-}	HCO_3^-	CO_3^{2-}	
Ca^{2+}	91	8	1	0.2	0.28
Mg^{2+}	89	11	1	0.3	0.36
Na^+	99	1.2	0.01	–	0.76
K^+	99	1	–	–	0.64

Note: Activity coefficient of Cl^- is 0.64.

Table 2.1-13(b) Chemical speciation of major components

Ion	Free ion (%)	Ca^{2+}	Mg^{2+}	Na^+	K^+	
SO_4^{2-}	54	3	21.5	21	0.5	0.12
HCO_3^-	69	4	19	8	–	0.68
CO_3^{2-}	9	7	67	17	–	0.20

(c) Nutrient salts in seawater

The nutrient salts in the form of inorganic bodies, which have very close relation to the production of oceanic life, include such nitrogen compounds as ammonium (NH_3, NH_4^+), nitrous acid (NO_2^-) and nitric acid (NO_3^-), phosphoric acid

(PO_4^{3-}) and silicic acid [$Si(OH)_4$]. Nitrogen and phosphorus have many types of organic compounds, organic carbon compounds such as saccharides and carbonate substances are complicatedly related with the production of oceanic life. Oceanic life, especially phytoplankton, propagates through photosynthesis by taking the above-mentioned nutrient salts from seawater. As a result, carbonate substances react and release oxygen as shown in the following formula, leading to the increase of pH.

$$nCO_2 + nH_2O + h\nu \rightarrow (CH_2O)n + nO_2$$

If zoo-plankton and hytoplankton die, the soma decomposes through a reaction which is almost the reverse of the above-mentioned reaction, consumes oxygen and then releases carbon dioxide into seawater. As a result, hydrogen ion (H^+) generates as shown below, leading to the decrease of pH.

$$CO_2 + H_2O \rightarrow H^+ + HCO_3^-$$

Generally, the content of nutrient salts fluctuates depending on the location of sea areas and the tidal current but is normally high in seawater of deep layers and seawater in high latitudes.

Table 2.1-14 Nutrient salts of oceanic seawater (μg.at./ℓ)

Nutrient salts	Surface layer seawater	Deep layer seawater	Maximum value
NH_3, NH_4^+-N	0 ~ 3	0 ~ 3	20
NO_2^--N	0 ~ 1	< 3	5
NO_3^--N	0 ~ 5	< 40	40
T-N	5 ~ 50	< 40	80
PO_4^{3-}-P	0 ~ 1	< 3	10
T-P	0 ~ 1	< 3	15
$Si(OH)_4$-Si	0 ~ 20	> 200	-

2.2. Regional Characters of Seawater

In the sea around Japan, the Kuroshio current of high salinity flowing in the northeast direction branches off at the south of Kyushu and part of the branched current becomes the Tsushima current flowing mainly along the Japan Sea coast in a northerly direction while becoming low in concentration under the influence of the East China Sea, and part of which flows through the Tsugaru Straits and moves down to the south along the Pacific coast. On the other hand, the Oyashio current of low salinity, which is abundant in nutrient salts, flows southward along the east coast of the Kurile Islands, moves to the northeast coast of Honshu Island and mixes with part of the main flow of the Kuroshio current which keeps flowing northward. This sea area where two currents meet each other often forms complicated water masses from a view point of both physical and chemical oceanography. Besides, the coast of Japan is featured by variable topography and water depths and where an upswelling current generates from deep layers, the content of both salinity and nutrient salts generally increases. In closed sea areas like the Seto Inland Sea and Ariake Sea, on the other hand, seawater often becomes low in salinity because of less circulation exchange of seawater and partly because of the effect of land drainage. In areas where industry, agriculture and stock raising have developed and in areas of high population density, the water quality deteriorates rapidly due to the effect of the influx of not only nutrient substances but also the so-called polluting substances. A similar tendency is seen in half-closed water areas such as Tokyo, Ise and Kagoshima Bays.

2.2.1. Composition of Seawater in the Offshore Area of Japan

Data on the composition of the main elements of seawater around Japan for specific months and years under specific conditions can be found here and there but the data showing quantitative values for the whole year are very few. Shown in Table 2.1-15 are mean values obtained by numerous analyses made in saltworks and other establishments (it was unavoidable that analyses were made mainly in the Inland Sea area).

For the purpose of a comparison, values of standard seawater of Cℓ 19 $^0/_{00}$ calculated from Table 2.1-11 are provided at the bottom of the table.

Table 2.1-15 Composition of coastal seawater

District	B.D. (15℃)	Temp (℃)	Ion concentration (N)						
			Cl^-	SO_4^{2-}	Ca^{2+}	Mg^{2+}	K^+	Na^+	total*
Tohoku East coast	3.59	19.7	0.535	0.053	0.019	0.104	0.009	0.457	0.586
Seto inland sea West side	3.19	19.8	0.502	0.051	0.018	0.099	0.008	0.427	0.553
Central (1)	3.34	18.7	0.500	0.052	0.018	0.098	0.009	0.427	0.552
(2)	3.50	19.7	0.517	0.051	0.019	0.101	0.009	0.439	0.568
(3)	3.38	20.2	0.512	0.054	0.020	0.104	0.008	0.434	0.565
Sagami Bay	3.50	19.1	0.537	0.056	0.020	0.106	0.010	0.458	0.593
Normal seawater	3.59	—	0.550	0.057	0.021	0.107	0.010	0.470	0.608

Note: For the sake of culculation sums of negative ions and positive ions do not coincide.

B.D.: Baumé degree

2.2.2. Chlorinity, pH, COD and the Nutrient Salts in Seawater in Offshore Areas of Japan

There are only a few cases in which measurements were taken during at least four seasons of the year and at many measuring stations in the same sea area. Of the result of observations made so far, the mean values of collected data are shown in the following table. These observations were made during a period from 1973 to 1977. The table shows that the seawater in inland seas and bays is slightly low in concentration and high in COD as compared with that of the open sea, with nutrient salts showing the same tendency.

Table 2.1-16 Chloride ion, pH, COD and Nutrient Salts

District	C (‰)	pH	*COD (ppm)	Nutrient salts (μg.at./ℓ)			
				NH_4^+	NO_2^-	NO_3^-	PO_4^{3-}
Hokkaido Uchiura bay	18.02	8.20	1.19	0.68	0.08	0.94	0.24
Tohoku North-east coast	18.24	8.19	1.62	–	–	–	
South-east coast	18.43	8.25	0.34	1.30	0.25	4.06	0.24
Tokyo bay for interior	16.17	–	–	52.5	4.0	16.6	2.55
Ise bay Central	16.23	8.30	0.77	3.97	1.01	8.65	0.54
Seto inland sea Central (1)	16.97	8.24	0.63	7.2	1.7	2.8	0.6
(2)	17.53	–	–	0.78	0.40	1.11	0.27
West side	18.14	8.25	1.61	–	–	–	
Japan sea East side	18.48	8.34	0.41	–	–	–	
West side	18.75	8.20	0.44	1.41	0.32	1.03	0.15
Kyushu Island Northern-east	18.77	8.31	0.46	1.05	0.06	0.48	0.11
West coast	18.34	8.29	0.45	0.94	0.22	0.47	0.12
South-west coast	18.92	8.29	0.31	0.67	0.11	0.97	0.15

Note: COD figures show those obtained by the Alkaline method.
C: Chlorinity

III. Various Processes of Seawater Desalination

Seawater is an aqueous solution, of which 96.6 percent is water and the remaining 3.5 percent is dissolved salts of various types. Seawater desalination means the removal of these dissolved salts from seawater and the method used for this process is called the seawater (saline water) desalination or saline water conversion, to remove salts from seawater, processes like the distillation process, crystallization process, reverse osmosis process, solvent extraction, electrodialysis process and ion exchange resin process, are well known. The distillation process and the crystallization process separate out the water content in seawater by making use of a phase change of water, while the reverse osmosis process and the electrodialysis process separate out the water content by means of pressure difference and potential difference by making use of specially made membranes (reverse osmosis membranes and ion exchange membranes, which are known generically as membrane processes. The solvent extraction process makes use of the difference in solubility of water in organic solvents with a change of temperature, while the ion exchange membrane process makes use of absorption of salt ions by an ion exchange membrane. The following is a brief description of the principles and characteristics of these processes.

3.1. Distillation Process

(1) Characteristics and types of the distillation process

In the distillation process, seawater is heated to evaporate the water content and the generated vapor is condensated to obtain fresh water. This process produces fresh water of better quality as compared with the crystallization process or the membrane process. Seawater boils at 100.5°C or overunder 1 atmospheric pressure but boils and evaporates below 100°C when the pressure of the evaporator is reduced as shown in Fig. 3.1-1. Evaporation of water requires evaporation heat (Fig. 3.1-1) which is retained by the generated vapor as latent heat. When the vapor is condensed, it given off its retained heat, which can be reused to heat (preheat) seawater or evaporate the water content of seawater.

The distillation process is divided into three methods, multiple effect distillation, multistage flash distillation and vapor compression distillation.

The distillation process has been used for many years as a means of concentrating solutions and is a technique with many proven results. In the case of multiple effect distillation, in particular, a three effect evaporator was already installed in Europe some 120 years ago and the vapor compression evaporator was adopted in Europe around 1910 and was applied to various industries in Switzerland and other countries. A test plant for concentration of seawater using this evaporator was conducted also in Japan in the last half of the 1920s. Multistage flash evaporation distillation is a new technique applied to saline water conversion, which was proposed by Silver of Great Britain in the 1950s when the research and development of saline water conversion was actively promoted.

The problems common to all types of the distillation process are scale deposition and corrosion of equipment by the effect of dissolved oxygen in seawater. For this reason, prevention of scale deposition and corrosion is essential technology in the design of a saline water conversion plant using the distillation process. As details of the distillation process and the main points of its technology described in Chapters 4 and 5, only a saline water conversion plant using the distillation process will be discussed briefly in the following.

(2) Multistage flash distillation

When the heated seawater is led to a vessel under reduced pressure, part of the water content of seawater takes evaporation heat from the seawater and then boils violently, thereby reducing the temperature of the seawater. This phenomenon of adiabatic expansion is called flash evaporation.

Fig. 3.1-3 is a simple diagram of a desalination plant using the multistage flash distillation. The plant consists of only three stills, but a plant for land use normally consists of several tens of stages of stills. Each still is divided into the evaporation room and the condensation room. The seawater supplied by pump is led through heat transfer tubes in the condensation room and is heated successively by vapor generated in each evaporation room. Then the seawater is further heated in the brine heater and is led to the first stage evaporation room. Each still is maintained under a specified vacuum condition by means of a vacuum device in the steam ejector and a pressure regulation orifice in the condensation room, and pressure difference is maintained in each evaporation room by means of orifices provided between stages of stills. As a result, the heated seawater flows from the high temperature first stage toward the low temperature third stage, while flash evaporating successively in each evaporation room, concentrates and then flows out as brine. The vapor generated in the evaporation room of each stage passes through a mist eliminator while releasing its retained heat into the seawater in the heat transfer tube, becomes condensed water and collects in the tray provided. When this condensed water moves from the high temperature stage to the low temperature stage, part of it flash evaporates, reduces its temperature by releasing its retained heat into the seawater in the heat transfer tube and is then discharged from the third (last) stage by a brine pump.

The multistage flash distillation process described above is called the once-through system, as seawater is discharged after only one pass through the plant. This system requires a large volume of seawater supply. For practical purposes, therefore, a multistage flash distillation plant using a brine circulation system is often employed. A plant of this system comprises a heat recovery section and a heat rejection section as shown in Fig. 4.4-1. The brine in the heat rejection section gives off its retained heat while it flash evaporates under the effect of cooling seawater and decreases its temperature and then circulates to the last stage heat transfer tube of the heat recovery section together with the make-up seawater. This operation increases the temperature difference between the brine and the fluid in the heat transfer tube and accelerates flash evaporation of brine. Moreover, the heat rejection section absorbs any change of the environment and temperature to ensure stable operation of the plant. Besides, this system requires deaeration, decarbonation and pH control, or the use of an inhibitor, for prevention of scale deposition only for make-up seawater and is advantageous because the use of chemicals can be held to a minimum.

(3) Multiple effect distillation

The multiple effect method is a process in which several evaporators for concentration of solution and connected in series and are designated as the first effect and the second effect evaporator and so on from the high temperature side. In this process, only the first effect evaporator uses steam from the boiler and the second effect evaporator and the subsequent evaporators use steam produced by the previous evaporators. In this manner, evaporation repeats corresponding to the number of evaporators to concentrate the solution to higher levels with high thermal efficiency or produce fresh water from the generated steam. A plant used in this process is called the multiple effect distillation plant. In this process, each effect evaporator reduces its pressure below saturation pressure according to the temperature of solution, and the evaporator of the lowest temperature is connected to a condenser and a vacuum device provided separately.

For general industrial use, evaporators of the vertical tube rising film type (the solution rises along the inner wall of a vertical tube from the bottom of the evaporator) are often employed. For desalination, the effect evaporator of the vertical tube falling film type, which is excellent especially in heat transfer coefficient, has been adopted and has proven its effectiveness.

Fig. 3.1-2 is a diagram of an effect evaporator of the vertical tube falling film type. This evaporator consists of an upper brine room, center calandria and a lower brine reservoir, with the brine room and brine reservoir linked to each other by calandria which is a group of vertical tubes. The pressure in the brine reservoir is held lower than that in the brine room by the effect of the orifice provided on the top of the calandria. The high temperature brine is pumped from the brine reservoir of the previous effect evaporator and fed into the upper brine room. When the brine passes through the orifice, part of the brine flash evaporates while reducing its temperature and flows down along the inner wall of the tube in the form of a film. This falling solution takes heat of vaporization from the high temperature steam produced in the previous effect evaporator by way of heat exchange, evaporates violently and falls into the brine reservoir. The steam in the calandria condenses and becomes fresh water. In the brine reservoir, brine and generated steam are separated from each other, flow into the brine room and calandria of the succeeding effect evaporator, respectively. In this manner, evaporation of brine and condensation of generated steam repeat consecutively.

The process described so far is the normal flow system in which solution and steam flow in the same direction. For multiple effect distillation, the reverse flow system, in which solution and steam flow in the reverse direction, is also adopted. For desalination, however, the normal flow system is adopted from the necessity of preventing scale deposition.

To prevent the deposition of hard scale, the seawater, after being treated by decarbonation for prevention of the deposition of soft scale, is fed to the first effect evaporator, namely, the low concentration brine is fed to the evaporator on the high temperature side. While the concentration ratio of brine (the ratio of concentration of seawater to that of brine) is limited to less than 2 with multistage flash distillation because of the possibility of hard scale deposition, a concentration ratio of about 3 is possible with multiple effect distillation.

(4) A combination of multistage effect distillation and multistage flash

A new system of desalination, which combines the multistage effect distillation and multistage flash distillation to make use of the heat rejection effect of the solution afforded by the latter and the high thermal efficiency of the former, has been proposed. Desalination plants of this combination include multi-effect multistage flash distillation (MEMS), by which multistage flash distillation is used as an effect evaporator and vertical tube multistage effect distillation (VTE/MSF), by which multistage flash distillation is used as a heater of make-up seawater.

The plant using the MEMS process is unique in that all the process of evaporation is flash evaporation, while a plant using the VTE/MSF process has an effect evaporator at every three or five stages of flash-still and brine in each effect evaporator returns to the original still as illustrated in Fig. 3.1-4. The multi-effect evaporator causes brine to flash-evaporate at the orifice on the top of vertical tubes to create the temperature difference necessary for heat transfer between the steam and the brine, while this plant has the portion corresponding to the orifice replaced with a 3 or 5 stage flash-still to provide the brine with sufficient temperature difference and the brine is fed to the effect evaporator and heated for evaporation by the steam produced in the previous effect evaporator. A trial calculation shows that the cost of desalination with this process is lower than that with the multistage flash evaporation, if applied to a large plant. For this reason, attention is now directed to the VTE/MSF process as a promising process for application to a large desalination plant of the future.

(5) Vapor compression method

When a gas is subjected to adiabatic compression, it generally shows a rise of temperature and condenses at a high temperature. Based on this theory, the vapor produced in an evaporator can be reused as heating steam for the same evaporator. This process is called vapor compression distillation. Fig. 3.1-5 is a diagram of a desalination plant using this process. The evaporator shown in the figure is heated initially with an external heat source, and once evaporation takes place, the temperature difference ($T_2 - T_1$) can be maintained with the use of evaporated steam as a heat source by operating a compressor. In this case, the electric power consumption is equivalent to several tenths of the evaporation heat (latent heat) and high thermal efficiency of the desalination plant can be expected. However, the compressor requires high quality energy such as high pressure steam or electric power, as its motive power. In this respect, vapor compression distillation is disadvantageous as compared with multistage flash distillation or multiple effect distillation, as the last two distillation methods can operate with steam of relatively low quality as a heat source. Because of this restriction, vapor compression distillation may be used only in combination with other distillation methods or may be applied only to a relatively small scale desalination plant.

3.2. Membrane Processes

The membrane process is a method which separates, concentrates, and desalts the solution by making use of the difference of osmotic (permeation) speed of substances through a membrane.

The membrane process is further divided into dialysis and osmosis. With the dialysis process, ions or solute are transmitted through a membrane, while with the osmosis process, a solvent, especially water, is transmitted through a membrane.

When a substance moves through a membrane, it requires some force or other. This force is called the driving force. Except for cases where accurate theoretical analysis is required, the following five phenomena may be considered as driving forces; potential difference, pressure difference, density difference, temperature difference and chemical reaction.

The reverse osmosis process uses pressure difference for driving force, while the electrodialysis process utilizes a potential difference for driving force.

The performance of a membrane can be expressed by the flow flux which indicates the transmission speed of a substance through a membrane and by the selectivity of material to be rejected. To express the flow flux of the dialysis process, the quantity of substances per unit time and unit area may be used, and for practical application, kg/m^2 day may be considered. The flow flux of the osmosis process is expressed by the volume per unit time and unit area, and m^3/m^2 day is used for practical purposes.

The selectivity is expressed by the transport number of cation in the cation exchange membrane, for example, in the case of the electrodialysis process and by the rejection rate of salt in the case of the reverse osmosis process.

The membrane process is said to be more suitable for brine than seawater. The brine referred to here means brackish water. Brackish water is water which is more dilute than seawater, but which is unfit to drink. The membrane process is best suited to desalination of brine and the ion exchange resin method is suitable for desalination of salt water of less than 500 ppm, which is more dilute than brine.

3.2.1. Reverse Osmosis Process

Here, the reverse osmosis process will be described only briefly, as details will be given at a later stage.

The reverse osmosis process is a method in which fresh water is produced with the use of a semipermeable membrane, which transmits mainly water, to gain a pressure difference against the osmotic pressure and utilizes this pressure difference as a driving force.

The semipermeable membrane of relatively high water permeability, made from acetyl cellulose powder by the Cast process, comprises a thin surface active layer and a thick porous support layer. This construction is suggestive of a new membrane provided with selectivity in the future. In other words, it is now possible to provide the thin surface layer with selectivity and increase the flow flux of the porous support layer. Membranes of the future are required to have a construction of different layers.

Materials used for reverse osmosis membranes include acetyl cellulose as a tubular type and spiral type module, and aromatic polyamide as a hollow fiber type module. These modules, however, are disadvantageous because they generally require full treatment of raw water in advance and have a relatively short life.

Nevertheless, the reverse osmosis process is characteristic in that it is not greatly influenced by the concentration of raw water. Of course, higher concentration of raw water increases the osmotic pressure, thereby reducing the effective pressure difference, but high-concentration is not detrimental in this case. This process, therefore, has been widely applied to experiments at desalination plants as a process of energy-saving. If the problem of performance is solved with time, this process will be adopted for practical use in the future.

3.2.2. Electrodialysis Process

(1) Introduction

The electrodialysis process is a method which provides electrolytic solution with a potential difference and uses ion exchange membranes having electivity as opposite (different) symbol ions for concentration and desalination of seawater.

For many years, Japanese soda manufacturers have been concerned about the supply of crude salt, which is entirely dependent on import and have been showing deep interest in the process which enables production of crude salt from seawater on an industrial scale.

The development of ion exchange membranes has brought a hope that the above-mentioned long standing question might be solved by the electrodialysis process. Though this method has not yet been completely successful despite the streneous efforts of soda manufacturers, salt for human consumption has been successfully manufactured as a by-product under the protection of the Salt Monopoly Law. The experience gained in the production of salt has contributed to the improvement of technology of the electrodialysis method to the highest level in the world and has provided valuable information on desalination.

(2) Principles of desalination and concentration of the ion exchange membrane electrodialysis process

It can be said that the cation exchange membrane transmits only cation and that the anion exchange membrane transmits only anion.

In the multi-chamber electrodialysis cell which has cation exchange membranes and anion exchange membranes arranged alternately as shown in Fig. 3.2-1, the application of DC voltage to the electrode on both ends together with the supply of brine into the cell creates a potential difference which causes cation to move to the cathode side and anion to the anode side. If this action is combined with the previously mentioned characteristics of the ion exchange membrane, the chamber in which ion is concentrated (concentration chamber) and the chamber in which ion is diluted (desalination or dilution chamber) appear alternately as shown in the figure.

Of the two electrode chambers, the anode chamber generates chlorine or oxygen and acidifies brine, while the cathode chamber generates hydrogen and alkalifies brine. To lessen this effect, a multi-chamber (300 pairs or so) is used.

(3) Types and characteristics of ion exchange membranes

Table 3.2-1 enemurates the characteristics of ion exchange membranes.

An ion exchange membrane is divided into the strong acidic cation exchange membrane (C) and the strong basic anion exchange membrane (A), and these two types are generally used for desalination. Besides, the strong acidic cation exchange membrane (SC) and the strong basic anion exchange membrane (SA), provided with selective transmission of monovalent ion, have been developed for concentration of seawater. In the future, it may be necessary to use these types of membrane for desalination.

When raw water contains a large amount of Ca^{2+} and $SO_4{}^{2-}$, for example, the concentration of which cannot be increased greatly in the concentration chamber, it becomes necessary to use a monovalent ion selective membrane having considerably low permeability of these ions.

Trade names and symbols of membranes are shown together like Selemion CMV or Aciplex Ca^{-2} in the table 3.2-1.

Since the table was prepared sometime ago, it may not be up to date. Asahi Chemical Industry Co., Ltd. has shifted to the production of reinforced membranes having a higher bursting strength.

The selectivity against ions of opposite symbols, which is the most important feature of ion exchange membranes, is expressed by the transport number, i.e. the transport number of cation in the case of cation exchange membranes and the transport number of anion in the case of anion exchange membranes. The table shows that a value of 0.9 or greater is constantly maintained even when there is a considerable increase of concentration.

As for the selective permeability among ions of the same symbol, the selective order of normal membranes is $K^+ > Ca^{2+} > Mg^{2+} > Na^+$ in the case of cation exchange membranes and shows a slightly greater value of $C\ell^-$ than $SO_4{}^{2-}$ in the case of anion exchange membranes, while the selective order of monovalent ion selective permeable membranes is $K^+ > Na^+ > Ca^2 > Mg^{2+}$ in the case of cation exchange membranes and $C\ell^- \gg SO_4{}^{2-}$ in the case of anion exchange membranes.

When the ion exchange membrane was applied to concentration of seawater in Japan, a low actual resistance was especially required. For this reason, the Japanese manufacturers made many attempts and have finally succeeded to reduce the actual resistance or the resistance per unit area ($\Omega \cdot cm^2$) to nealy 2 to 4. In the case of the membrane manufactured by Ionics of the U.S.A. which is used only for desalination of seawater, stress is put rather on durability and the actual resistance is as high as 12 $\Omega \cdot cm^2$. The Japanese made membrane has some strength and may excel Ionic's membrane for application to desalination, as it shows excellent performance in other aspects.

The thermal resistance-heterogeneity ion exchange membrane, though not listed in Table 3.2-1, has been developed by Mitsuibishi Petrochemical Co., Ltd. and is now in use for desalination. Even the conventional ion exchange membrane is capable of withstanding raw water at 40°C but this membrane is provided with properties to withstand a temperature of at least 80°C. Since the seawater in the Middle Near East has a considerably high temperature, the membrane is required to withstand a temperature of at least 40°C. All ion exchange membranes manufactured in Japan can withstand this temperature and in this respect, the electrodialysis method is superior to the reverse osmosis process. For application to desalination of seawater of this temperature range, the power requirement with the electrodialysis process is approximately 8 kWh/m^3, which is about the same level as that required by the reverse osmosis process. For ordinary seawater, however, the power requirement of the electrodialysis process is approximately 15 kWh/m^3, and in this respect, the electrodialysis process is no match for the reverse osmosis process.

(4) Electrodialysis cell and a flow chart of the desalination process

The electrodialysis cell is a filter press type (clamp type) cell consisting of an anion exchange membrane, a gasket with spacers forming a dilution chamber (desalination chamber), a cation exchange membrane, a gasket with spacers forming a concentration chamber and an anion exchange membrane, all clamped in this order from the anode side. The spacers provide passages of solution. Passages are formed in the dilution chamber to allow the flow of diluted solution (desalted solution) and passages are formed in the concentration chamber to allow the flow of only concentrated solution. A flow chart of this process is shown in Fig. 3.2-2.

The raw water is sand filtered, and the concentrated solution and the diluted solution (desalinated solution), after adjustment of concentration by raw water, are circulated in the concentration chamber and dilution (desalinated) chamber respectively. A portion of desalinated water is taken out in the form of fresh water as a product and a portion of condensed solution is completely discarded or led through the anode chamber or cathode chamber in part as an electrode solution and the acidified and alkalified solutions are merged again into one flow, which is circulated as an electrode solution after pH adjustment, while part of the solution is discarded.

The most suitable concentration of raw water for treatment by the electrodialysis process is 1,000 - 2,000 ppm. The electrodialysis process is most effective among many desalination process to produce fresh water of about 200 ppm from brine of this concentration. The higher concentration of raw water requires a higher electric current for removal of ions and the lower concentration results in the excessive dilution of desalinated water and causes an increase of electric resistance. In both cases, there is an increase of power consumption. In this respect, the electrodialysis process is disadvantageous as compared with other desalination process.

(5) Conclusions

The electrodialysis process used in Japan holds a unique position in the world. The reason is that while researchers of other countries where concentrating their efforts on the application of the electrodialysis process only to the production of fresh water from the sea, the Japanese researchers were actively engaged in the development of a process for production of salt through concentration of seawater. As a result, the concentration technique using the electrodialysis process in Japan has advanced to a level which is now unrivalled by others. It cannot be denied, however, that there was very little research done for production of fresh water from the sea in the past because of a very small demand for it in Japan and that there was an easygoing attitude on the part of Japanese researchers who believed that desalination can be accomplished as a natural consequence of concentration of seawater. Full-scale research has been under way since several years ago for both desalination and production of fresh water from the sea and the technique in these fields has now attained world level and the manufacturing technique of membranes now ranks first in the world. Great expectations are being placed in the future development of this process.

3.3. Crystallization Process

In the crystallization process, seawater is maintained at a low temperature to precipitate crystals of ice or gas hydrate and fresh water is produced by the separation of crystal sofice or gas hydrate. The method used to obtain fresh water by the separation of ice crystals is called the freezing process and the method which produces fresh water by the separation of gas hydrate is called the gas hydrate process. Ice crystallizes when the temperature of seawater is maintained below the freezing point (-1.8°C) and gas hydrate crystallizes at a temperature of around 10°C, with the temperature varying somewhat depending on the kind of hydrate agent used. Since these crystals do not include dissolved salt, the brine adhered to the surface of the crystal is washed off and the crystal is then melted by increasing the temperature to produce fresh water. The crystals of gas hydrate are first decomposed to the original hydrate agent and water, and are then separated to produce fresh water. However, the quality of water produced by the crystallization process is dependent on the shape or size of crystals and also on the technique of separation from the solution and is, therefore, inferior to that of water obtained with the distillation process.

The salinity of fresh water produced with this process is in the range of 100 ppm to 500 ppm when the most advanced technique is applied. The heat of crystallization of gas hydrate or ice is in the range of 100 Kcal/kg to 80 Kcal/kg as compared with the heat of vaporization of 540 Kcal/kg. Such being the case, there is substantial by less thermal-conferment and the theoretical energy requirement of the freezing process is calculated at about 3.5 Kcal (4 kWh/m^3 - fresh water) per kilogram of fresh water. Besides, corrosion of equipment is very slow because the operating temperature is below normal and since scale deposition on the heat transfer surface is eliminated completely eliminated only deaeration for pre-treatment of seawater is required.

3.3.1. Freezing Process

(1) Desalination facilities by freezing process

The freezing process is divided into the indirect freezing process which cools seawater indirectly by way of a thermal transfer surface and the direct freezing process which freezes seawater by way of direct contact with a refrigerant. The direct freezing process is further divided into the vacuum freezing process, which creates a vacuum (less than 3 mmHg) inside the plant and freezes seawater by way of evaporation of the water content and the secondary refrigerant freezing process, which uses refrigerants insoluble in seawater (butane or freon P-114, which do not form gas hydrates). Of these two, the indirect freezing process has a restriction on production capacity of the heat transfer surface, and ice crystals produced are of poor quality. Such being the case, this process was considered unfit for desalination of seawater and efforts were made mainly for the development of a saline water conversion plant using the direct freezing process. Main desalination plants using the direct freezing process, which have so far been developed and tested, are shown in Table 3.3-1.

Fig. 3.3-1 is a diagram of the basic process of desalination plant using the direct freezing process.

Seawater is conducted through the deaerator and the heat exchanger to the crystallizer, where granular ice crystals are formed by the evaporation of secondary refrigerants and seawater is concentrated to about twice the concentration of normal seawater. The ice crystals are separated from brine and washed by a counter flow of part of the fresh water produced in the wash column and is then transformed to fresh water in the melting tank. Part of the brine separated in the wash column circulates to maintain slurry concentration in the crystallizer at about 15 to 20 percent and the rest is discharged from the heat exchanger.

The secondary refrigerant evaporated in the crystallizer is compressed by the operation of the main compressor, condenced in the melting tank by way of heat exchange with ice, separated from the melted ice (fresh water) and returned to the crystallizer. Since the process is a low temperature operation, the heat from the compressor, pumps, agitator and external thermal sources must be absorbed. The secondary refrigerant freezing process requires an auxiliary compressor and the vacuum freezing process, which uses water as refrigerant, requires an auxiliary refrigerator.

(2) Vacuum freezing process

The vacuum freezing process is divided into the vapor absorption process and the vapor compression process. The vapor absorption process is unique in that the plant does not require a compressor but requires a vapor absorption column and an absorbent regeneration column instead. A desalination test plant (57 m^3/day) using 50% lithium bromide as absorbent was operated. The conclusion was that the cost of installation is too high and the process is unfit for large installations, with the cost of fresh water being higher than that by other processes.

From the requirements of the desalination plant using the vapor compression process to operate the entire system at a steam pressure close to a vacuum condition, a system integrating a compressor, a crystallizer and a melter into one unit (hydro converter) was developed jointly by Israel and the U.S.A. A test plant (228 m^3/day) of this system was operated. The excellent results included an average water production of 350 m^3/day, which was greater than the design capacity, a total power consumption of 12.7 kWh/m^3 and an average salt concentration of 323 ppm in the fresh water. In practical application, however, the maintenance of the system under a vacuum condition below 3 mmHg required a heavy load and as a result, a new system which houses a hydro converter and wash column in one unit has been proposed.

(3) Secondary refrigerant freezing process

With the previously mentioned vacuum freezing process, the amount of fresh water production is restricted by the capacity of the compressor which must treat a large volume of steam at a low pressure. With the secondary refrigerant freezing process which uses butane and freon compounds, whose saturation pressure at temperatures of seawater close to freezing point is approximately one atmospheric pressure, as secondary refrigerants, the specific volume of steam is so small that the amount of steam treated by compressor increases more than 200 times that treated with the vacuum freezing process, and then the amount of fresh water production increases by ten times. This process, therefore, is said to be suitable for a large desalination plant.

The butane freezing process, which uses butane as secondary refrigerant, was developed in Japan, and based on this development, a pilot plant (56 m^3/day) using the butane freezing process was installed and operated in the United States. The published result shows a salinity of less than 500 ppm in the fresh water produced, a loss of wash water in the range of 2 to 16 percent of the amount of fresh water produced and a power consumption of 9.2 kWh/m^3 for a plant having a capacity of 3,000 m^3/day. Also in Japan, research was carried out to apply this process practically, and after experiments with a test plant having a capacity of 45 m^3/day, a process which uses liquefied natural gas (LNG) instead of a butane compressor for cooling heat source has been developed.

As a result, the relationship between LNG consumption and power consumption in desalination plant using the butane freezing process has been established as shown in Table 3.3-2.

Butane is most suitable secondary refrigerant for use in a desalination plant, but because of their combustible and explosive nature, they require special cautions for safety of the plant. In this respect, freon compounds of R-C318 or R-114 are non explosive.

In the United States, a study was made on the use of these freon compounds as secondary refrigerants and a desalination plant using R-C318 as secondary refrigerant was designed for analysis. The conclusion was that R-C318 was an expensive refrigerant, with the cost of fresh water calculated for this design being higher than that by the vacuum freezing vapor compression process.

In the United States, development of a desalination process using R-114 as a secondary refrigerant being in 1970, succeeding the previously mentioned technique of the secondary refrigerant freezing process. Since the refrigerant R-114 has a greater density than seawater, the crystallizer of the plant is so designed that the refrigerant and seawater are mixed together by spray nozzle and injected into the crystallizer from the top. Calculation of the cost of fresh water produced by this plant shows a power consumption of 12.2 kWh/m^3 for a plant having a capacity of 380 m^3/day and 10 kWh/m^3 for a plant having a capacity of 3,800 m^3/day. Application of this process for concentration of levels up to a eutectic point for treatment of industrial waste water is also being considered.

3.3.2. Hydrate Process

(1) Principles of the hydrate process

Chlorine, carbon dioxide, hydrocarbons (excluding normal butane) of up to C_4 class, including methane, and ethane, and freon compounds are known to form hydrates in coexistence with water. These compounds, which form hydrates, are called hydrating agents. The hydrate is a cage type clathrate compound, in which water molecules envelope the molecule of hydrating agent and form a crystal, and has two different crystal structures, type I and type II, depending on the molecule size of the hydrating agent. Table 3.3-3 shows the properties of main hydrating agents and hydrates.

Formation of hydrates is dependent on the temperature and pressure. Fig. 3.3-2 is a phase diagram of the water + hydrating agent system. The line EBF represents a saturation vapor pressure line of the hydrating agent, line BC is a hydrate formation line and line CD indicates a freezing line of water. Point B is an invariable point (degree of freedom = 0), where four phases - water (liquid phase), liquid phase and vapor phase of the hydrating agent and hydrate (solid phase) coexist and where the pressure and temperature show their inherent values in each hydrating agent, and is the critical decomposition point of gas hydrate. The pressure and temperature of each hydrating agent at the critical decomposition point are shown in Table 3.3-3. Point C is an invariable point where four phases - hydrate and ice of the solid phase, water of the liquid phase and hydrating agent of the vapor phase, coexist, and the temperature at this point is 0°C. In other words, the area surrounded by line ABCD represent the range of hydrate formation. In the case of the solution-hydrating agent system, it is known that the hydrate formation line shifts to the low temperature side as shown by the dotted line A'B'C' depending on the concentration of the solution.

(2) Development of a desalination plant using the hydrate process

Of the various hydrating agents, propane or freon compounds are hardly soluble in water and have the nature of refrigerants. It is possible, therefore, to mix them directly with seawater, utilize their vaporization heat to maintain the temperature of seawater within the range of the formation of hydrate and obtain crystals of hydrate. For this purpose, a desalination plant using propane or freon R-12 or freon R-21 as a hydrating agent has been developed.

A flow sheet of a basic desalination plant using the hydrate process is shown in Fig. 3.3-3. When the liquefied hydrating agent is brought into contact with seawater under temperature and pressure suitable for the formation of gas hydrate in the crystallizer, the generated heat for formation of a hydrate is removed by the heat of vaporization of the hydrating agent. The slurry of hydrate and brine is transferred to the wash column and the brine is washed off with a counter flow using part of the fresh water produced. Hydrating agent vaporized in the crystallizer is compressed and condensed in the melter and at the same time the hydrate is resolved into fresh water and hydrating agent. The liquefied hydrating agent is returned to the crystallizer and the hydrating agent mixed in fresh water in a very small amount is removed by separator.

The seawater, after being emptied of air and other gases by deaerator, is pre-cooled with fresh water and brine before being transferred to the crystallizer. Part of the brine is circulated to the crystallizer for adjustment of the amount of solids in the slurry. An auxiliary compressor is required to remove heat from the atmosphere, pumps and main compressor as in the case of the previously mentioned freezing process.

Since the hydrate generally forms at temperatures higher than the freezing point of seawater as shown by the critical decomposition point in Table 3.3-3, this process uses higher operating temperatures than the freezing process and then is economically advantageous. However, the hydrates produced with this process do not grow to large crystals, for example, the hydrates of propane and freon R-12 are small crystals of 10 to 50 microns. These crystals also form compressible cakes, which makes separation and washing operations extremely difficult. The development of a desalination plant using the hydrate process has been directed mainly to the separation and washing techniques and tremendous efforts are being made to reduce the concentration of salt in the production fresh water to a level below 1,000 ppm. This problem, however, has not yet been solved technically.

3.4. Other Processes

3.4.1. Ion Exchange Resin Process

(1) Introduction

The ion exchange resin was the first material applied to the desalination process on an industrial scale without the use of heat.

The ion exchange resin process was overestimated initially but is now being evaluated correctly and the technology had advanced to a level so that practically pure water can be obtained from a relatively dilute solution (100 ppm, for example) with a process on an industrial scale which completely desalts the solution.

This process, however, is not suitable for desalination of seawater (including brine) because of its excessive concentration and is used only for desalination of salt water of less than 500 ppm. The process is often applied to the manufacture of pure water from dilute solutions produced by other desalination processes.

The problem of the ion exchange resin process is regeneration of the resin. The problem of regeneration in the case of the mixed bed desalination process was solved by separation of cation exchange resin and anion exchange resin with the use of the specific gravity difference but the use of acid and alkali for regeneration has brought about a new problem waste water.

To cope with this situation, a new method, which uses a resin having a nature of weak acidity and weak basicity to desalinate salt water and also uses hot water for regeneration, has been developed recently and is drawing attention. This process is said to be capable of desalinating salt water of up to 3,000 ppm.

(2) Principles and method of desalination with ion exchange resin

If seawater is treated with the hydrogen type cation exchange resin and the hydroxyl type anion exchange resin, desalination of seawater expressed by the following formula may be accomplished.

$$HR+R'OH+Na^{+}+C\ell^{-}\rightarrow NaR+R'C\ell+H_2O$$

The methods used up to now are the mixed bed type and the dual bed type.

With the mixed bed type, cation exchange resin and anion exchange resin are mixed in one column for desalination and very pure desalted water can be obtained. With the dual bed type, on the other hand, salt water is treated firstly with cation exchange resin for desalination,

$$HR+Na^{+}+C\ell^{-}\rightarrow NaR+H^{+}+C\ell^{-}$$

and is then treated with anion exchange resin,

$$R'OH+H^{+}+C\ell^{-}\rightarrow R'C\ell+H_2O$$

While the purity of desalted water is not so high, regeneration in this method is simple and economical. For application to desalination, the dual bed type is generally employed. The resin used for regeneration of hot water is the amphoteric ion-exchange resin having a nature of weak acidity and weak basicity, and the principles of desalination with this method are the same as those in the mixed bed type.

(3) Method of resin regeneration

Since the exchange volume (capacity) of resin is fixed, resin must be regenerated after a fixed amount of fresh water is obtained.

Regeneration of resin in the dual bed type is simple and easy. After treatment until the break-through of ions, the cation exchange resin can be regenerated with the use of hydrochloric acid and the anion exchange resin can be regenerated with the use of an aqueous solution of sodium hydroxide according to a fixed rule.

Regeneration of the mixed bed type cannot be operated in the conditions of the mixed bed. After supplying seawater into the column from the bottom to separate high density cation exchange resin in the lower section and low density anion exchange resin in the upper section, these resins must be regenerated separetely.

This regeneration process, however, also generates waste water containing acid and alkali and presents a problem of waste treatment. Then, interest was directed to the hot water regeneration method.

This is a method by which a resin having properties of both weak acid and weak basicity is used for desalination based on the principles shown below.

$$RCOOH + R'N + Na^+ + Cl^- \underset{80℃}{\overset{20℃}{\rightleftharpoons}} RCOONa + R'NHCl$$

In other words, salt water is treated at 20°C (low temperature) for production of fresh water and then treated with hot water at 80°C for regeneration.

For pretreatment, the raw water is first softened with the Na type cation exchange resin as required and after hydrogen carbonate ions (HCO_3^-) have changed to carbon dioxide (CO_2) with the addition to acid, it is emptied of oxygen and carbon dioxide through vacuum deaeration and is then neutralized. When a large quantity of suspension exists, it must be removed according to a fixed rule. For the resin column, a fixed bed is used and cold salt water is supplied from top to bottom to obtain fresh water and then hot water is supplied from bottom to top for regeneration.

This method is used for desalination of brine, and the fresh water produced is further treated by the mixed bed desalination process for production of pure water as required.

The hot water regeneration method is being studied and developed jointly by the Seawater Conversion Promotion Center (Foundation) and Mitsubishi Chemical Industries, Ltd. at the Minami-sunamachi water treatment plant of the Tokyo Metropolitan Government for desalination of city sewerage.

(4) Conclusions

For desalination by ion exchange resin, a method using H-type cation exchange resin and a method using O-type anion exchange resin are employed. Though these methods presented some problems with regard to regeneration of resin and treatment of waste water of the regeneration process, use of the hot water regeneration method has made it possible to apply this process to desalination of salt water and interest is being directed to the future development of this process.

3.4.2. Solvent Extraction Process

A study of desalination by the solvent extraction process was first carried out by D.W. Hood et al. of Texas A & M Research Foundation under contract with the Office of Saline Water of the United States in 1953. They first referred to as many as 400 kings of organic solvent. A solvent suitable for desalination must have properties to dissolve water when brought in contact with saline water and change greatly in solubility according to the change of the temperature but minimize dissolution of the salt. Triethl amine, for example, dissolves water at the rate of 30 percent at a temperature of 20°C but dissolves water at the rate of only 2.5 percent at a temperature of 50°C as shown in Fig. 3.4-1. If the solvent is brought into contact with saline water at 20°C for a sufficient time, and then the solvent phase is separated and heated at 50°C, the greater part of water dissolved in the solvent phase can be separated as an aqueous solution phase. Since this aqueous solution phase contains about 3 percent dissolved amine, the aqueous solution phase flowing in the counter-current with steam in the column separates amine for recovery.

Fig. 3.4-2 is a flow sheet of a desalination plant using the solvent extraction process. In the extraction column of this plant, a cooled solvent supplied from the bottom flows in the counter-current with raw water supplied from the top extracts the water content of raw water and flows out from the top. The solvent which has dissolved water exchanges heat with the hot solvent and produces fresh water circulating from the phase separator and then flows into the phase separator, where it separates most of the dissolved water. The separated solvent is recirculated to the extraction column after heat exchange. The fresh water separated in the phase separator, on the other hand, passes through a steam stripper after heat exchange to remove and recover a very small amount of dissolved solvent and is taken out as production water. The brine flowing out from the bottom of the extraction column also passes through the steam stripper before being discharged, as it also contains a dissolved solvent.

The extraction column is not necessarily a single stage column but can be a several-stage column for improvement of the extraction efficiency. If the extraction temperature is lower than the ambient temperature, a process, which operates a compressor provided between the extraction column and the phase separator to evaporate part of the solvent in the extraction column, and absorbs in the phase separator part of the heat of the solution accompanying the extraction, may also be employed. The heat of the solution in this case is approximately 40 Kcal/kg H_2O, which is about half the heat of fusion of ice.

The advantage of the extraction process is that the operating temperature is close to ambient temperature and that heat energy of lower grades can be utilized. It is also possible to adapt the process to the environment with the use of appropriate solvents shown in Fig. 3.4-1.

It is also advantageous in that it minimizes energy loss, as the energy transfer (heat of solution) is smaller than that of the distillation process (530 Kcal/kg H_2O) and the freezing process (80 Kcal/kg H_2O) and that it is simple in operation, as conversion taken place only in the liquid phase. On the other hand, its disadvantages are; (a) that solvent dissolves not only water but also a part of the salt, requiring the repeated extraction of solvent if high concentration saline water is to be obtained; (b) that a large heat transfer area is required for recovery of heat; (c) that it requires operations to recover dissolved solvents in the production fresh water and in the brine to be discarded. A chemical reaction of the components in raw water may also take place at times. The amine shown in Fig. 3.4-1, which has strong basicity, causes separation of magnesium hydroxide in the presence of magnesium ions.

With the solvent extraction process, there are cases in which water is extracted at temperatures lower than that required phase-separation. When phenol is used as a solvent, for example, extraction of water takes place at 25°C, while phase-separation take place at temperatures of 0 to 11°C, which is the reverse of the preceding process. In other cases, benzene was used as a solvent and extraction was made at high temperatures and pressure. In all cases, the operating method differs depending on the type of solvents used. To date, solvents which are superior to amine or mixed solvents of amine have not been discovered, and application of the solvent extraction process to desalination has seldom been attempted because of its major disadvantage that the solvent dissolves in the aqueous solution phase.

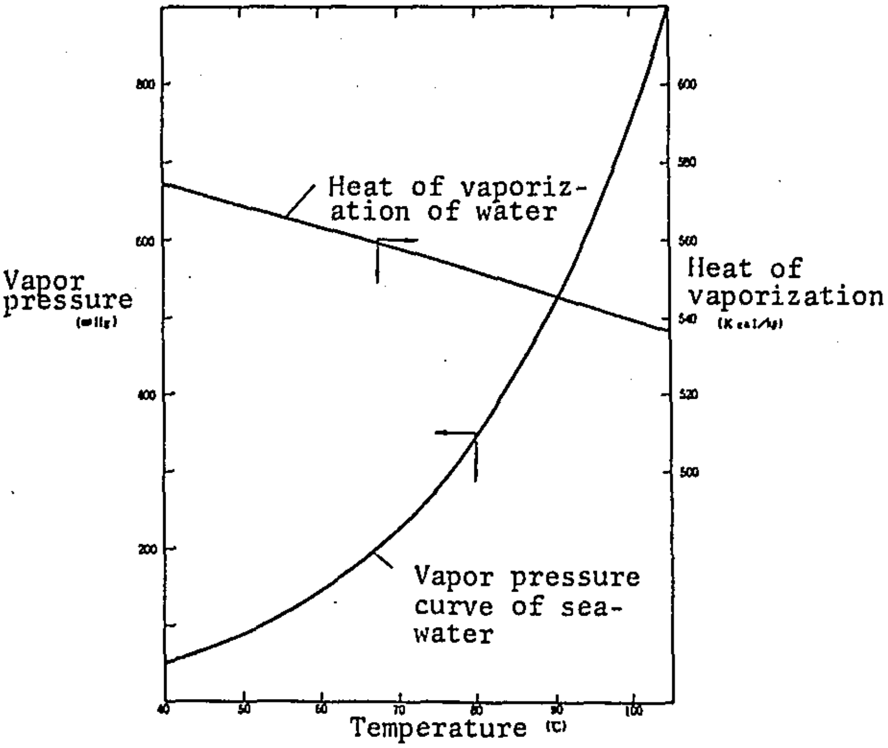


Fig. 3.1-1 Vapor pressure curve of seawater (salt concentrate 3.5 WT%) and heat of vaporization

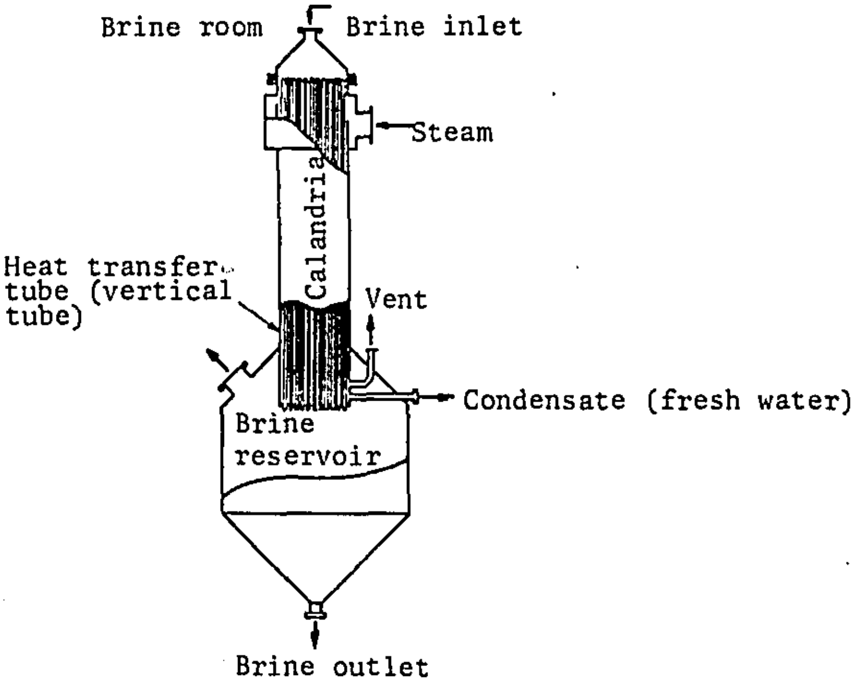


Fig. 3.1-2 Schematic diagram of thin film vertical tube effect evaporator

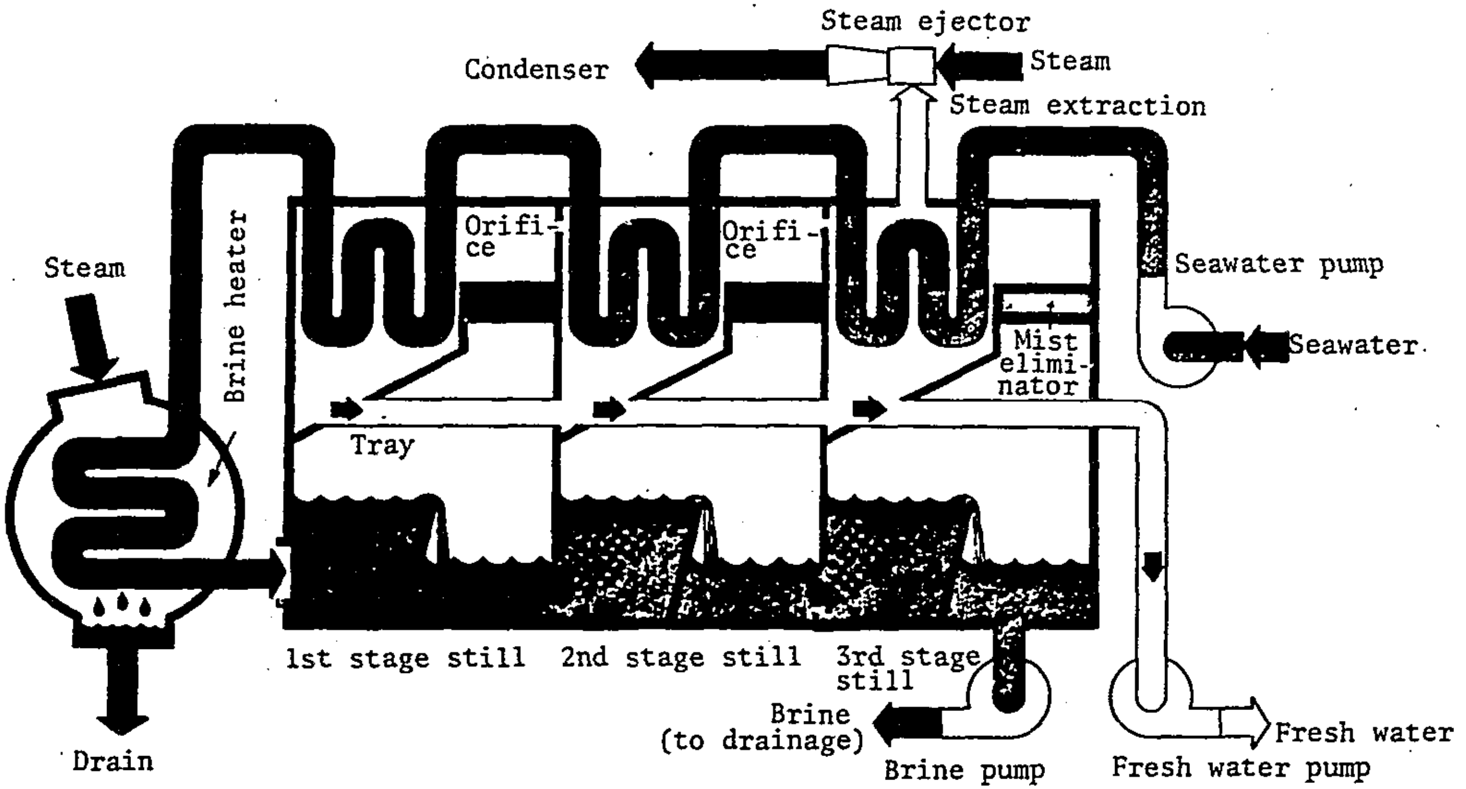


Fig. 3.1-3 Schematic diagram of multistage flash distillation process

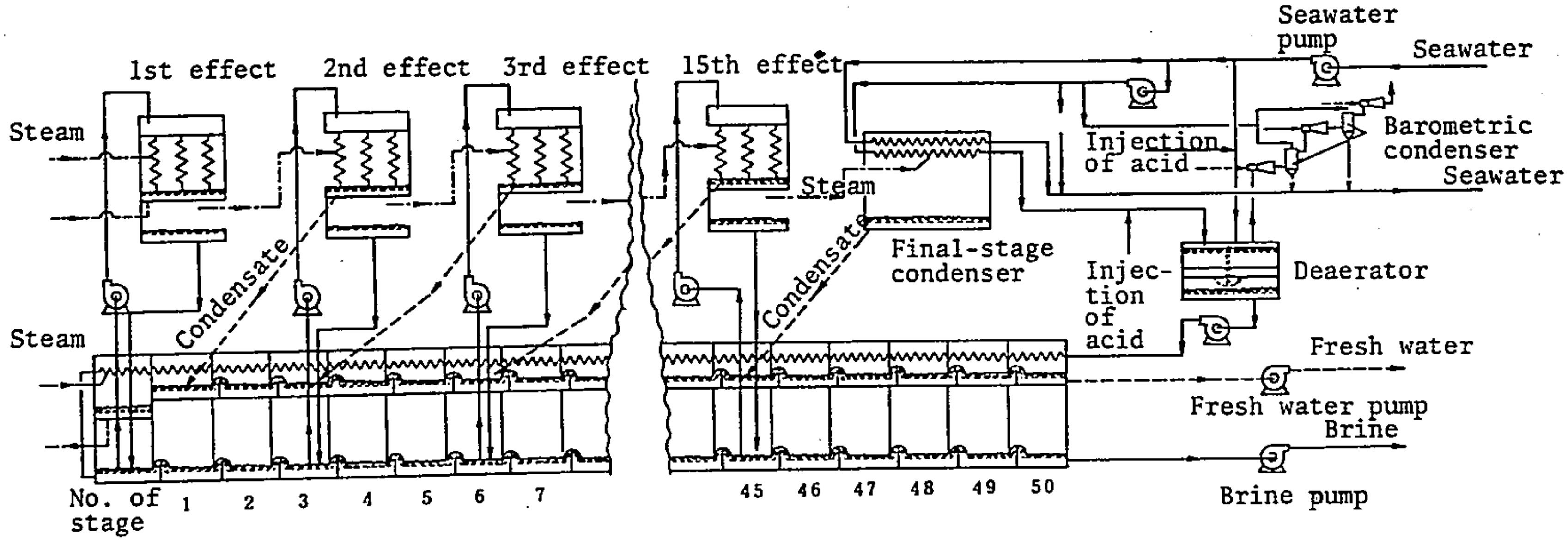


Fig. 3.1-4 Schematic diagram of desalination plant by VTE/MSF process

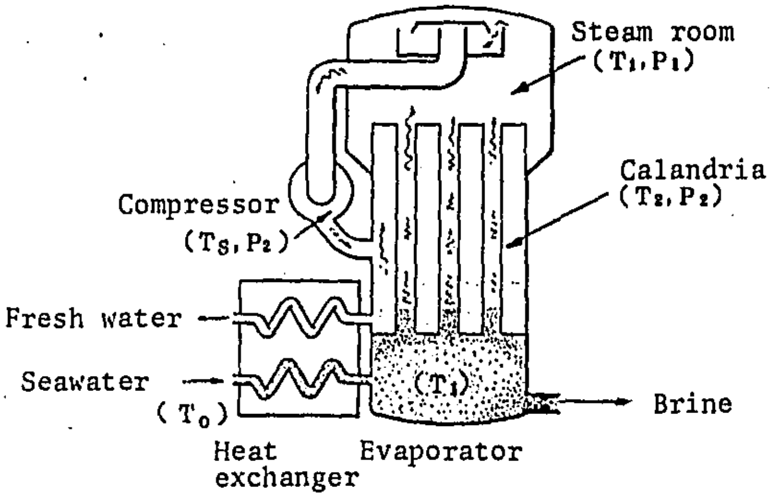


Fig. 3.1-5 Scehmatic diagram of desalination plant by vapor compression to process

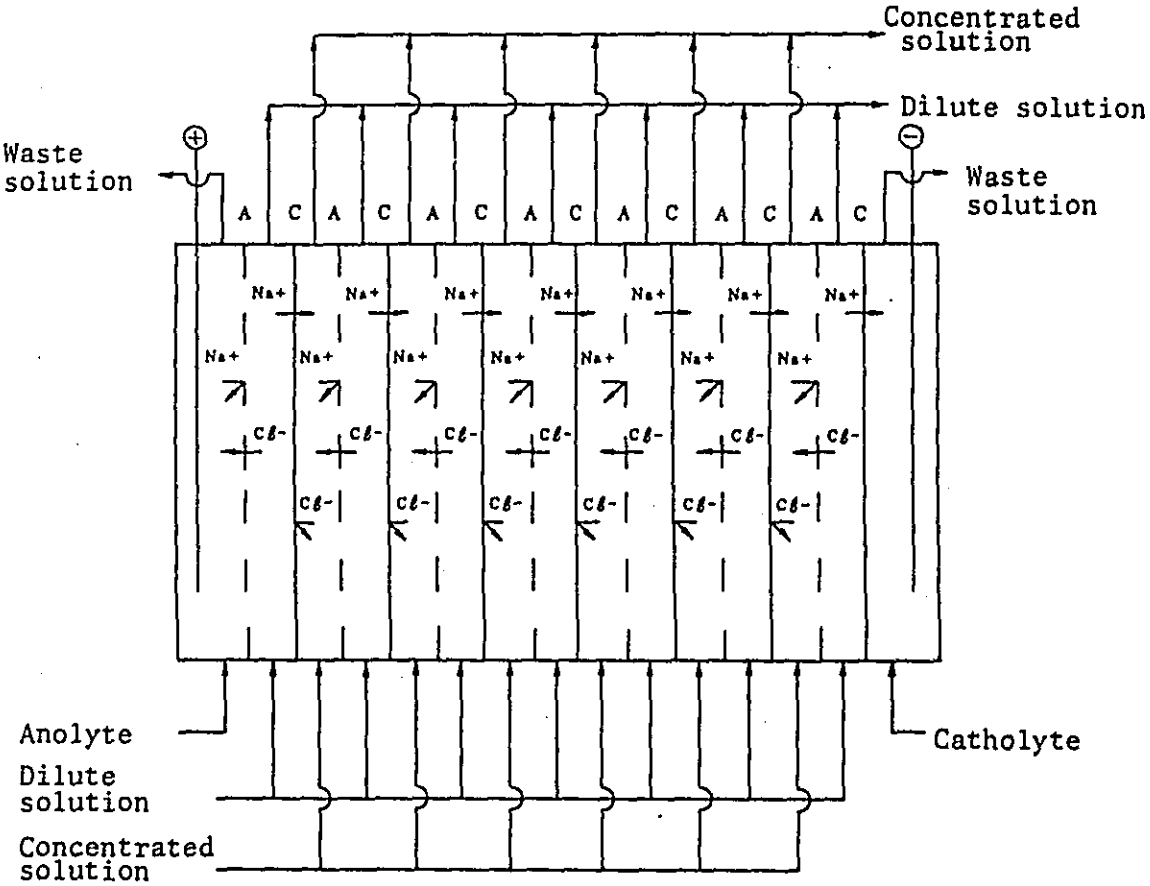


Fig. 3.2-1 Concentration and desalination of solution by ion-exchange membrane electrodialysis process

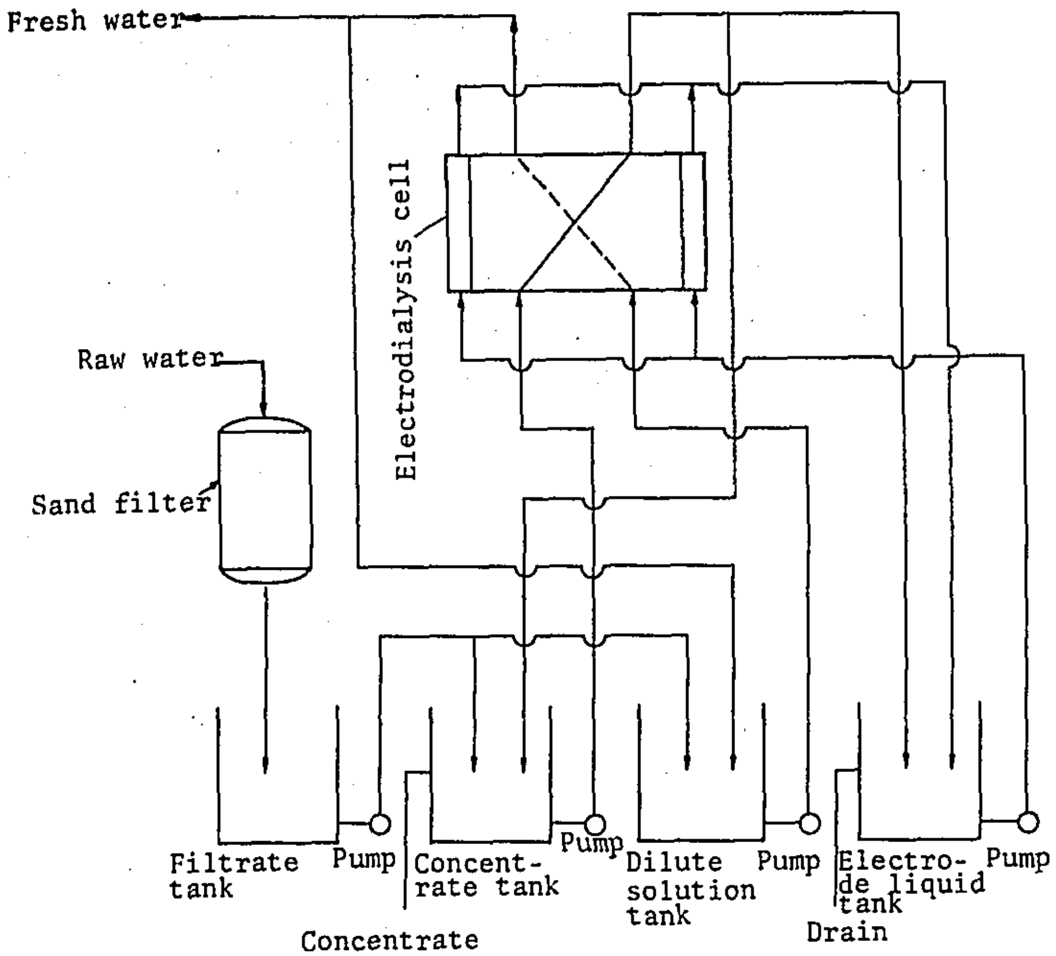
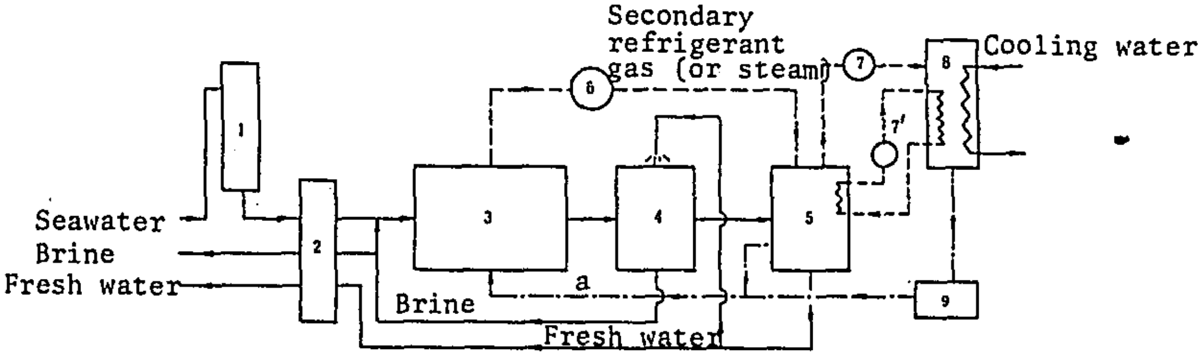


Fig. 3.2-2 Flow sheet of electrodialysis method



Note: a; Secondary refrigerant liquid
1) Deaerator 2) Heat exchanger 3) Crystallizer 4) Wash column 5) Melter 6) Main compressor (Absorption column in case of vapor absorption process) 7) Auxiliary compressor (in case of secondary refrigerant freezing process) 7') Auxiliary refrigerator (vacuum freezing process) 8) Heat exchanger 9) Refrigerant storage tank

Fig. 3.3-1 Basic flow sheet of direct freezing process

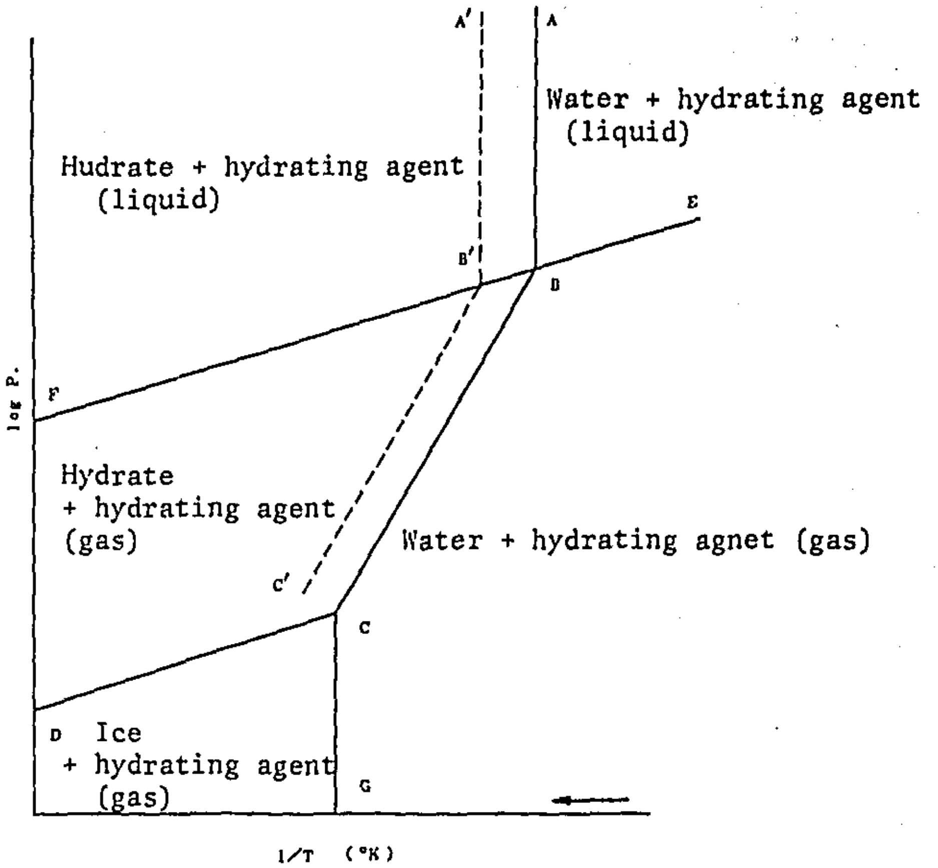


Fig. 3.3-2 Phase diagram of hydrating agnet + water system

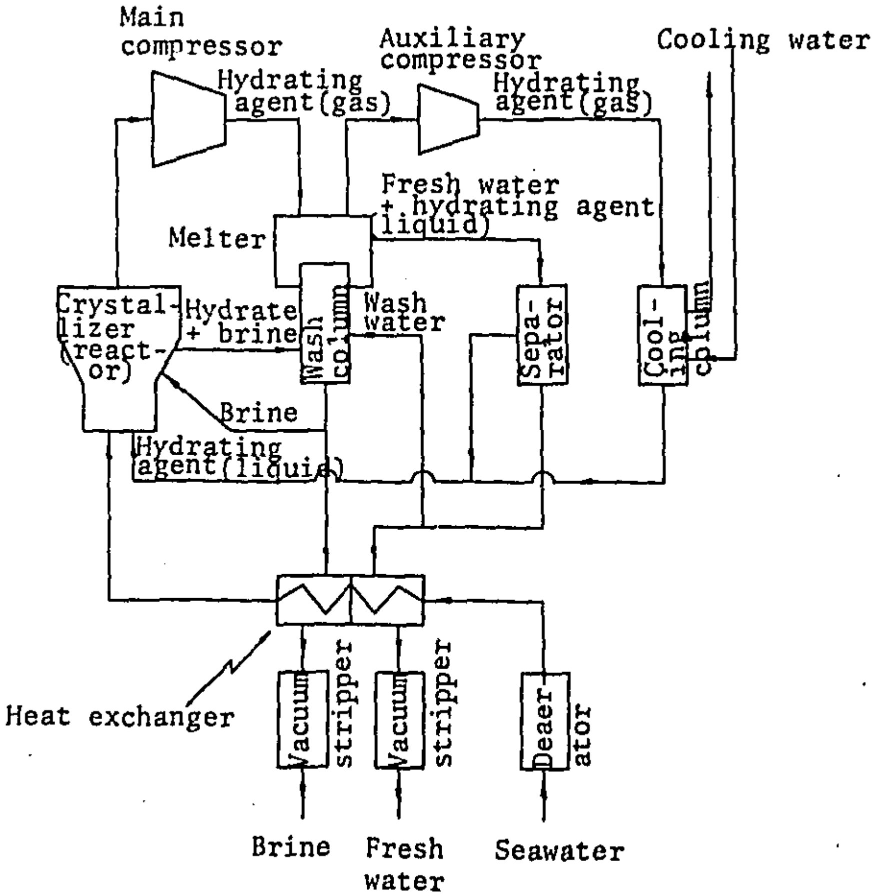


Fig. 3.3-3 Basic flow sheet of desalination plant by hydrate process

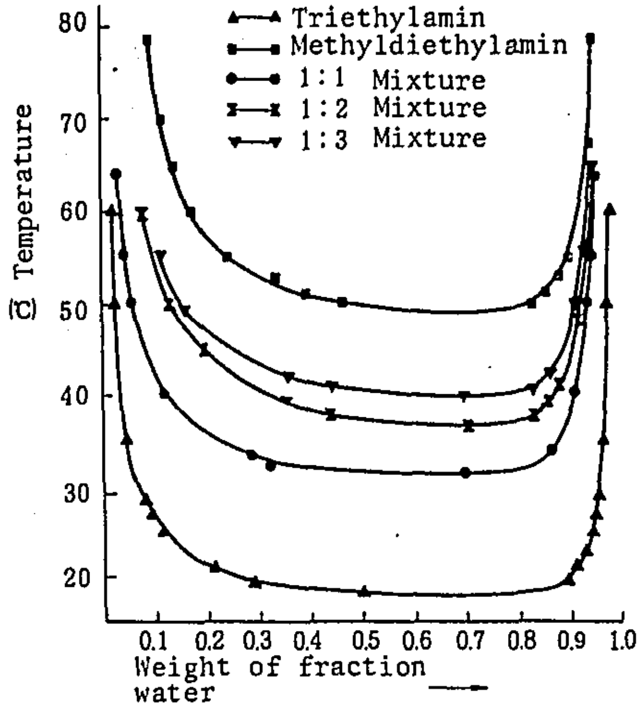
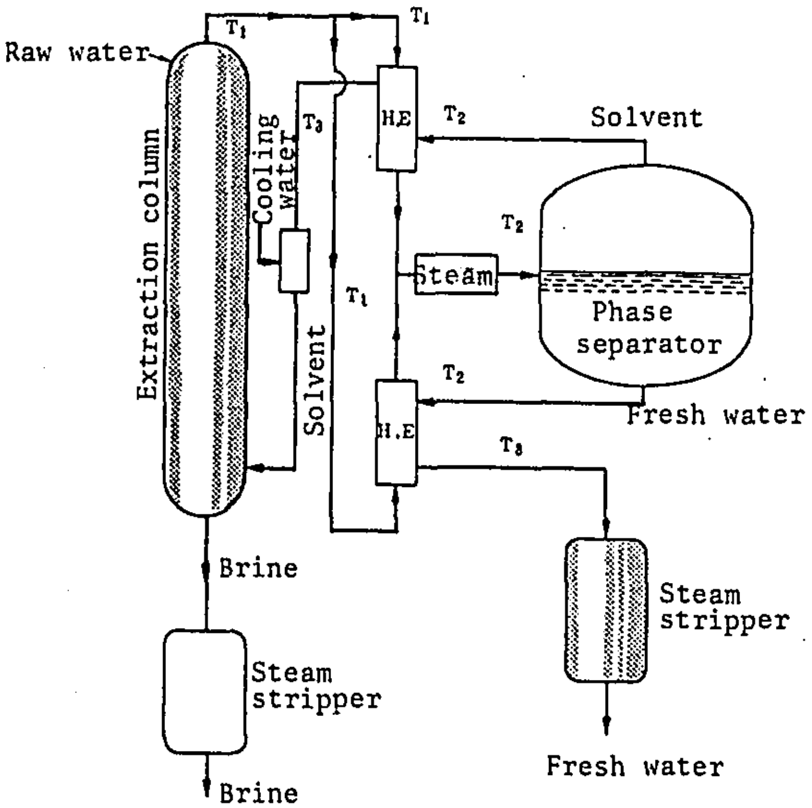


Fig. 3.4-1 Solubility curves for triethylamin, methyldietylamin and mixtures of the two



Note: H.E.; Heat exchanger

Fig. 3.4-2 Flow sheet of desalination plant by extraction process

Table 3.2-1 Property of ion-exchange membranes

Manufacturer	Trade name	Symbol	Reinforcing material		Size (cm × cm)	W.T. (mm)	E.R. (Ω.cm²)	T.N.	B.S. (Kg/cm²)
Asahi Glass	Selemion	CMV	C	PVC	98×150	0.15	3.0	0.91	4
		AMV	A	〃	〃	〃	3.5	0.93	〃
		ASV	SA	〃	〃	〃	4.5	0.95	〃
Asahi Chemical	Aciplex	CK-1	C	Non	112×112	0.23	3.3	0.91	2
		CK-2	SC	〃	〃	〃	〃	〃	〃
		CA-2	A	〃	〃	0.21	2.3	0.98	〃
		CA-3	SA	〃	〃	〃	〃	〃	〃
Tokuyama Soda	Neosepta	CL-25T	C	PVC	100×150	0.17	3.0	0.98	4
		CLS-25T	SC	〃	〃	〃	〃	〃	〃
		AF-4T	A	〃	〃	0.20	2.0	〃	〃
		AFS-4T	SA	〃	〃	〃	3.0	〃	〃
Ionics	Nepton	CR 61 AZL 183	C	Dynell	—	0.60	12.0	0.91	8
		AR 111 BZL 183	A	〃	—	〃	〃	0.96	〃

Note:
W.T.; Wet thickness
E.R.; Effective resistance
T.N.; Transport number
B.S.; Bursting strength

Table 3.3-1 List of desalination test plants by direct freeaing process

Process	Imprementing organization	Production output	Period implemented	References
Vacuum freezing				
Vapor absorption	Carrier Co.	(15×10^3 GPD) 57m³/D	1960~1963	OSW.No113 (1964)
Vapor compression	Colt Industries	(60×10^3 GPD) 2274m³/D	1965~1968	OSW.No295 (1968)
Ejector-absorption	〃	(6×10^3 GPD) 228m³/D	1970	OSW.No744 (1971)
Secondary refrigerant freezing Butane	National Chemical Laboratory for Industry	100m³/D	1955~1959	Chemical Engineering 27, 218 (1963)
〃	Struthers Co.	(15×10^3 GPD) 57m³/D	1963~1968	OSW.No416 (1968)
〃	Mitsui Shipbuilding Co.,	45m³/D	1972~	Bulletion of the Society of Sea Water Science, Japan
〃	Simon-Courves (UKAEA)	(10×10^3 GPD) 45m³/D	1963~1972	The 4th International Symposium (1973) Vol.3, P.281
〃	Blow-Knox	(35×10^3 GPD) 133m³/D	1960~1964	Saline Water Conversion Report for 1964 (OSW)
C-318 (C_4F_8)	Corrier Co.	(37.5×10^3 GPD) 142m³/D	Design 1964	OSW.No256 (1967)
R-114 ($CC\ell F_2-CC\ell F_2$)	AVCO.Co.	(75×10^3 GPD) 285m³/D	1973~1976	The 4th International Symposium (1973) Vol.3, P.371

Table 3.3-2 LNG Consumption and Electric Power Consumption Desalination Plant by Butane Freezing Process with LNG as Cold Thermal Source
(Production output: 1,000 m^3/D)

Use of compressor	LNG evaporator located at:	LNG consumption (t/m^3 - fresh water)	Power consumption (kWh/m^3 - fresh water)
Main compressor and auxiliary compressor (conventional type)	None	0	11 ∿ 12
Main compressor only	Auxiliary compressor	0.12 ∿ 0.13	8 ∿ 9
None	Main compressor and auxiliary compressor	0.5 ∿ 0.6	3 ∿ 4

Table 3.3-3 Properties of hydrating agents and hydrate

Hydrating agents	Molecular weight	Boiling point (℃)	decomposition point Tc(℃)	Pc(mmHg)	Solubility(Wt%) (Tc, Pc)	Heat of formation (kcal/mole)	Hydrate composition Moles H_2O / Moles agent	Hydrate crystal structure
Ethane	30.068	−89.0	14.5	$>10^4$	−	16.3	5¾·H_2O	I
R−22 ($CHClF_2$)	86.483	−40.4	17.8	6,452	0.75 (5℃)	20.3	7⅔·H_2O	II
CH_3Br	94.95	− 3.56	14.7	1,151	1.85	19.5	〃	〃
CH_3Cl	50.491	−24.2	21.0	3,800	3.00	18.1	〃	〃
Propane	44.094	−42	5.7	4,141	0.06	32	17·H_2O	I
R−21 ($CHCl_2F$)	102.933	8.92	8.68	760	1.82	32.7	〃	〃
R−11 (CCl_3F)	137.369	23.8	8.0	419	0.11 (27℃)	29~33	〃	〃
R−12 (CCl_2F_2)	120.924	−29.2	12.1	3,435	0.16	32.9	〃	〃

IV. Saline Water Conversion Using the Evaporation Process

The evaporation process has developed with the growth of chemical or process industries including the salt manufacturing industry, sugar refining industry, pulp industry and soda manufacturing industry and it has made a remarkable improvement in the operating efficiency and production cost in recent years with the progress of systematization of engineering, especially in the field of chemical engineering. It can be said, therefore, that improvement of equipment design and operating methods suitable to the type of materials handled and the characteristics of the process was made after much trial and error in the initial stage, but is supported by engineering analysis these days.

Historically, the modern evaporation process made its shart with the single effect evaporator, in which boiler steam was used as a heat source and the steam generated from the boiling fluid was condensed for concentration of the object solution or for production of solid crystals. The next type was the multiple effect evaporator which used the steam generated in an evaporator as a heat source for subsequent evaporators which were kept at a lower pressure. This multiple effect evaporator is now being used in many of the industries mentioned above. For saline water conversion, the single effect evaporator or submerged tube evaporator was used first but the multiple effect evaporator is now employed.

The evaporator, on the other hand, started with a type similer to an iron pot, which is still used for solidification of electrolytic caustic soda, changed to the calandria type, which uses a group of vertical heat transfer tubes to improve the rate of heat transmission and finally developed to the present evaporator system, which uses various combinations of length and diameter of vertical tubes, horizontal tubes and diagonal tubes according to the characteristics of scale deposition, thermal properties, corrosion and viscosity of the solution to be treated for concentration or crystallization.

If the steam generated in the evaporator is to be reused as a heat source, the steam must be condensed on the external or internal surface of calandria tubes at temperatures higher than boiling temperature or heating temperature inside or outside the heating tube. The method which compresses the generated steam to a specified level in a single evaporator for use as its heat source is the vapor compression process. This process was used first by a beet sugar refinery in Europe and had been used in Japan until the three salt manufacturing plants, from seawater, each having a capacity of 30,000 tons per year, shifted their process to electrodialysis process. Naturally, application of this process to saline water conversion is also conceivable.

As compared with the above two evaporation processes, the multistage flash evaporation method has been developed only in recent years for application to saline water conversion.

These three evaporation processes were adopted by the Office of Saline Water (O.S.W.), Department of the Interior of the United States, for saline water conversion under its overall development program. As a result, a demonstration plant having a capacity of I MGD (3,800 m^3/day) was constructed and operated in Free Port, Texas for experiments on multiple effect evaporation, in Roswell, New Mexico for experiments on double effect-vapor compression evaporation and in San Diego, California for two experiments on multistage flash evaporation. These demonstrations provided valuable data, which are summerized in the Research & Development Progress Report of O.S.W. Later, O.S.W. was merged into the Office of Water Research & Technology (O.W.R.T.) of the Department of the Interior but it still carries out its original function.

While research and development was carried on, the technology of saline water conversion of the world has come to be developed, and by January 1975, a total of 1036 saline water conversion plants, each having a capacity of 94.6 m^3/day or over, were in operation or under construction for a target production of 1,990,000 m^3/day of fresh water, of which the plants using the evaporation process account for 664 or 64% in number and 1,690,000 m^3/day or 85% in the production of fresh water.

Of the plants using the evaporation process, those using the multistage flash method account for 285 or 42.9% in number and 1,332,000 m^3/day or 78.7% in the production of fresh water, with the average production rate per plant being 4,674 m^3/day, the largest figure among the plants using the evaporation process, while those using the vertical multiple effect process account for 104 in number and 193,000 m^3/day or 9.7% in the production of fresh water, those using the horizontal multiple effect process account for 12 in number and 19,000 m^2/day or 1% in the production of fresh water, and those using the submerged tube evaporator, which were constructed in the early stages, account for 138 in number and 87,000 m^3/day or 4.4% in the production of fresh water. The evaporation process, which made its start with the submerged tube evaporator was improved greatly in thermal efficiency with the adoption of the multiple effect process and has since been changed to the present multistage flash evaporator in nearly all cases because of the characteristics of scale deposition and for economical reasons in scaling up the operation. A great number of large scale plants, for operation with low level heat sources, have been constructed with a huge amount of oil dollars, especially in countries of the Middle East, to cope with a sharp increase of water demand for industrialization and improvement of the standard of living. All of these plants adopt the multistage flash evaporation process. In this chapter, the engineering aspect, which is the basis of the entire evaporation process and the process flow of multiple effect, vapor compression, and multistage flash evaporation processes and of a combination of these processe will be discussed. Saline water conversion using solar heat will also be discussed.

4.1. Evaporator and Its Operation

The standard evaporator, which may be called a prototype evaporator used for the evaporation process, is shown in Fig. 4.1-1. The evaporator consists of a calandria which condenses steam, a heat source, and transmits the heat retained by the steam to evaporator fluid through heating tubes; a vapor chamber in which the generated steam and vessel fluid are separated from each other; an eliminator which separates vessel fluid splashes from the steam flow; a steam discharge pipe; a steam condenser (installed separately); a heating steam supply pipe running to the calandria; a vent which discharges non-condensed gas contained in the heating steam and a drain discharge pipe. Provision of a device to prevent and remove scale deposition is required regardless of the type of evaporator and the method of process. Operation of the evaporator used for saline water conversion is described below.

4.1.1. Deposition of Scale and Its Prevention

When the vessel fluid is heated and concentrated, certain components of the fluid separate out according to their solubility characteristics and coprecipitate solid fine grains in the evaporator under certain circumstances. Very often, these particles deposited on the surface of the heat transfer tube form a scale, which provides resistance against heat transfer or causes a flow resistance, eventually leading to the plugging of the piping system. It is essential, therefore, to provide a means of preventing and removing the deposition of scale when the evaporation process is adopted.

In the evaporation process of seawater, in particular, heating and concentration of HCO_3^-, Mg^{++}, Ca^{++} among the components shown in Table 4.1-1 inevitably lead to the precipitation of $CaCO_3$ and $Mg(OH)_2$ through reactions shown in Fig. 4.1-2 and causes scale deposition of the $CaSO_4$ system through reaction (5) at higher concentrations.

In other words, HCO_3^-, when heated, decomposes (1) and generates CO_2 and CO_3^{--}, the reaction (2) precipitates $CaCO_3$ and the OH^- generated by the reaction (3) and the reaction (4) cause the deposition of $Mg(OH)_2$ scale. The concentration of CO_3^{--} and OH^- is determined by the solubility product of $CaCO_3$ and $Mg(OH)_2$. In the case of water at 25°C, the concentrations at equilibrium are given by formulae (2) and (4) in the figure, but slightly different values are obtained depending on the temperature and the concentration of solutions. In the case of water below 82°C, the reaction (2) is predominant and for water above 82°C, the reactions (3) and (4) are predominant. Fig. 4.1-3 shows the limiting concentration of $CaCO_3$ and $Mg(OH)_2$ concerning pH and temperature, and from both values, pH shown by the axis of abscissas and the concentrations shown by the axis of ordinates, whether any solutions are lower than the solubility with regard to $Mg(OH)_2$ can be determined. The solubility of $CaCO_3$ is a function of the total alkalinity expressed as $CaCO_3$. The dotted line is a solubility curve of plotted concentration and saturated total alkalinity for untreated concentrate of seawater without the loss of alkalinity. With the untreated seawater, at least $CaCO_3$ separates out at pH 8 and $Mg(OH)_2$ separates out at a temperature above 80°C regardless of the concentration.

Hard scale of the calcium sulfate family takes the form of anhydrate salt ($CaSO_4$), hemihydrate salt ($CaSO_4 \cdot 0.5\ H_2O$) and dihydrate salt ($CaSO_4 \cdot 2H_2O$) depending on the temperature and concentration as shown in Fig. 4.1-4. Symbols shown in the figure are 67 pieces of data plotted by O.S.W., which indicates that the scale deposition does not occur at levels close to the saturation line of $CaSO_4 \cdot 0.5\ H_2O$ in the multistage flash evaporator, while the scale deposition does occur at the level above the saturation line of $CaSO_4$ in other evaporators and that considerable stable supersaturation may be maintained for $CaSO_4$ in the M.S.F. process. Contrary to the soft (alkaliner) scale, $CaCO_3$ and $Mg(OH)_2$, which can be easily removed with acid, the hard scale of the $CaSO_4$ family, once deposited, requires suspension of plant operation for removal. It is important, therefore, to develop a means to avoid this deposition in designing the process. In the case of the multistage flash process, the operation is limited to a temperature below 121°C and concentration below 2 at present.

For prevention or removal of soft scale such means as mechanical or chemical means, seeds sludge recirculation of scale components, ion exchange, addition of a scale inhibitor and pH control are available. For a large plant, the last two methods are employed.

Hagevap is a mixture of a compound of Sodium polyphosphate - Lignin sulfonate and ester of polyalkyl glycol, which is effective for prevention of scale deposition when added in the amount of 2 to 4 ppm at temperatures up to 85°C and is being used in desalination plants. Besides, new scale prevention agents or inhibitors of various types have been developed in many countries.

For prevention of scale deposition in large plants, such means as stripping of CO_2 and pH control of make-up seawater and pretreatment of raw water centering on vacuum deaeration for removal of dissolved oxygen, which causes corrosion of materials, have been studied for processes using the multistage flash evaporation process in the National Research and Development Program, "Seawater Desalting and By-Product Recovery".

In the reaction shown in Fig. 4.1-2, a process which adds H_2SO_4 to evaporate the carbonic acid content to the air as CO_2 has been studied, using packed column, perforated plate tower and cascade systems.
A special perforated plate tower has been developed for use in deaeration. In addition, development researches have been made on inhibitors and the intermittent wiping process using sponge balls and on the material of balls. Obata et al. have conducted bench scale experiments to determine the level of vacuum and the capacity of deaeration towers. They have found that, in the case of a packed tower, the attained dissolved O_2 concentration drops drastically when the operating temperature at the operating pressure comes close to the saturation temperature at a gas velocity of G_{ss} = 378 - 529 [kg/m^2·hr] and at liquid velocity of L = 38.7 [t/m^2·hr], and that there is a decrease of $(H.T.U.)_{OL}$, which may be called the reciprocal of the capability of the tower. In other words, it has been confirmed that the desoration of dissolved oxygen is accelerated suddently when the phase changes from the liquid phase to the vapor phase as in the case of flash evaporation. This suggests the advantage of deaeration in M.S.F. plants.

4.1.2. Heat Transfer Tubes and Heat Transfer Rate

In the saline water conversion plant using the evaporation process, the cost of materials resistant against seawater corrosion takes a large share in the total construction cost, and for this reason, the discovery of materials of low cost and the improvement of the heat transfer rate are major subjects of development research.

The heat transmission rate q[Kcal/h] is given by the following;

$$q = UA(T-t) \cdots\cdots (1)$$

where U: Overall coefficient of heat transmission [Kcal/m^2·hr·°C)

A: Area of heat transfer [m^2]

T-t: Temperature difference of heat transfer [°C]

T: Temperature of heating fluid [°C]

t: Temperature of fluid being heated [°C]

In the model of heat transfer between two fluids separated by a heat transfer tube, the heat transfer resistances related to the tube, two films of fluid and layers of scale on both sides of the tube, shown in Fig. 4.1-6, are added together like electric resistance in series and the total of this resistance is expressed as 1/U. The thin layer of fluid of a certain thickness adjacent to the tube wall is called a film, in which only laminar flow parallel to the tube wall occurs and flow perpendicular to the tube wall does not occur, with no convection heat transfer. If the radiation is ignored, heat transfer occurs only as a result of heat conduction through the film. The thickness of the films changes according to the flow condition on both sides. A film of water formed by condensation of steam on the heating side in the case of saline water conversion using the evaporation process, a film of heat receiving seawater or concentrated seawater in the case of the flash evaporator or in the case of the evaporator which avoids boiling and a film of two phase flow of vapor and liquid, produced as a result of boiling in the conventional evaporator, function as heat transfer resistors. The reciprocal of this resistance is called the film heat transfer coefficient (refer to Fig. 4.1-6).

Resistances related to the steam side film, the heat transfer tube, two soil layers and on the liquid side film are $1/hs$, x/λ (x is the wall thickness of tube in m, λ is heat conductivity of tube in Kcal/m·hr·°C), $1/h_{SD}$, $1/h_{LD}$ and $1/h_L$, respectively.

Since i/h_{SD} can be ignored normally, the total heat transfer resistance, $1/U$, is as shown below, if the area on both sides and the average area are corrected to A_1, A_2 and A_{av} [m^2], respectively.

$$q = \frac{T - t}{\frac{1}{A_1 h_S} + \frac{x}{A_{av} \lambda} + \frac{1}{A_2 h_{DL}} + \frac{1}{A_2 h_L}} \quad \cdots\cdots\cdots\cdots \quad (2)$$

If the area is corrected with inside and outside diameters of the tube and the average diameter taken as D_1, D_2 and D_{av}, respectively,

$$\frac{1}{U_1} = \frac{1}{hs} + \frac{x}{\lambda}\left(\frac{D_1}{D_{av}}\right) + \frac{1}{h_L}\left(\frac{D_1}{D_2}\right) \quad \cdots\cdots\cdots\cdots \quad (3)$$

A_{av} and D_{av} are the logarithmic mean of A_1, A_2 and D_1, D_2, respectively, and the difference between the logarithmic mean and the arithmetic mean is less than 4% if the ratio of each is less than 2.

A profile of condensation heat transfer is shown in Fig. 4.1-7, in which the distance between the two temperature lines is the driving force of heat transfer. Since the temperature is normally measured at both ends of temperature lines, the temperature difference between the two ends of the line, the log-mean temperature difference of ΔT_1 and ΔT_2 or $\Delta t1$, and Δt_2 is taken and the arithmetic mean is taken if the ratio is less than 2. Since the transfer of latent heat of condensation occurs when the steam reaches saturation temperature and the specific heat of steam and condensed water is far below the latent heat, the mean value of the difference between the saturation temperature of steam and the heat receiving temperature should be taken as the actual temperature difference to be exact. Therefore, care should be taken in diverting the data of U when the temperature profiles are not considered to be identical.

To reduce the required heat transfer area, A, it is necessary to minimize 1/U or each heat transfer resistance in the equation (3). In order to determine the extent of the effect of these heat transfer resistances on 1/U or the total temperature difference, the values calculated from the theoretical formula and empirical formula were applied, with the data obtained from test plants of O.S.W., as shown in Fig. 4.1-8.

The axis of ordinates in the figure represents heat transfer resistance expressed as a heat transfer temperature difference and the distance between curves corresponds to each resistance, though not shown in the order of their magnitude. The resistance of films of steam and liquid ($1/h_s$, $1/h_L$) and heat transfer tubes, x/λ, was obtained from the equation shown in Chapter 5 and the boiling point raise (B.P.R.) and the non-equilibrium temperature difference (N.E.T.D.), which will be described at a later stage, were obtained from the reported equations. While the latter two values may not necessarily be accurate values, the former three values should be reasonably accurate.

Tendencies of these five values are that B.P.R. of case (1) is about 1°C against the total temperature difference of about 4°C, which is not adjustable except for concentration as far as evaporation of seawater is concerned, that the total of $1/h_L$, x/λ, $1/h_s$ of cases (3), (4), (5) is approximately 2°C, the decrease of which is not at all impossible but is rather related with other operating conditions, and that N.E.T.D. of case (2) and the resistance due to non-condensate gas and scale of case (6) can be reduced. Especially the scale in case (6), the extent of which has not been determined, can be reduced through complete removal, and N.E.T.D. of case (2), through its empirical formula is not accurate, can be reduced through appropriate operations.

For B.P.R., the following two equations were used in the aforementioned study.

The equation sought by Nakajima et al. from the measured value obtained by Uchida et al.:

$$B.P.R.=C^{1.178}(0.1120+0.180\times10^{-2}\times t)$$

The equation approximately by Nakajima et al. from the equation developed by Stoughton:

$$B.P.R.=C^{1.168}(0.0550+0.00065t)$$

where C: Concentration of salt (weight %)
t: Boiling point of water (°C)

The boiling point rise (B.P.R.) of a solution is due to the presence of solute. When salts are present, the solution boils at a temperature higher than the boiling point of water under the same pressure and the steam thus generated does not condense till it comes down to the saturation temperature of steam (boiling point of water under that pressure). It is important, therefore, to know the B.P.R. In the case of a conventional evaporator in which boiling occurs on the heat receiving side, there is another B.P.R. due to the increase of static pressure according to the depth of liquid and the temperature on the heat receiving side exceeds the boiling point of water at the pressure of the vapor chamber of the evaporator by the sum of these two values of B.P.R. Normally, the value of B.P.R. for the average depth of solution is added to the value of B.P.R. owing to the pressure of solute.

For analysis of operation data of an evaporation plant, the difference of saturation temperature between the heating side and the heat receiving side at the pressure of both chambers is taken as a apparent heat transfer driving force (Δt_a), and then the apparent overall heat transfer coefficient, U_a, obtained from the heat transfer equation $U_a = q/A \cdot \Delta t_a$, is often adopted. For use of U_a for other evaporators, however, these values of B.P.R. should be given due consideration.

While there are many reports on B.P.R. of seawater, the data on the change of concentration of salts through heating and evaporation and the reliability of measuring method vary from one report to another. Tawara and Hakuta summarised physical properties of seawater and brine and Hakuta et al. prepared a table of values of measurement (Table 4.1-2) taken with a twin ebulliometer and a graph showing the relationship between the temperature, pressure and B.P.R. (Fig. 4.1-9). The formula used by them is as follows.

$$\begin{aligned}\text{B.P.R.} =\ & 0.528764\times10^{-1}\,x+0.826030\times10^{-5}\times xT\\ & -0.315082\times10^{-7}\times xT^{2}+0.320553\times10^{-2}\times x^{2}\\ & -0.144367\times10^{-4}\times x^{2}T+0.184416\times10^{-6}\times x^{2}T^{2}\end{aligned}$$

As shown by the example of the multistage flash evaporation process, the effect of B.P.R. increases the area of heat transfer, and for this reason, an accurate calculation must be made when a precise cost estimate is required. The above-mentioned non equilibrium temperature difference (N.E.T.D.) often makes evaporating seawater fail to attain the saturation temperature (temperature with the addition of B.P.R., to be exact) at the pressure of the stage depending on flow velocity and temperature in the flash chamber of the multistage flash evaporator and as a result, the insufficiently evaporated and superheated brine flows out to the next stage. The difference between the temperature at the outflow and saturation temperature plus B.P.R. is called the non-equilibrium temperature difference (N.E.T.D.). N.E.T.D. in the case of flash evaporation is described in detail in 5.3.1.

The value x/λ can be decreased through an increase of λ and a decrease of x or through use of a thin-wall heat transfer tube made of materials of high heat conductivity. However, the type of materials, which can be used for this purpose, is limited because of the requirement for corrosion resistance, mechanical strength and corrosion margin.

It was for this reason that experiments on Ti 0.3 mm thin-wall tubes were carried out to determine its corrosion resistance and mechanical strength under this R & D project. The result of such experiments is highly evaluated. While h_s, h_L will be taken up when discussing the equation in section (5.3.2) dealing with heat exchange in relation to the multistage flash evaporation process, the problem is the decrease of h_s due to the presence of non condensate gas and the increase of the number of tubes in large plants. The value of h_L in the case of no boiling of the heated solution can be calculated in the same manner as that for other evaporators and heat exchangers.

Since the value of h_L in the case of boiling of the solution varies depending on the type of evaporator and operating conditions of the evaporator, i.e. the forced and natural circulation and non-circulation, equations which make calculation possible in general have not yet been established. Therefore, only the vertical falling film type long tube evaporator, on which QSW has concentrated their development efforts, will be discussed here. From Keville's milk concentration experiment, the following may be established as an equation between nondimensional numbers such as the modified Reynolds number $Re(=\frac{Du_v \rho_L}{\mu_L})$, Prattle numbers $Pr(=\frac{Cp \mu_L}{\lambda_L})$, and $(\frac{P_L}{\sigma_L})$.

When Re > 7,000

$$\frac{h_L D}{\lambda_L} = 5.40 \times 10^{-3} (Re)^{0.5} (Pr)^{0.3} (\frac{PL}{\sigma})^{0.9}$$

when Re < 7,000

$$\frac{h_L D}{\lambda_L} = 1.98 \times 10^{-3} (Re)^{0.33} (Pr)^{0.3} (\frac{PL}{\sigma})^{0.9}$$

In the above, ρ, μ, λ, σ, Cp and density (kg/m^3), viscosity (kg/m·sec.), heat conductivity (Kcal/m·hr·°C), surface tension (kg/m), and specific heat (Kcal/kg·°C), respectively and D means inside diameter of tube (m), L is the length of tube (m), P is the pressure (kg/m^2), unit kg is the force, kg is the mass, and inferior figures L and v are liquid and vapor, respectively.

Of U, soil coefficient, h_{LD}, requires numerical data but the scale deposition varies greatly depending on the process characteristics such as solubility characteristics of deposition components, composition of corrosion products and other co-precipitating solids, surface properties of the material used for heat transfer tubes, flow velocity inside the tube and tube surface temperature, and for this reason, no generalized expression can be made in providing measures against scale deposition, even when the value of h_{LD} may be very small. Normally, the most economical washing cycle is determined with the change of h_{LD} or U against operating hours used as data for calculation. The result obtained under this project is described in detail in 5.1.

4.2. Multiple Effect Evaporator

Heat economy has improved greatly by the shift from the immersion tube system, in which steam was conducted through immersed coiled pipes to boil seawater, to the multiple effect evaporator, and the system developed by QSW. in the test plant was a twelve effect plant using long tube vertical evaporators. In this system, the boiling seawater flows down through heating tubes to form a falling film to reduces the resistance of heat transfer. At first, the plant was like the one shown by the flow chart in Fig. 4.2-2, which consisted of two rows of 6 single units shown in Fig. 4.2-1, but latter, an experiment was made again with a 17 effect evaporator system which had an additional five box-type evaporators.

The points shown at the lower left of $CaSO_4$ solubility line at the lower left of Fig. 4.1-4 are the data obtained during this operation, a normal flow operation in which raw seawater was supplied to the high temperature low concentration evaporator and then conducted to the last evaporator where the concentration reached the maximum level. For this process, it was necessary to preheat raw seawater and provide a heat exchanger between each evaporator. It was also necessary to provide a means of final preheating using a small amount of steam to raise the temperature of seawater to the boiling point of the first evaporature of the highest pressure. For generated steam of the lowest pressure, a condenser must be provided separately. Needless to say, a vacuum device must be provided to remove non condensable gas and lower the pressure of the last evaporator as much as possible because there is a limit on the maximum temperature and concentration for prevention of scale deposition.

According to Porteous' equation, which assumes that the amount of seawater supply is 1.5 times the amount of fresh water production and that the latent heat of evaporation and specific heat of seawater, steam and concentrated brine are constant under the condition of each evaporator, the amount of fresh water production M_D (kg/hr) per steam requirement

M_s (kg/hr) of n multiple effect evaporator or the ratio of fresh water production $R_n = M_D/M_s$ may be determined approximately with the following equation.

$$R_n = \frac{2n\gamma}{2\gamma + 3n(\Delta + \alpha)}$$

where γ : Latent heat of evaporation [Kcal/kg]
α : Boiling point raise [°C]
Δ : Temperature rise of preheater in each evaporator [°C]

If $\gamma = 556.7$ Kcal/kg, $\alpha = 0.83$°C, $\Delta = 3.34$°C, then the cost of fresh water production of a 13 multiple effect evaporator is calculated to be 11.34.

According to Porteous' calculation method which uses a daring assumption for calculation of the required heat transfer area of a multiple effect evaporator, the latent heat held by the heating steam M_{SO} (kg/hr) of the first evaporator is transferred to the boiling seawater at the temperature difference of $(\Delta T_X - \alpha)$ [°C] and therefore, the required area S_1 [m^2] is

$$S_1 = \frac{M_S L}{U_e(\Delta T_X - \alpha)} \qquad (5\text{-}14)$$

In the above, U_e is the average overall heat transfer coefficient [$Kcal/m^2 \cdot hr \cdot °C$] in the evaporation section and ΔT_X is the temperature difference of steam between inlet and outlet of each evaporator.

In the second ~ n-th evaporators, the preheat portion of raw seawater $M_F \cdot \Delta T_X$ is deducted as follows.

$$S_{2-n} = (n-1)\left[\frac{M_{SO} \cdot \gamma - M_F \cdot \Delta T_X}{U_e(\Delta T_X - \alpha)}\right] \qquad (5\text{-}15)$$

Then the required area of the raw seawater preheater is

$$\frac{nM_F}{U_n}\log\left[\frac{\Delta + \Delta T_X}{\Delta}\right] \qquad (5\text{-}16)$$

In the above, U_n is the overall heat transfer coefficient [$Kcal/m^2 \cdot hr \cdot °C$] of the n-th preheater.

From the sum of the preceeding 3 equations, the total heat transfer area A_T [m^2] per unit of fresh water production, when the amount of raw seawater M_F is assumed to $(3/2)M_D$, is calculation as follows.

$$\frac{A_T}{M_D}=\frac{n(\gamma-\Delta T_T)}{Un(\Delta T_T-n\alpha)}+\frac{1.5}{U_n}\log\left[\frac{\Delta+(\Delta T_T/n)}{\Delta}\right]$$

In the above, $\Delta T_T = n\Delta T_X$

The test plant shown in Fig. 4.2-2 achieved an operating rate of 74%, annual operating days of 274 MG, and a production rate of 0.88 MGD against the design production of 1 MGD (3.785 m^3/day). The maximum temperature was 127°C and the evaporation temperature of the 12th evaporator was 46.2°C. The problems reported include the deposition of soft scale in the low concentration evaporators, deposition of hard scale on the condensation side of high concentration evaporators due to mist entrainment, corrosion of equipment and uneven distribution of seawater in the falling tubes of calandria. Through such measures as the improvement of preheating operations, lining of the inner surface of evaporators, reform of distribution nozzles on the top of heating tubes and pH control, the operating rate improved to 84% and the fresh water production rate increased to 9.7. An alloy of 90:10 Cu-Ni and 2%-Aℓ brass showed a relatively high corrosion resistance.

A report of O.S.W. states that under the present state of multiple effect plants using falling film vertical evaporators, the largest production plant with a single aim of fresh water production or dual aim of fresh water production and power generation, among the plants of 1 - 250 MGD scale, is a 1 MGD (3,800 m^3/day) plant of 11 multiple effect evaporators with a water production rate of 9.3. The largest water production rate obtained in the experiment is 13.2 for a plant of 17 multiple effect evaporators. The recently studied combined system (ME/MSF), in which a multistage flash evaporator is used as a preheater in the multiple effect plant, is hopeful for improvement of the economy of the plant with respect to the materials of equipment, construction and operation of the plant. The study shows that the enhanced tube contributes to the improvement of evaporation and condensation heat transfer and that the concrete lined steel plate has an improved corrosion resistance for operation over a long period of time. The plant constructed with the present technology has a concentration ratio of 2.5 - 3.0 and a water production rate of 0.75 - 0.85 per effect evaporator and uses 2 or 3 inch-diameter heat transfer tubes made of Aℓ-brass or an alloy of 90:10 Cu-Ni. Further improvement of the process, materials of equipment and manufacturing techniques is expected to lower the cost of saline water conversion.

In this sense, the development of fluted tubes has increased the competitive power of the multiple effect evaporator with vertical tubes. In the falling film evaporator where boiling heat transfer occurs, the boiling seawater and condensed water both inside and outside the ridge section of the flute form thin films by the effect of surface tension, which increases the value of h_L, h_s considerably and constantly dampens the ridge section inside the tube to facilitate uniform flow in the direction of the tube. As a result, the total heat transfer coefficient increases four times to 12,000 [$Kcal/m^2 \cdot hr \cdot °C$] in terms of the standard diameter of a smooth tube. Though the manufacturing cost is said to be higher by 30% than the cost of smooth tubes by, a substantial cut in the cost can be expected. Together with the development of fluted tubes, a tube top seawater distribution method has been developed.

Cox has reported a new attempt to provide grooves inside the tube to improve the rate of heat transfer in the multiple effect evaporator for evaporation outside and condensation inside the horizontal tube. In the flow chart shown in Fig. 4.2-3, which is similar to the flow chart of a two-effect evaporator shown in Fig. 4.3-4, evaporation on the top outside the tube and grooves having dimensions of 0.254 mm x 0.8 mm inside the tube improve evaporation and condensation heat transfer respectively as in the previous case. It is said that the increase of flow velocity of steam inside the tube to 2.45 m/sec. or over to force out the condensed water has resulted in an increase of the value of U to two times the value of a vertical smooth tube.

In the initial stage of development of a multiple effect evaporator by OSW., the relative merits of climbing film and falling film were studied in relation to the operation of long vertical tubes, and development effort was concentrated on the latter thereafter. However, a comparative study of the two for a plant of the 2.5 MGD (9,500 m^3/day) scale was commissioned to a private organization at a later stage. The result of this study shows that in the case of climbing film, which requires a slightly greater heat transfer area but does not require a circulation pump for each evaporator, construction cost can be reduced by 6% and the cost can be reduced further up to a plant size of 10 MGD. The reduction of water production costs, including power costs, is 0.05$ from 1.13$/1,000 gal. The heat transfer tube of the evaporator, which is 2 inches in outside diameter and 10 to 14 ft in length, has flutes made of Aℓ-brass both inside and outside. PH control was used for prevention of scale deposition and the maximum temperature of brine was 127°C.

A comparative study of multiple effect (VTE) and multistage flash (MSF) was also made for saline water conversion plants of 2.5 MGD and 250 MGD scale. In large plants, the 15 effect evaporators - 50 stage (2 stages for heat rejection) flash evaporator link system showed a water production ratio of 12.89, while the system in which the effect evaporator had a concrete floor and a steel ceiling and the flash evaporator had a concrete structure (12 inch) with 1/4" steel plate lining, showed a water production ratio of 12.84 at a maximum temperature of 122°C. In small plants, the water production ratio was 11.6 for a 14 effect evaporator and 11.5 for a 42 stage (3 stages for heat rejection) flash evaporator.

4.3. Vapor Compression Process

The vapor compression process by which the energy equivalent to fractions of latent heat of vapor is utilized to compress the steam generated in the vessel to increase the condensation temperature and recover the heat retained by steam, has been studied or has been in practical use for many years as a very handy method.

The required power used for compression of vapor may be expressed as follows.

$$W=\frac{1}{\eta}\cdot\frac{1}{\eta_{ad}}\cdot W_{ad}\ [\mathrm{Kg}\cdot m/\mathrm{hr}]\cdots\cdots\cdots\cdots\cdots(\ 1\ )$$

$$W_{ad}=\frac{\gamma}{1-\gamma}P_1\ V_1\left[\left(\frac{P_2}{P_1}\right)^{\frac{\gamma-1}{\gamma}}\right]\ [\mathrm{Kg}\cdot m/\mathrm{hr}]\cdots(\ 2\ )$$

where η: mechanical and electrical efficiency [-]

η_{ad}: adiabatic compression efficiency [-]

γ: $C_p/C_v = \frac{\text{specific heat at constant pressure}}{\text{specific heat at constant volume}}$

= (1.333 for 3 atom gas) [-]

P_1, P_2: vapor suction and discharge pressure [kg/m^2]

V_1: amount of vapor at suction [m^3/hr]

Since the condensation temperature is necessary to know the heat transfer rate of the plant, a graphical calculation which uses i-S chart of steam as shown in Fig. 4.3-1 is convenient.

The work of compression required for compression of the vapor at t_I, which has been superheated above the saturation temperature at P_1 by BPR, to P_2 when $\eta_{ad} = 1.0$ and $\eta_{ad} < 1.0$, are respectively $(i_2 - i_1)$, $(i_2' - i_1)$ (Kcal/kg) and $i_2' - i_1 = (i_2 - i_1)/\eta_{ad}$. Since the compressed vapor condenses at its saturation temperature of the compressed press, T_2, the effective heat transfer temperature difference is $(T_2 - t_1)$.

The required compression power, when the amount of evaporation is taken as E [kg/hr], may be expressed as follows.

$$W=\frac{1}{\eta}(i_2{}'-i_1)E \qquad [\mathrm{kcal/hr}]$$

$$=\frac{1}{\eta}\cdot\frac{1}{\eta_{ad}}(i_2-i_1)E \qquad [\mathrm{kcal/hr}]$$

$$=\frac{1}{\eta}\frac{1}{\eta_{ad}}(i_2-i_1)E\frac{1}{860} \qquad [\mathrm{kW}]$$

The value of η_{ad} is 0.75 for a turbo compressor and about 0.85 for an axial flow compressor.

The present National Chemical Laboratory for Industry conducted pilot tests on the process for concentration of pretreated seawater for industrial use with a plant of the 100 m^3/day scale using vapor compression evaporation in fiscal 1951 through 1953 in a series of process development researches for comprehensive use of seawater.

A flow chart of the process used then is shown in Fig. 4.3-2. In this process which used a natural circulation type climbing film long tube vertical evaporator and an axial flow compressor, which was the first unit developed for this purpose, the vapor at a pressure of 1.1 [kg/cm^2· absolute] generated in the evaporator was compressed to a compression ratio of 1.6 and the pretreated seawater was concentrated to 190 (g-NaCℓ/ℓ]. The optimum effective temperature difference across the tube on the economic base then was approximately 6°C. The plant was equipped with an evaporator which had an evaporation heat transfer area of 146 m^2, comprising heat transfer tubes each having an outside diameter of 50 mm and a length of 6 m, and a compressor having revolutions of 10,000 rpm or 200 HP, with approximately 85% adiabatic compression efficiency. The plant intrumentation was designed for one-man operation and the plant was operated with priority given to safety. The water production ratio to power consumption was approximately 16. Seawater preheaters which recovered the heat of condensed water and concentrated brine were of the counter flow type and had a heating area of 74.6 m^2 and 17.6 m^2, respectively.

The energy loss included a mechanical and electrical loss of 13.5%, an external loss of 31.2%, a preheater terminal loss of 28.5% and a leakage loss of 23.2%.

The process adopted for the No.4 demonstration plant of OSW. of the Department of the Interior was that the vapor generated in the second effect of the two evaporators was compressed for use as heating steam of the first vessel as shown in Fig. 4.3-3. In this process, underground brine having a total salt concentration of 15,000 ppm (Ca 582, Mg 162, SO_4 1,500, SiO_2 15 ppm) was conducted through an ion exchange resin column to remove Ca^{++} and then deaerated to prevent scale deposition and corrosion of equipment. The plant size was I MGD (3,800 m^3/day) and an axilial flow compressor was used as in the preceding case. The required power for compression from 1.07 [kg/cm^2 absolute] to 1.3 [kg/cm^2 absolute] was 2,500 HP. The evaporator was a vertical long tube evaporator but was of a type which does not cause boiling in the heating tube.

This plant was operated from 1963 to 1968, during which time many problems were encountered. To solve these problems, a small test plant was constructed and steady operation was achieved after necessary measures were taken but the plant failed to maintain stable operation over a long period of time. However, the energy requirement for water production in normal operation was 16.8 [$kW{\cdot}hr/m^3$], which is quite high in plant efficiency.

The most serious problems were scale deposition, corrosion and mist entrainment. These problems, in correlation with each other, contributed to the decrease of the number of operating days. From the experience with this demonstration plant, the following were suggested.

1) Complete provision of a compressor, a mist eliminator for prevention of corrosion of the evaporator body and a monitor of dissolved oxygen concentration; 2) maintenance of an appropriate and uniform flow velocity in the tube and optimum temperature of calandria for prevention of scale deposition and sufficient depth of solution to prevent boiling inside the tube; 3) introduction of a plate heat exchanger made of Ti for prevention of scale deposition and corrosion and 4) adoption of a sludge recirculation process, as the scale problem of SiO_2 and oxides of Ca^{++}, Mg^{++}, Fe^{++} cannot be solved through pH control.

Depending on the composition and concentration, underground brackish water was not less serious than seawater in the magnitude of scale deposition and corrosion, and the difficulty of experiment with a plant of the 3,800 m^3/day size using the technology, which was successfully applied to the operation of a small test plant, is an indication of the difficulty for practical application of the method because of such problems as non-uniformity of flow velocity in the tubes, and flow of the heating steam in calandria and the increase of cost with the adoption of sludge recirculation required for prevention of scale deposition.

A. Porteous says that a saline water distillation plant similar to that of OSW shown in Fig. 4.3-4 is now in operation, which requires 11.1 kWh/m^3 to increase the temperature by 5.5 - 6.7°C. The evaporator used in this plant is a horizontal tube evaporator, which feeds steam into the heat transfer tubes and forms a film of seawater outside the tube to increase the heat transfer coefficient. However, the plant size is said to be restricted to 4.5 - 90 m^3/day because of the high cost of compressors and preheaters.

In Japan, a stack type vapor compression plant has been developed, in which a three - six effect evaporator, incorporating evaporators similar to the previously mentioned evaporator stacked vertically, is driven by a steam ejector. Plants of this type have been manufactured for shipment to such arid areas as the Middle East for production of water for use in construction work.

4.4. Multistage Flash Process

A flow chart of this process and a temperature curve are shown in Fig. 4.4-1. However, the application of this process to saline water conversion has the following restrictions.

1) Pretreatment of make-up seawater is required for prevention of the formation of soft scale [$CaCO_3$, $Mg(OH)_2$].

2) Though the precess is advantageous as compared with other processes in respect of hard scale deposit as it is possible to heat concentrated seawater inside the heat transfer tube under supersaturation conditions as shown in Fig. 4.1-3, there is naturally a limit to maximum temperature and concentration.

   At present, the limit of the concentration ratio (concentration of evaporator brine/(concentration of raw seawater) is less than 2 and the limit of the temperature is 120°C.

3) In the relation (amount of evaporation) x (evaporation latent heat) = amount of heat transfer ≑ (range of flash temperature) x (flow of flashing seawater) x (specific heat of seawater), the minimum temperature is higher than the temperature of raw seawater, while the maximum temperature is restricted (item 1 above) and the evaporation latent heat is more than 500 times the specific heat of seawater. For this reason, the amount of flashing seawater flow must be greater than the amount of evaporation by several times.

4) A vacuum device is required to hold the pressure of the lowest pressure stage to a minimum so as to increase the range of flash temperature. Moreover, heat must be rejected at low temperatures for maintenance of overall heat balance and a large quantity of seawater which has be preheated to some extent by heat transfer tube of the heat rejector must be discharged.

   When the economy of the process and plant is taken into consideration for design under these restrictions, the reduction of cost of materials used for heat transfer tube and vessel body, cost of pretreatment and power, as well as specific steam consumption and unit cost of steam, from the optimum cost of saline water conversion shown in Fig. 4.4-2 will be a prerequisite for overall cost reduction. The illustration is based on the economic framework of 1969 and has become almost meaningless after the oil crisis with respect to absolute values but may be used for reference of relative value.

Various design methods and analytical works for flash evaporation plants have been reported by Frankel, Steiner, Brice et al. Though somewhat different conclusions have been obtained depending on the method used for simplification, the approximate calculation used by Höffer is quoted in the following.

When the number of flash stages of the heat recovery section is N, the following assumptions may be established.

1) If the condensation temperature of steam in the arbitrarily chosen n-th stage is taken as Tn [°C] and the temperature of preheated circulating seawater coming out the condenser of the same stage as t_n [°C], the difference is considered as constant for each stage.

$$T_n - t_n = \varDelta t = \text{constant} \qquad (1)$$

2) The enthalpy of circulating seawater is constant and is equal to the enthalpy of pure water

3) The boiling point raise of the circulation fluid can be ignored. The pressure loss in the steam path of each stage can also be ignored.

If the calculation is simplified under these assumptions, the boiling point of circulating fluid in n-th stage can be obtained from the following equation.

$$T_n = [(N+1-n)T_0 + n(t_F + \varDelta t)]/(N+1) \qquad (2)$$

If Δt is assumed, the preheating temperature in the n-th stage can be obtained from equations (1) and (2).

Therefore, the temperature drop between the n-th stage and the (n-1) stage may be expressed as follows.

$$T_{n-1} - T_n = \Delta\theta = (T_0 - t_F - \Delta t)/(N+1) \qquad (3)$$

In the above, T_0 is the temperature of circulating seawater entering the first stage and T_F is the temperature at the inlet of the heat recovery section.

When the flow rate of circulation is taken as F [kg/hr] and the evaporation latent heat as α, the amount of condensed water (amount of evaporation) D [kg/hr] obtained from the entire process is as follows:

$$D = \sum_{n=1}^{n=N} D_n = F\Delta\theta \left\{ \frac{1}{\alpha_1} + \left(1 - \frac{\Delta\theta}{\alpha_1}\right)\frac{1}{\alpha_2} + \left(1 - \frac{\Delta\theta}{\alpha_1}\right)\left(1 - \frac{\Delta\theta}{\alpha_2}\right)\frac{1}{\alpha_3} + \cdots \right.$$
$$\left. + \left(1 - \frac{\Delta\theta}{\alpha_1}\right) \cdots\cdots \left(1 - \frac{\Delta\theta}{\alpha_{N-1}}\right)\frac{1}{\alpha_N} \right\} \qquad (4)$$

If the above equation is transformed with the use of the average value, α_m of α_1, α_2 ....., $_n$ in each stage, the ratio of the amount of evaporation to the amount of make-up seawater may be expressed as follows;

$$D/F = 1\left(1 - \frac{T_0 - t_F - \Delta t}{\alpha_m (N+1)}\right)^N \quad [\mathrm{Kg/Kg}] \qquad (5)$$

When the number of stages is infinite, the limiting value of D/F is;

$$\lim_{n\to\infty}\frac{D}{F} = 1 - e^{-(T_0 - t_F - t)/\alpha_m} \qquad (6)$$

If the concentration of circulating seawater is taken as X_F, the concentration of discharged fluid, X_n, is;

$$x_n = x_F \frac{1}{\left(1 - \frac{T_0 - t_F - \Delta t}{\alpha_m (N+1)}\right)^N} \quad [\mathrm{wt\%}] \qquad (7)$$

The required calories in the heater is;

$$qH = F\frac{T_0 - t_F + N\Delta t}{N+1} \quad [\mathrm{kcal/hr}] \qquad (8)$$

The required calories per kg of evaporation is;

$$\frac{qH}{D}=\frac{T_0-t_F+N\Delta t}{(N+1)\left[1-\left(1-\frac{T_0-t_F-\Delta t}{\alpha_m(N+1)}\right)^N\right]}\ [\mathrm{kcal/kg}] \qquad (9)$$

The theoretical limiting value for the infinite number of stages is;

$$\lim_{n\to\infty}\frac{qH}{D}=\frac{\Delta t}{1-e^{-(T_0-t_F-\Delta t)/\alpha_m}}\ [\mathrm{kcal/kg}] \qquad (10)$$

Fig. 4.4-4 shows the characteristics of the multistage flash evaporation process, in which the relationship between the temperature of circulating seawater, T_0, entering the first stage and the required calories is illustrated. The parameter in this case is the number of all stages and the figure may be used as an index to the assessment of approximate heat economy of the process.

When materials of high cost are used for heat transfer tubes of condensation and heat transmission units, the construction cost of a plant is largely dependent on the area of heat transmission. The difference in cost can be obtained from the following. If the arithmetic mean, Δm, is used under the previously mentioned assumption, the effective temperature difference across the heat transfer tube is;

$$\Delta_m=\Delta t+\frac{T_0-t_F-\Delta t}{2(N+1)}\ [℃] \qquad (11)$$

Then the approximate value of the heat transfer rate, q, is;

$$q=D\alpha_m\ \ [\mathrm{kcal/hr}] \qquad (12)$$

The equation of the heat transfer rate is;

$$q=U\cdot A\cdot\Delta_m\,[\mathrm{kcal/hr}] \qquad (13)$$

Therefore, the required heat transfer area, A, is as follows;

$$A=\frac{D\alpha_m}{U\Delta_m}=\frac{D\alpha_m}{U\left[\Delta t+\frac{T_0-t_F-\Delta t}{2(N+1)}\right]} \qquad (14)$$

Adoption of a large final temperature difference (T_0 - t_n) in the design results in the increase of construction costs of the plant and the decrease of the required amount of heating steam. Adoption of a small value has the reverse effect as in the case of the relation between the number of vessels and the cost of the multiple effect process. It is necessary, therefore, to select the most economical final temperature difference in designing a plant.

Fig. 4.4-3 is an example of an economic final temperature difference calculated by Brice for a combination of a boiling water reactor and a multistage flash evaporator for production of fresh water from the sea. This optimization is for a case where 100,000 - 240,000 m^3/day of desalinated water is produced without power generation at a maximum circulating seawater temperature of 98.0°C for prevention of scale deposition, and a concentration ratio (concentration of discharge seawater/concentration of raw seawater) of 2.0, in the 50 stage flash evaporation process with a supply of 370 heat MW steam from the reactor. Under this condition, the most economic temperature difference is 2.3°C as shown in the figure.

In the above assumption, the number of stages is specified. The number of stages also has a great influence on the required steam for evaporation as shown in Fig. 4.4-4 and has some effect on the required heat transfer area. Fig. 4.4-5 is an example of the latter case taken up by Frankel.

The example which uses the amount of steam consumption as a parameter, shows that the required heat transfer area changes greatly with the change of the number of stages, that this effect diminishes when the number of stages is outside a certain range, that there is a final theoretically required area and that the boiling point rise, which was neglected in Höffer's assumption, should not be ignored in practice.

4.5. Combination Process

With the development and practical application of multiple effect, vapor compression and multistage flash processes, coupled with changes in plant size, energy situations such as fuel and power supply and water consumption, a combination of these three processes has been suggested or studied and necessary pilot tests of fundamental technology, conceptional design and cost study of various forms have been carried out. The following is a summary of these studies and experiments.

The basic consideration of the development of a combination of the multiple effect process and multistage flash process is to cover the following demerits of the former process. They are that the existing multiple effect evaporator requires preheating of seawater to the temperature of the first effect of the maximum temperature for normal flow operation and the provision of a preheater between each effect and that no water is produced from the heating surface of these preheaters. The concept of using a multistage flash evaporator as a preheater, as shown in Fig. 4.5-1 or the development of a combination of the multiple effect process and the multistage flash process to solve this problem has long been held by O.S.W. and is now in progress at various stages.

One of such efforts is an experiment with the use of a module, which integrates six effect evaporators and an 11 stage flash evaporator, conducted at Free Port during the period from 1971 to 1972.

The stability of the combination was satisfactory with easy start up and the object of the experiment was achieved in several hours.

The share of the cost of heat transfer tubes in the total investment was approximately 40 percent for a large plant and 20 percent for a small plant, with the former showing a higher thermo-dynamic efficiency and a smaller total heat transfer area depending on the combination.

Development researches with a pilot plant of the previously mentioned horizontal three effect vessels have been carried out by O.S.W., which has reported the findings on heat transfer coefficient and characteristics of the process against fluidized scale and on the other hand, a test has been conducted in Wrightsville Beach with a pilot plant equipped with heat transfer tubes and an evaporator body made of Aℓ alloy, which was developed as a combination of multiple flash-vertical tube multiple effect-horizontal tube multiple effect (MSF-VTE-HTE). With the use of H_2SO_4 for a maximum temperature of 121°C and a scale inhibitor (Darex 40) for a maximum temperature of 87°C, corrosion resistance of 3003, 3004 and 5052 has been obtained and the effect of sacrifice anode has been confirmed. The result of process cleaning shows that the VTE-HTE combination with MSF as a preheater contributes to the reduction of water production cost of 25.59$/1,000 gal and simplifies plant design. If enhanced tubes can be used for HTE, the efficiency of this combination will be improved further.

As an advanced form of this combination, the conceptional design of the vapor compression - vertical tube effect - multistage flash (VC-VTE-MSF) process has been commissioned to Fluor Corp. In this process, a vapor compressor driven by a gas turbine, a 34 stage flash evaporator and a quadruple effect evaportor are combined as shown in Fig. 4.5-2, by which the heat of exhaust gas from the gas turbine is recovered by a heat recovery boiler and the steam generated in the boiler is used as a heat source for the multiple effect evaporator.

Based on the study of the results of tests and developments, conceptional designs, and final specifications of recent years and of the size of existing plants, the total construction cost of a plant of the 8 MGD (30,000 m^3/day) scale is estimated at 11.84×10^6 ($) and the cost of water production is estimated at 59 - 67¢/1,000 gal with the fuel cost of 30 - to¢/10^6 Btu.

From a study of the conceptional designs for various combinations of processes and the result of pilot tests of various elements, it is easily recognized that the limiting factors in the design of a seawater desalination plants are the deposition of soft and hard scale and corrosion of component equipment. If these two problems can be solved by some creative means or technical innovation, there will be a major change in the operating conditions and the phase of a process or a combination of processes depending on the environment in which the plant is located, and the result of such change is expected to contribute greatly to the improvement of economy and welfare of the people. The effect of the improvement in this respect will be tremendous as long as the object of development is water, one of the most important elements for the survival of mankind.

The continued development and research in various sectors even after the completion of this big project is much hoped for.

4.6. Solar Heat Process

The same as other evaporation (distillation) processes, the solar heat process involves evaporation and condensation, but it is a well known fact that the process requires no main energy sources as long as sufficient solar heat is provided. It is a process which is capable of converting up to 45 m^3/day of water regardless of the type of raw water used. However, the application of this process is limited to regions of strong solar radiation, high fuel cost and very little or no labor force, such as large deserts.

Transparent materials such as glass transmit the short wave portion of solar radiation. Any surface placed under a glass cover can be heated by this solar radiation. Since the wave length of reradiation from the surface is only in arange of small radiation energy, there is no rise of temperature in the glass cover. This effect is applied to distillation of seawater which uses solar heat to produce fresh water from the sea.

The mechanism of solar heat radiation through glass is shown in Fig. 4.6-1, in which the greater portion of the heat transmits directly through the glass. Typical examples of solar radiant heat on the earth through the year are 4,400 - 5,500 [Kcal/m^2 day] in Batticaloa Cailon (07°43'N, 81°42'E), 3,240 - 5,950 [Kcal/m^2 day] in Nairobi, Kenya (01°18S, 46°45'E), 420 - 4,280 [Kcal/m^2 day] in Cambridge (52°13'N, 00'06'E), and 1,390 - 4,990 [Kcal/m^2 day] in Boston, U.S.A. (42°21'N, 71°04'W).

Since eighty-five percent of solar heat reaches the object at an incidence angle of up to 35°, application of solar heat to saline water conversion is generally limited to a narrow strip on both sides of the equator line bound by 35°N and 35°S, including large deserts.

The solar radiant heat in Alice Spring, Australia (23°48'S, 133°53'E) is 3,370 - 6,460 [Kcal/m^2 day].

The still used for saline water conversion is very simple in construction and is a brine sink consisting of a radiation absorption lining, a supporting structure and a glass or plastic cover. Fig. 4.6-2 shows a brine sink with a glass cover, which was installed at a research station on Aldabra Island of the Indian Ocean by Porteous, and which was a standard pre-fabricated construction already used by local fishermen. The dimension of the sink is $42\,m^2$ x 4 modules. A trough was provided to collect rainwater so that the production during the wet season of less radiant heat could be increased.

Evaporation starts when the temperature reaches 50° to 65° and before reaching a saturation state inside the cover, the vapor condenses and forms drops over the inclined inside surface of glass, which collect in the trough because of the wetting nature of glass and is then taken out of the still. Evaporation continues until the concentration of seawater reaches twice the original level and the seawater is discharged and replaced with a fresh supply. Normally, the depth of the sink is changed three to seven times a day according to the incidence of sun light. While low cost construction of the still consisting of a plastic roof, light supporting structures and polyethelene lined sink is recommended by Porteous, nothing can replace a glass roof, durable supporting structures and a heavy lining for a longer life, reliability and easy maintenance according to him. Plastic material is hard to wet, unfit for draining water, and the mist formed inside the still prevents the incidence of solar heat. Besides, plastic materials have a short life, as they decompose under the effect of ultra-violet rays during a period of high radiation.

The key point of continuous operation is how to control a very thin film of seawater running down beneath the 20 m long glass cover at a slope of 1/60. If any portion of the lining of the sink dries up, the heated section expands gradually and scale deposition increases rapidly.

The amount of water production is dependent on the intensity of solar radiation and still area. The published data on the stills in practical use are given in Table 4.6-1 [Porteous, p.89.1]. There is little difference in the rate of water production between a large system and a small system and the approximate capacity is 5 [$\ell/m^2 \cdot$day] when the amount of radiation is 7,000 Kcal/$m^2 \cdot$day.

The theoretical capacity is determined by the amount of radiation and heat loss in the ground. The heat loss in the ground is shown in Fig. 4.6-3. The heat loss in sand and sand gravel layer is 5.7 W/m^2°C.

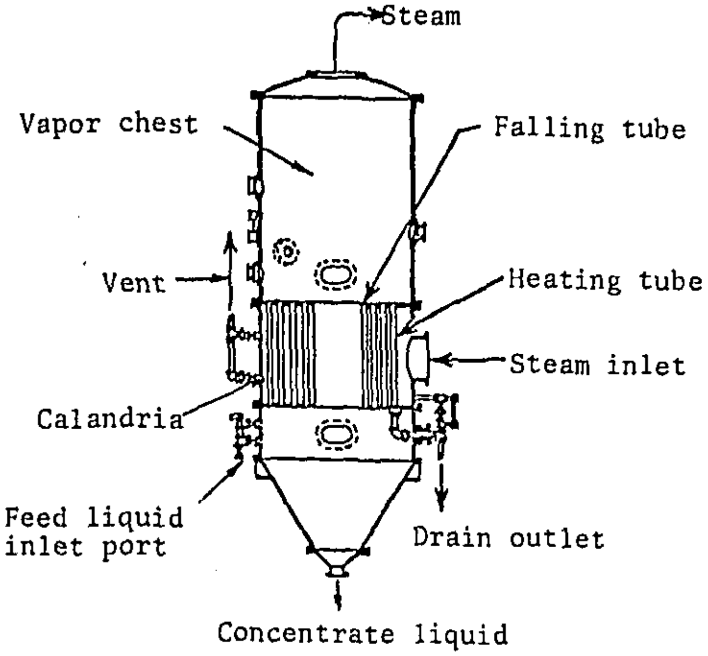


Fig. 4.1-1 Steam boiler, standard type

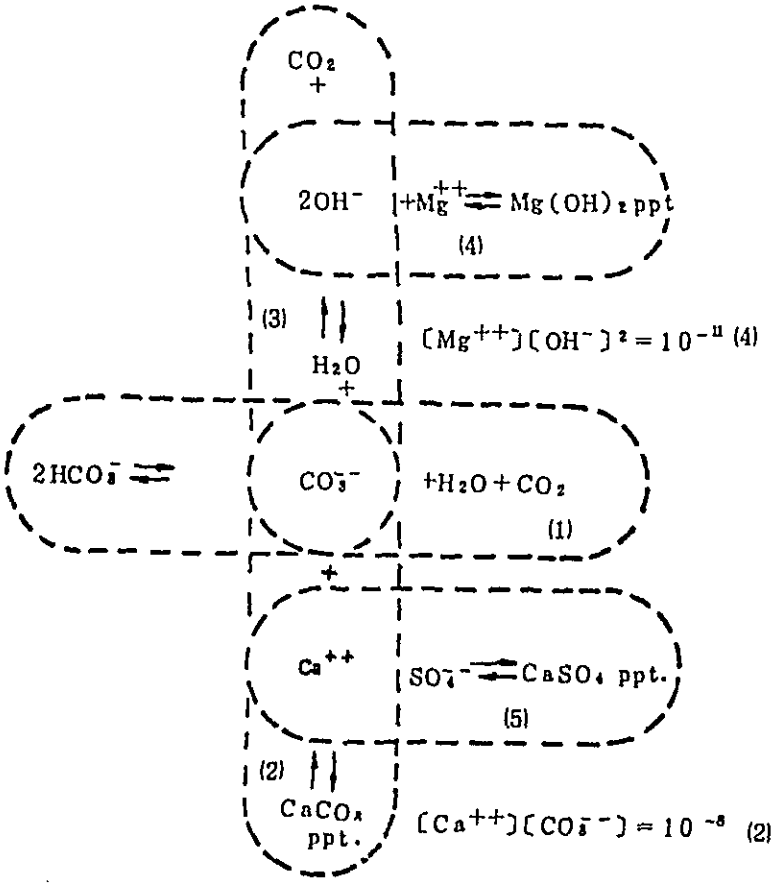


Fig. 4.1-2 Reaction of seawater scale deposition

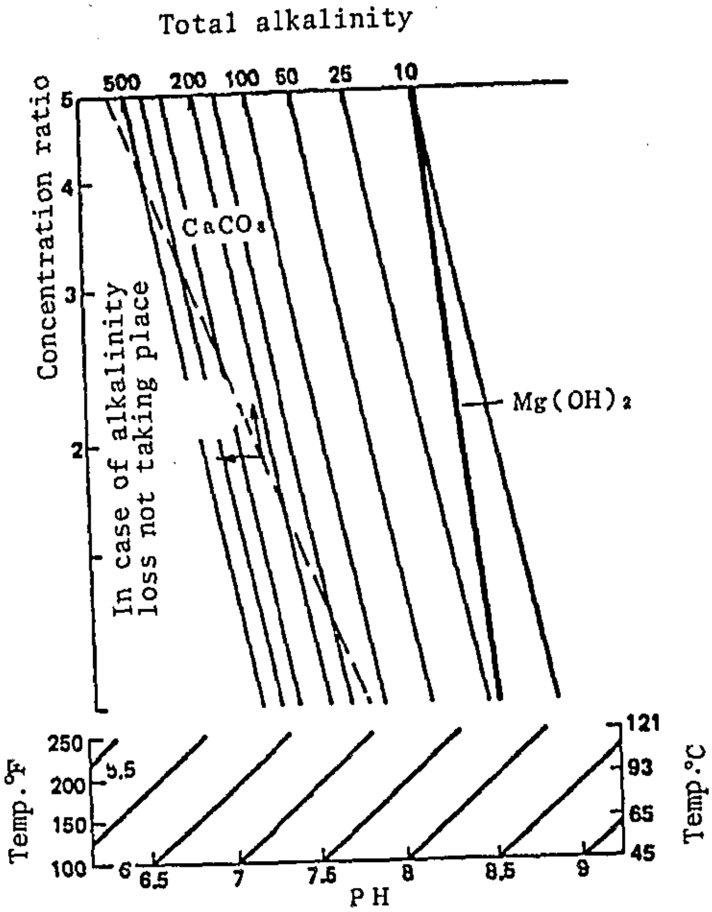


Fig. 4.1-3 Solubility of $CaCO_3$, $Mg(OH)_2$ in seawater

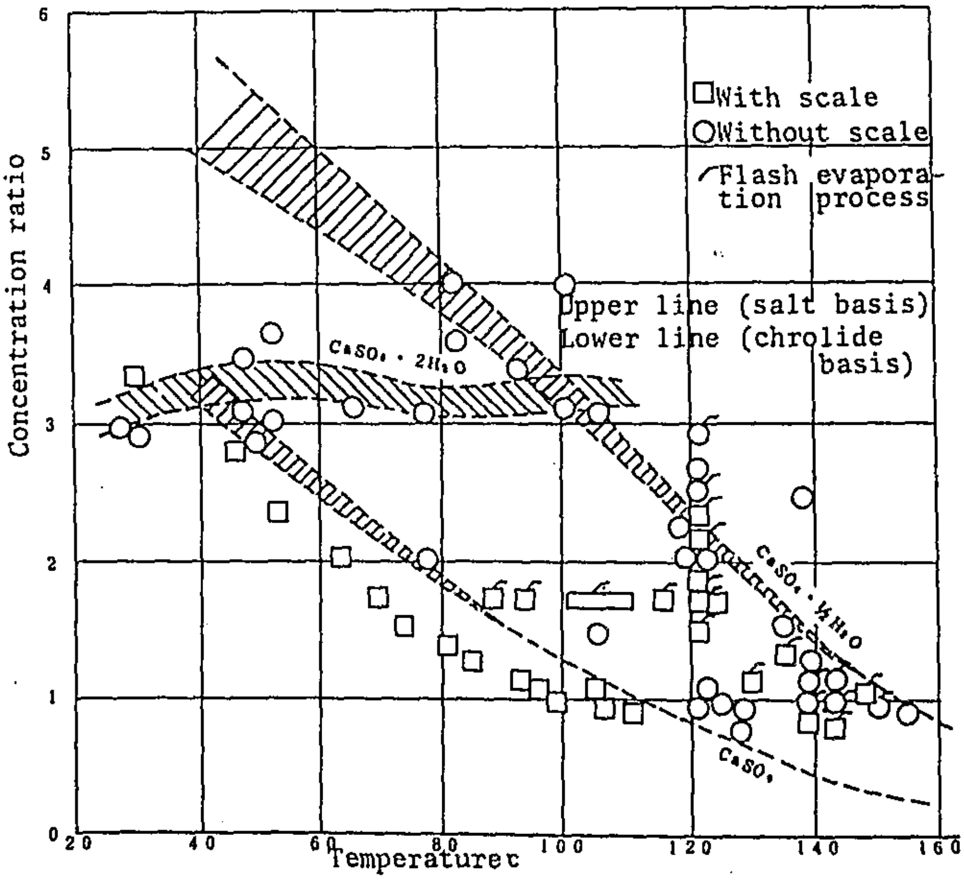


Fig. 4.1-4 Scale deposition and solubility of $CaSO_4$

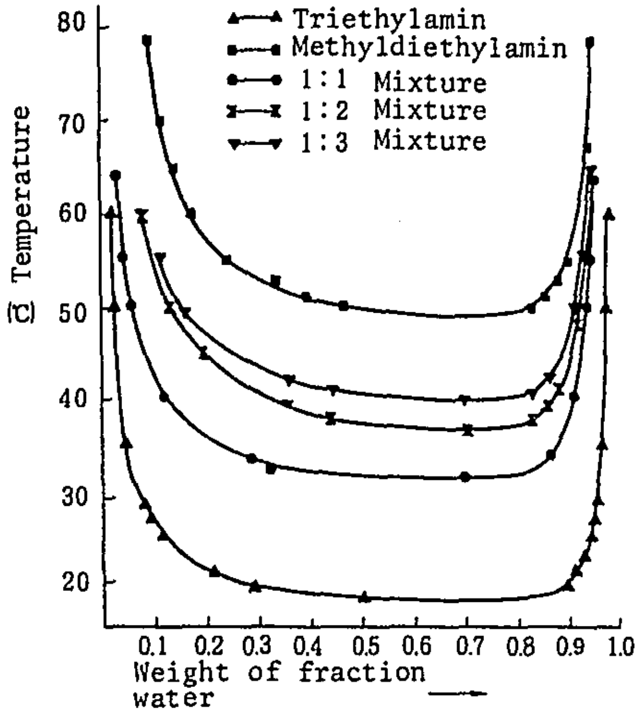
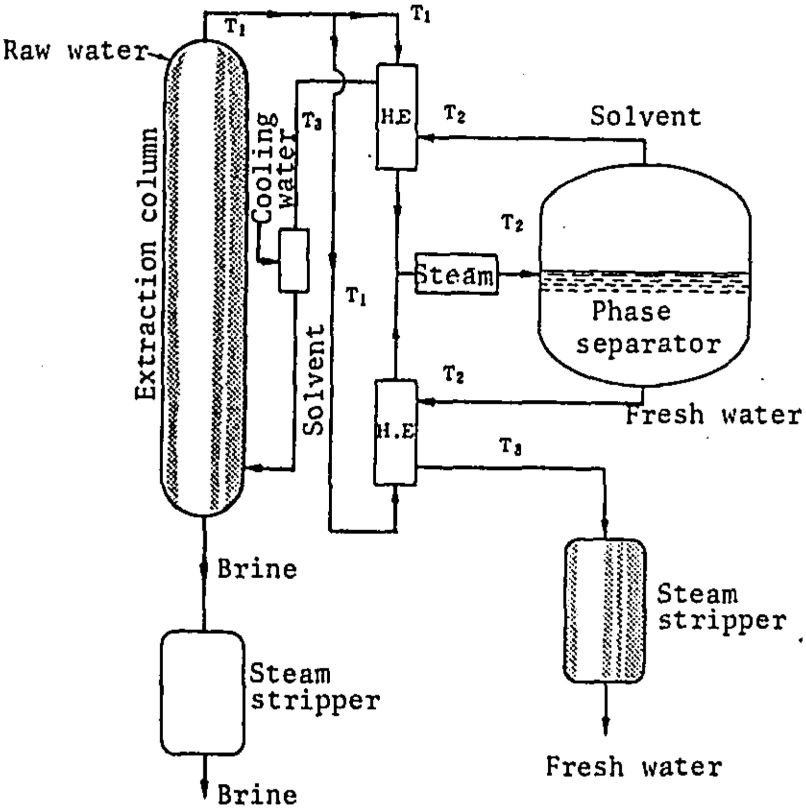


Fig. 3.4-1 Solubility curves for triethylamin, methyldietylamin and mixtures of the two



Note: H.E.; Heat exchanger

Fig. 3.4-2 Flow sheet of desalination plant by extraction process

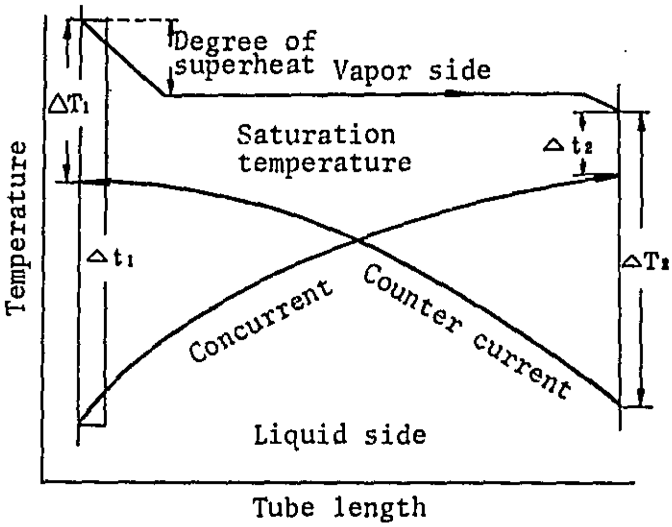


Fig. 4.1-7 Profile of condensed heat-transmission temperature

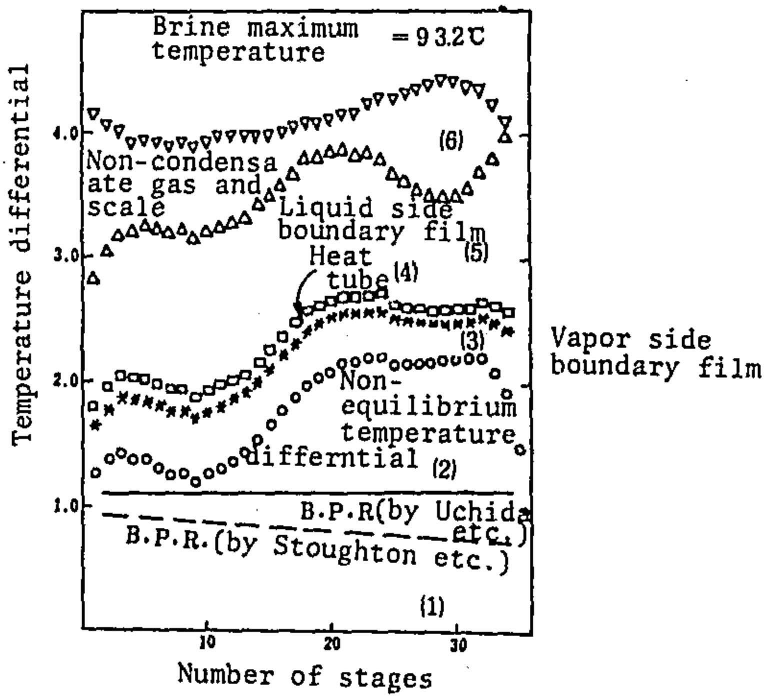


Fig. 4.1-8 Example of heat-transfer resistance in M.S.F.

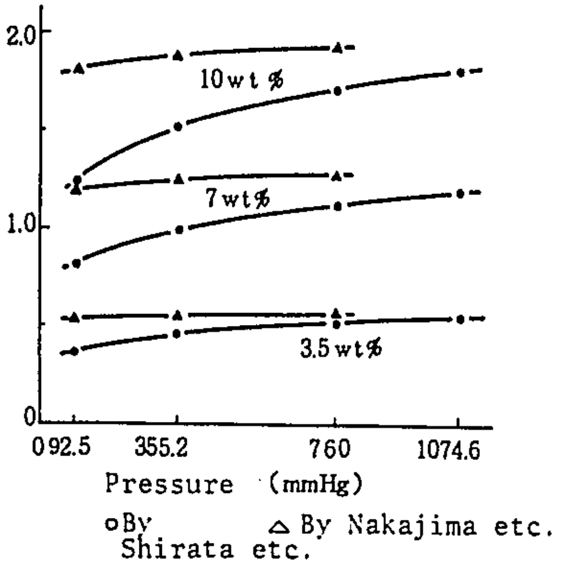


Fig. 4.1-9 Barometric pressure and concentration versus boiling point of seawater

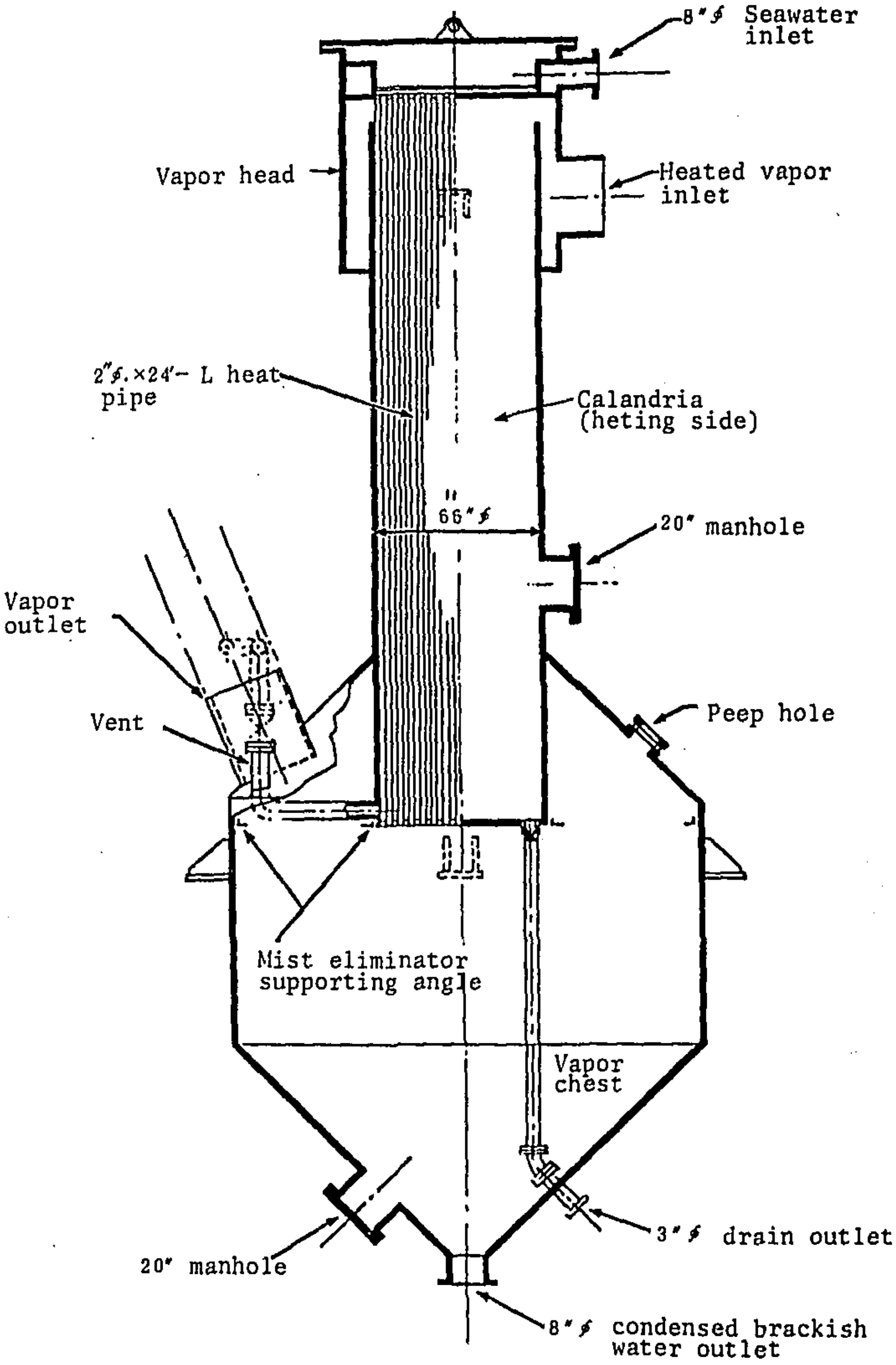


Fig. 4.2-1 Sketch of LTV

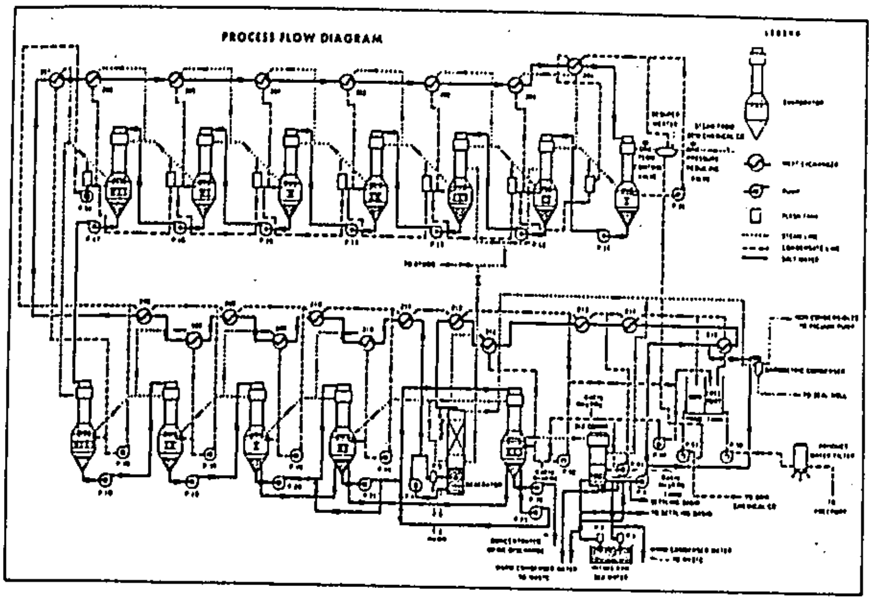


Fig. 4.2-2 Process flow diagram of dual-effect pilot plant

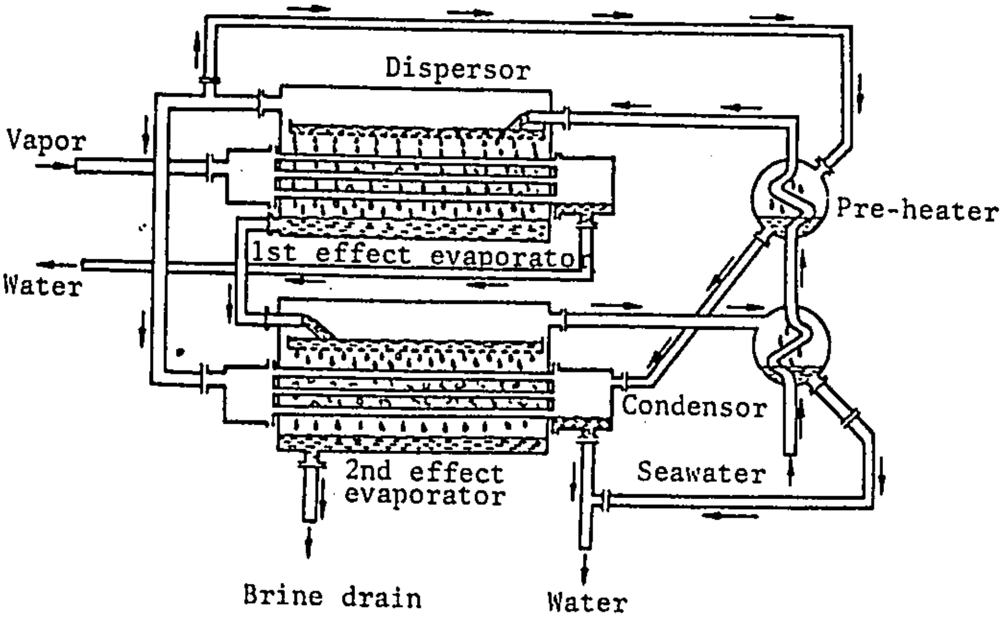


Fig. 4.2-3 Dual-effect evaporator by horizontal tube process

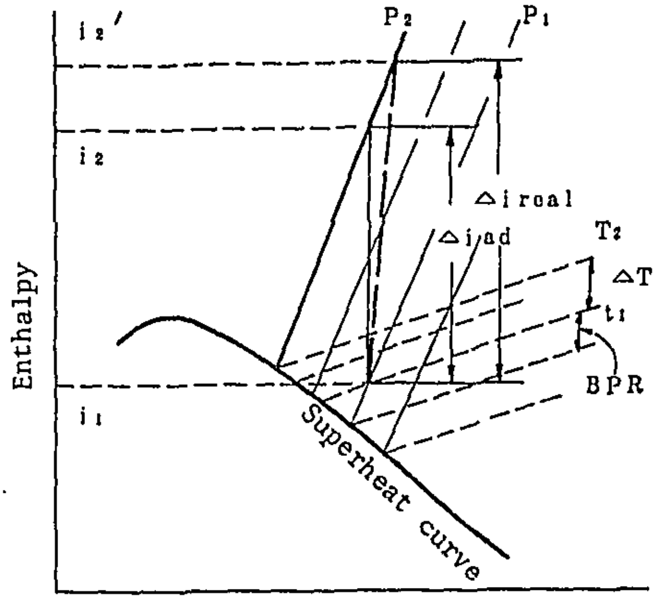


Fig. 4.3-1 i - S chart of vapor compression process

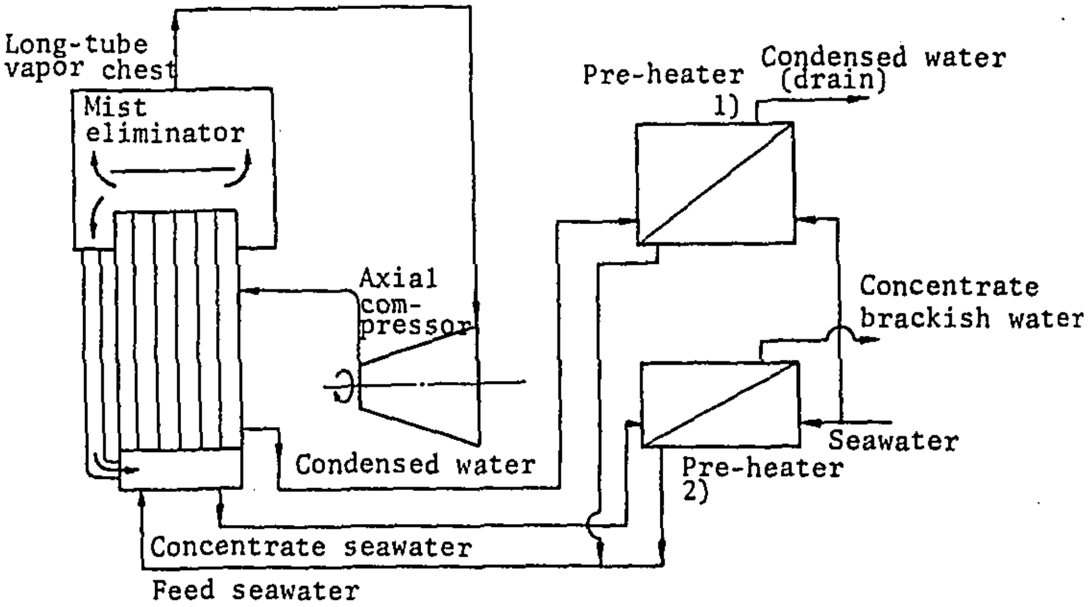


Fig. 4.3-2 Development of diverse-form utilization process of seawater, flow diagram of vapor compression process, pilot test on process concentration of seawater for industrial use by NCL

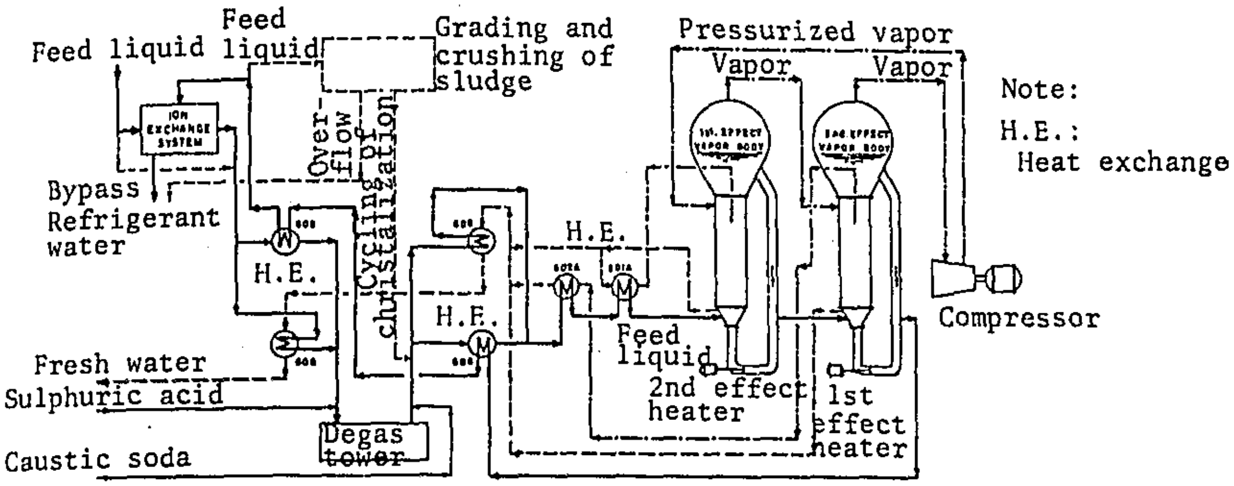


Fig. 4.3-3 Pilot plant by dual-effect vapor compression process, built by O.S.W. (Rosewell)

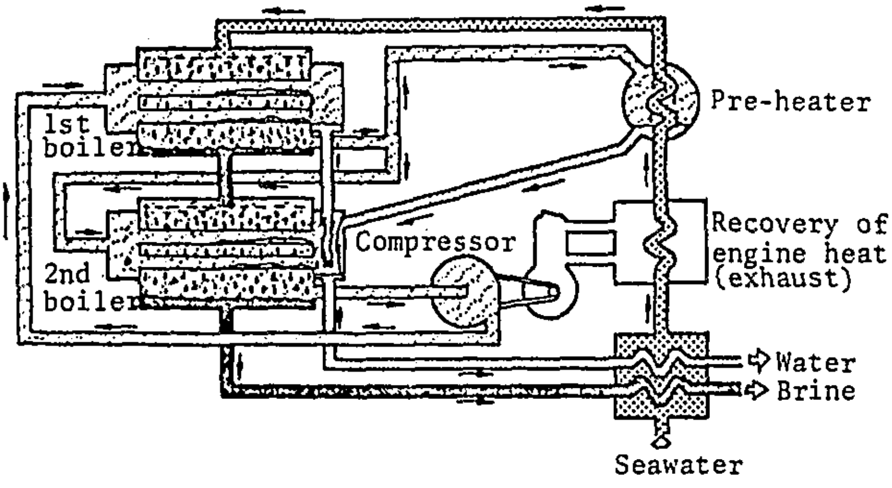
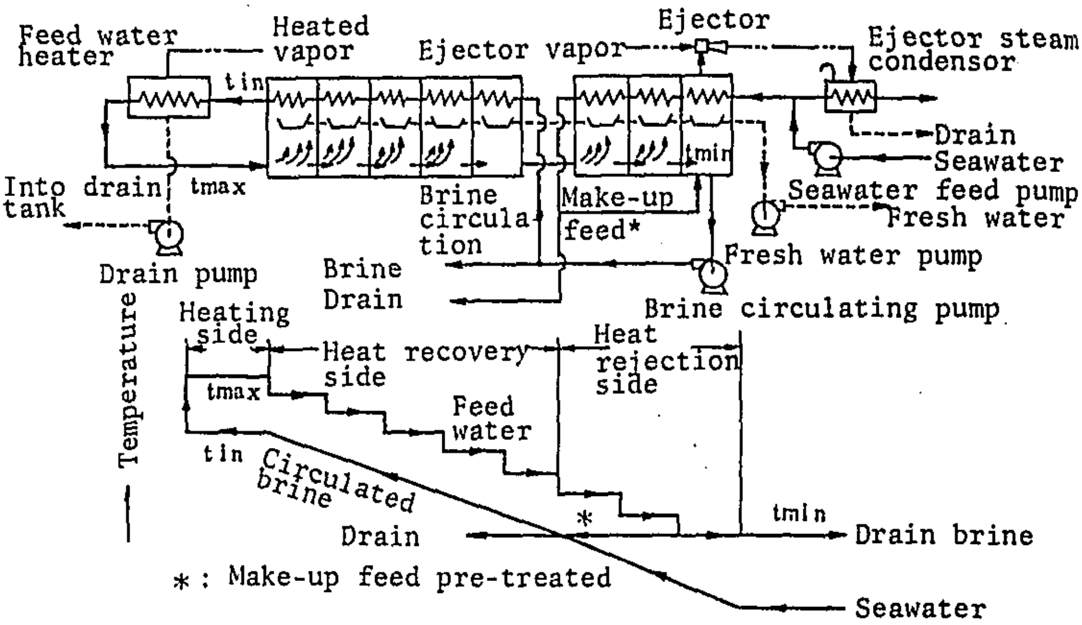
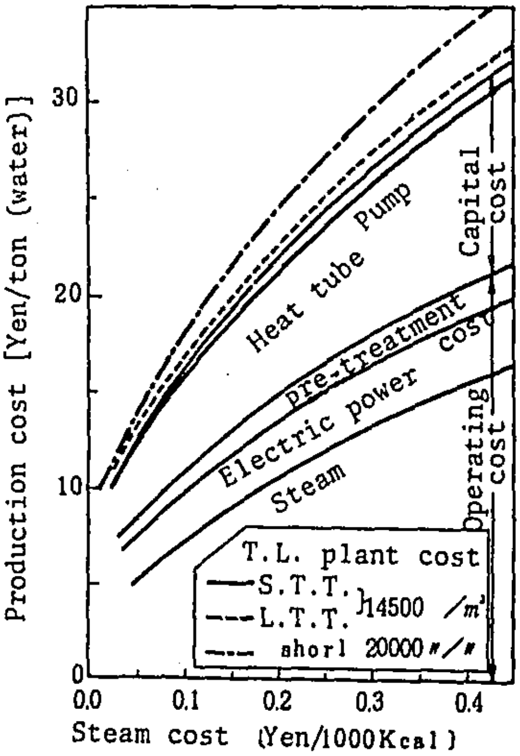


Fig. 4.3-4 Multiple effect vapor compression process



*: Make-up feed pre-treated

Fig. 4.4-1 Process diagram of multistage flash evaporation plant (brine recycle type)



Note:
T.L.; Tube length
S.T.T.; Short tube type
L.T.T.; Long tube type

Fig. 4.4-2 Optimum production cost by MSF process versus vapor cost, by estimation

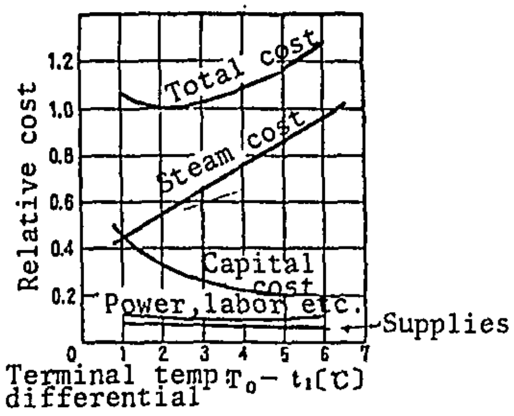


Fig. 4.4-3 Desalination cost by 50-stages flash evaporation process vs. terminal temperature differential (by nuclear power, without power generation, 100 - 240,000 ton/day)

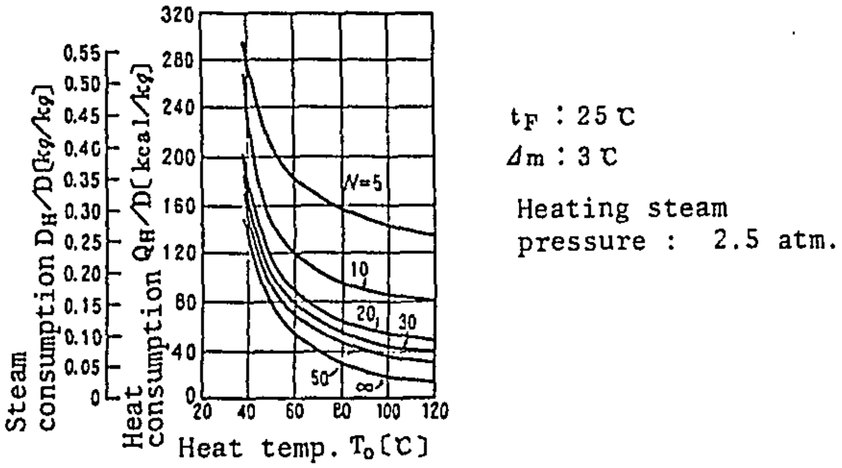


Fig. 4.4-4 Heat temperature vs. heat quantity by MSF process (multistage flash evaporation)

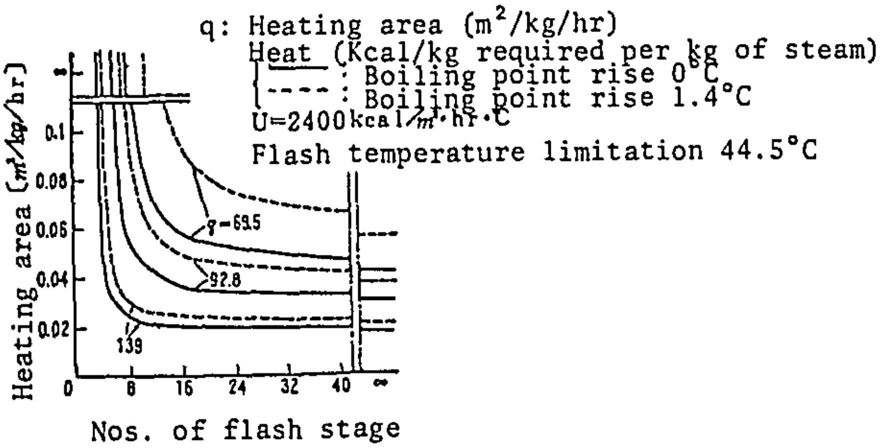


Fig. 4.4-5 Heating area vs. nos. of flash stage

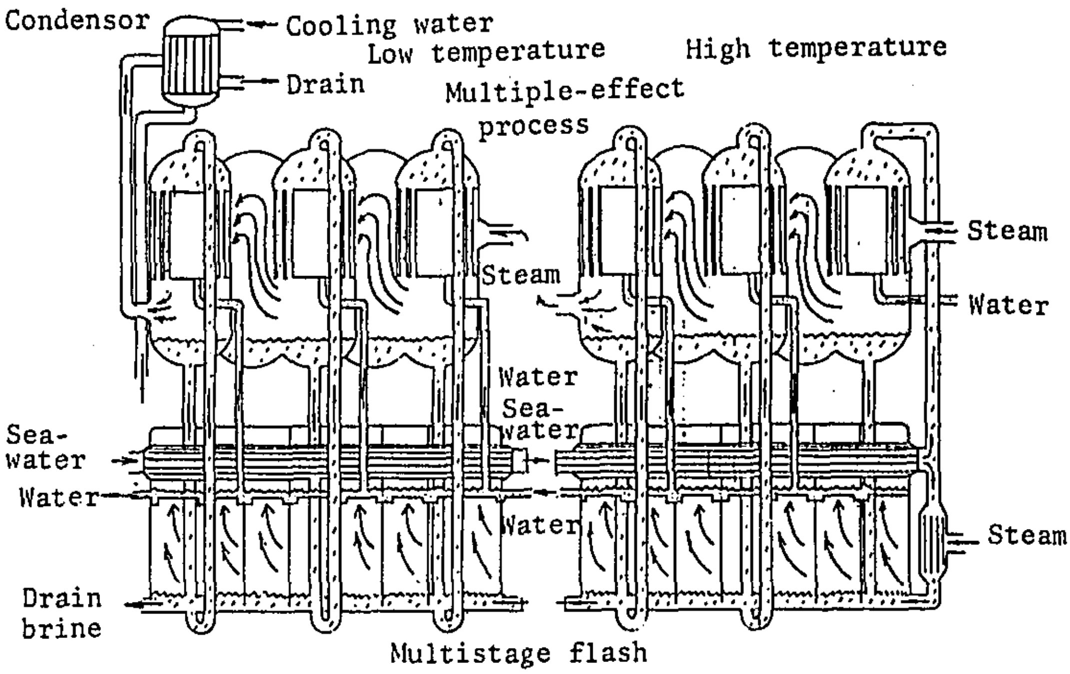


Fig. 4.5-1 Diagram of combination, multiple effect multistage flash process

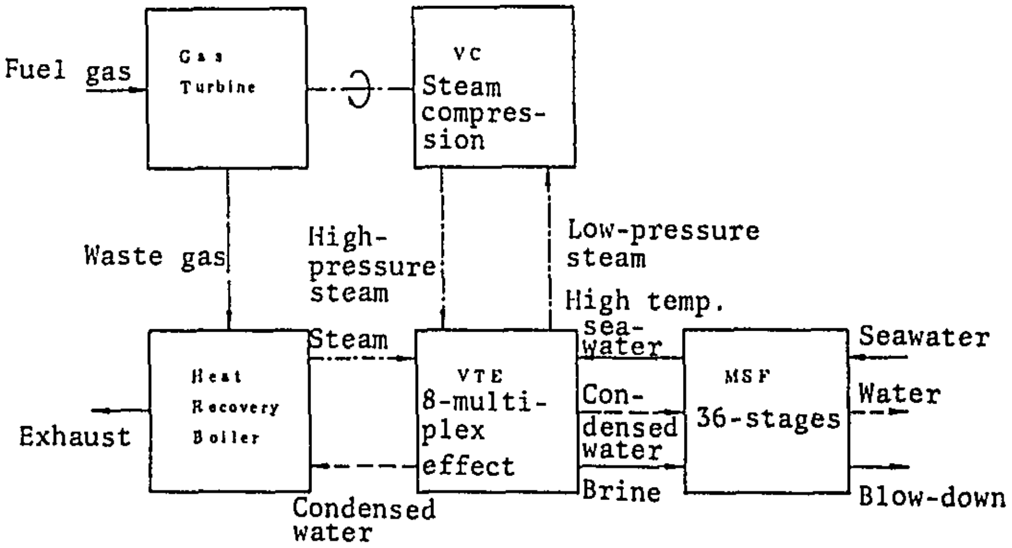


Fig. 4.5-2 Steam compression - Multiple effect - Multistage flash, combination

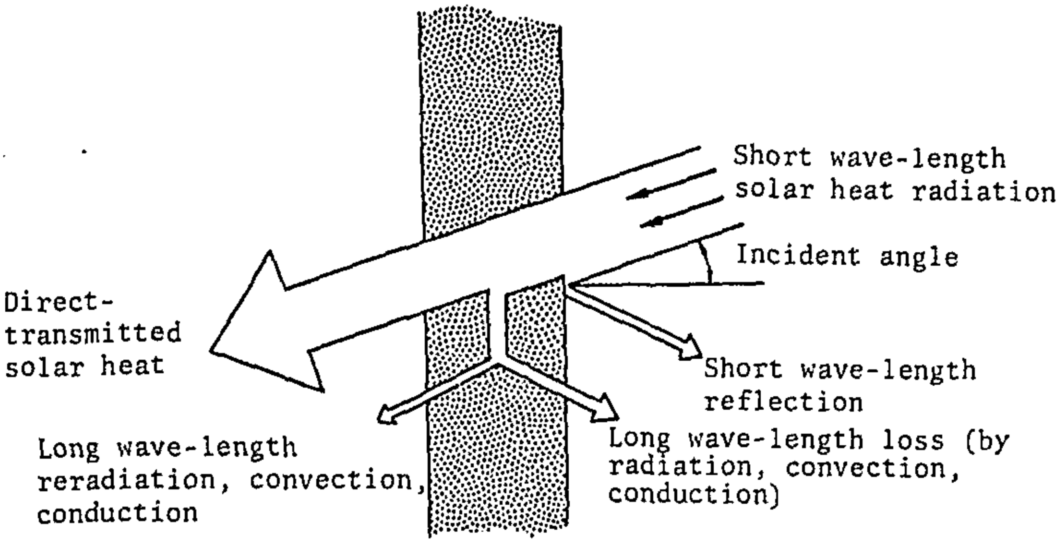


Fig. 4.6-1 Principle of solar heat radiation through glass

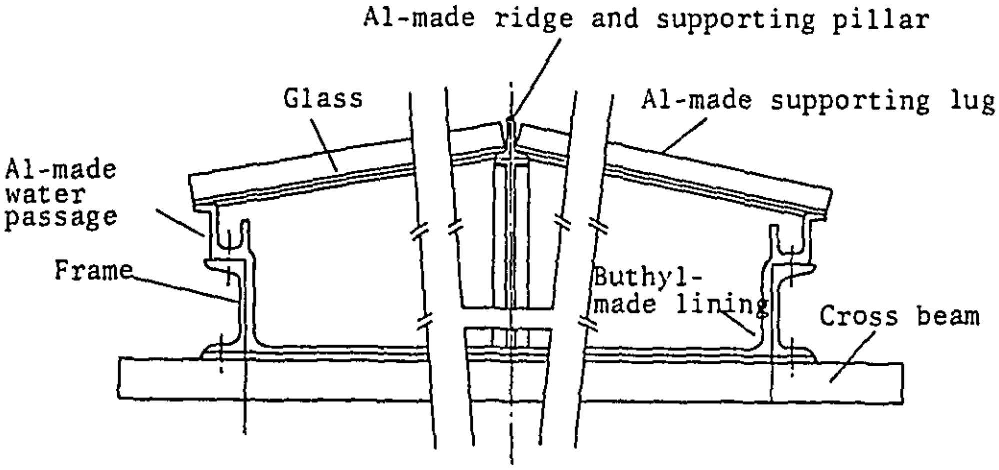


Fig. 4.6-2 Water reservoir type desalination plant by solar heat

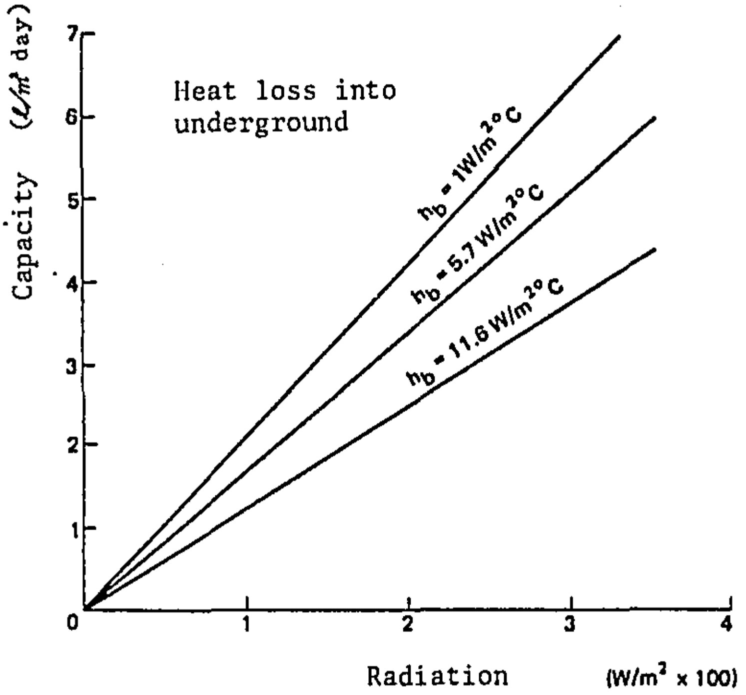


Fig. 4.6-3 Desalination plant capacity by solar heat vs. radiation

Table 4.1-1 Property of main ingredients of seawater

	g/kg	g/l	gmol/l
Total salts	35.1	36.0	
Sodium (Na)	10.77	11.1	0.482
Magnesium (Mg)	1.30	1.33	0.0554
Calcium (Ca)	0.409	0.42	0.0105
Potassium (K)	0.388	0.39	0.010
Chrolide (Cl)	19.37	19.8	0.558
Sulphate (SO_4)	2.71	2.76	0.0288
Bromide (Br)	0.065	0.066	0.0008
Bicarbonate (HCO_3)	0.149	0.152	0.0025

Table 4.1-2 Boiling point rise of seawater

S/T	50°C	60	70	80	90	100	110	120	130	140	150
2.00%	0.200	0.217	0.234	0.250	0.269	0.285	0.302	0.320	0.337	0.355	0.373
3.00	0.309	0.334	0.360	0.385	0.412	0.438	0.465	0.492	0.519	0.547	0.575
3.50	0.365	0.395	0.425	0.455	0.486	0.517	0.549	0.581	0.614	0.646	0.680
4.00	0.424	0.457	0.492	0.527	0.562	0.598	0.635	0.672	0.710	0.748	0.787
5.00	0.544	0.587	0.630	0.675	0.720	0.766	0.813	0.861	0.910	0.960	1.010
6.00	0.671	0.722	0.775	0.829	0.884	0.941	0.999	1.058	1.118	1.180	1.243
7.00	0.803	0.863	0.926	0.990	1.055	1.123	1.192	1.263	1.336	1.411	1.487
8.00	0.941	1.011	1.083	1.157	1.233	1.312	1.393	1.477	1.562	1.650	1.741
9.00	1.085	1.164	1.246	1.331	1.418	1.509	1.602	1.698	1.798	1.900	2.005
10.00	1.236	1.324	1.415	1.511	1.610	1.712	1.819	1.928	2.042	2.159	2.280
11.00	1.391	1.489	1.591	1.697	1.808	1.923	2.043	2.167	2.295	2.428	2.565
12.00	1.553	1.660	1.773	1.890	2.013	2.141	2.275	2.413	2.557	2.706	2.860
13.00	1.721	1.838	1.961	2.090	2.225	2.367	2.514	2.668	2.828	2.994	3.166
14.00	1.895	2.021	2.155	2.296	2.444	2.599	2.761	2.931	3.108	3.291	3.483
15.00	2.074	2.211	2.355	2.508	2.670	2.839	3.016	3.202	3.396	3.598	3.809

Table 4.6-1 Data of seawater desalination by solar heat

Location	Water reservoir area (m^2)	Constructed year	Type	Average water product $1/m^2/day$	Average solar heat $Kcal/m^2/day$	W/m^2	Source
Indian Ocean (Aldabra Island)	167	1970	Reservoir type	3.8	5,160	250	Porteous' data from Aldabra Research Station
Chile: Andes (1,300m above sea level)	4,700	1872	Reservoir type	5.76 (max)	8,150	395	Harding (1883)
Greece:							
Patmos	8,667	1967	Reservoir type	3.0	Not given	-	Delyannis and Piperoglou (1967)
Symi	2,700	1964	Reservoir type	2.62	3,743	182	Delyannis and Piperoglou (1968)
Australia:							
Coober Pedy	3,500	1966	Continuous type	3.22	5,100	246	Morse (1967)
Muresk	416	1936	Continuous type	2.2	5,100(Porteous presumption)	246	Morse (1967)
Pacific Island	465	1966	Reservoir type	4.3 (max)	6,300(Porteous presumption) from Lof 1966	305	Howe and Tleimat (1967)

V. Saline Water Conversion By The Multistage Flash Distillation Method (Process)

5.1. Characteristics of The Multistage Flash Process

5.1.1. Multistage Flash Process

Since the multistage flash distillation plant has already been described, its operational requirements will be analyzed here on the basis of the temperature and flow conditions given in the chart of operating principles in Fig. 5.1-1. The steam in the flash chamber preheats seawater inside the heat transfer tube,

$$T_v \geq T_{Smax} \qquad - (1)$$

$T_v = T_{Smax}$ holds only when the heat transfer area is infinite, which is impossible in practice. The same applies to a brine heater as follows.

$$T_h \geq T_{Bmax} \qquad - (2)$$

If the remaining heat and a heat source are taken into consideration.

$$T_{Smax} \geq T_S \qquad - (3)$$

$$T_{Bmax} \geq T_{Smax} \qquad - (4)$$

The equal sign is for a case where there is no heat transfer through heat transfer tube. Since flash evaporation has already occurred in the meantime,

$$T_{Bmax} \geq T_b \qquad - (5)$$

The equal sign is for a case where there is no flash evaporation. Since there is a boiling point rise in seawater.

$$T_b \geq T_v \qquad - (6) \qquad - (6)$$

These temperature relations may be given graphically as a temperature distribution curve as shown in Fig. 5.1-1. If the specific heat of seawater is taken as constant and expressed by C_p and the evaporation latent heat of water is taken as constant and expressed by λ, the following equation may be derived from the balance relation around the heat transfer tube in the flash chamber.

$$W \cdot \lambda = W_S \cdot C_p \cdot (T_{Smax} - T_S) \qquad - (7)$$

From the balance relation around the flash brine, the following may be derived in the same manner.

$$W \cdot \lambda = W_s \cdot C_p \cdot (T_{Bmax} - T_b) \quad - (8)$$

In the above, the amount of fresh water production is considered to be negligibly small as compared with the amount of seawater.
From equations (7) and (8);

$$T_{Smax} - T_s = T_{Bmax} - T_b \quad - (9)$$

or

$$T_{Bmax} - T_{Smax} = T_b - T_s \quad - (10)$$

Equation (9) shows that all of the heat lost by seawater through flash evaporation has been utilized for preheating of seawater and has been recovered. Equation (10) shows that all of the external heat applied by the brine heater has been released by discharged seawater. From equation (8), the following may be derived:

$$W/W_s = C_p \cdot (T_{Bmax} - T_b)/\lambda \quad - (11)$$

From this, it is known that the amount of fresh water production per unit flow of seawater is given as the ratio of the amount of sensible heat change to evaporation latent heat. Since the evaporation latent heat of water at 100°C is at a high level of about 540 Kcal/kg, a large amount of sensible heat change should be provided by increasing the range of flash temperature as much as possible. Otherwise, only a small amount of fresh water is available from a large amount of seawater.

On the other hand, the following equation may be derived from the heat balance at the brine heater:

$$W_h \cdot \lambda = W_s \cdot C_p \cdot (T_{Bmax} - T_{Smax}) \quad - (12)$$

In the case of a single stage flash evaporator, $Tb \geq Tsmax$ holds because of the relation between equations (1) and (6). From equation (8) and (12), therefore, the following may be derived.

$$W_h \geq W \quad - (13)$$

This means that after all the amount of fresh water production is smaller than that of steam utilized.

Since this is meaningless, a large number of flash chambers of different pressure are linked together in order of high pressure to flash evaporate concentrated seawater repeatedly to provide a wider range of flash temperature and, at the same time, the ratio of Tb to Tsmax is decreased drastically to produce fresh water in an amount several times greater than the amount of heating steam. This is the theory of the multistage flash distillation plant. As mentioned above, the flash process makes it possible to produce a large amount of fresh water as compared with the amount of heat applied if the temperature of seawater at the outlet of the brine heater is maintained as high as possible and the temperature of discharged seawater is maintained as low as possible. The multistage flash evaporator has two types. One is the through flow type shown in Fig. 5.1-2 and the other is the recirculation type shown in Fig. 5.1-3. As seen from the figure, the through flow type has no circulation flow and wastes the total amount of flash brine discharged from the last stage. In the case of the recirculation type, on the other hand, the flash brine discharged from the last stage is divided into two flows, of which one is wasted and the other is lead to the heat transfer tube. The through flow type is not in use at present, as it requires pretreatment of the total amount of seawater entering the heat transfer tube, thereby increasing the cost of pretreatment per unit of water product as will be explained in detail in the next section. However, its application to very limited special purposes, such as a saline water conversion plant, in which a large quantity of cooling seawater used by other industries is directly utilized without being heated for flash evaporators or a multistage flash evaporator which preheats seawater in the multiple effect type saline water conversion plant, is being considered. This is only a special case and multistage flash evaporators used today are of the recirculation type in nearly all cases and all of the plants delivered to foreign countries centering around the Middle East are of this type. In the following, only the recirculation type will be described.

The multistage flash evaporator consists of four large sections, namely, brine heater, heat recovery section, heat rejection section and make-up seawater pretreatment section as shown in Fig. 5.1-3. The brine heater, as already shown by Fig. 5.1-1, heats the fluid in the heat transfer tube with externally supplied heating steam and provides a driving force for flash evaporation. The heat recovery section causes the flash brine heated by the brine heater to repeat flash evaporation while losing its sensible heat as evaporation latent heat and at the same time condenses the generated steam on the surface of heat transfer tubes and recovers this latent heat as a sensible heat of the fluid in the heat transfer tubes. Through this distillation process, fresh water is separated from brine. This section plays a central role in the multistage flash evaporator. The heat rejection section releases the heat provided by the brine heater at low temperature levels because of the relation shown by the previous equation (10). Different from the through flow type in which the total amount of flash brine is discharged from the last stage, this system releases heat with the cooling seawater. In other words, this section conducts seawater into the heat transfer tube as cooling water and discharges the greater portion of preheated seawater directly to the outside while separating fresh water from brine and supplies the remainder to the pretreatment equipment as new make-up water. The make-up seawater pretreatment section removes CO_2 dissolved in seawater or adds a scale inhibitor to prevent scale deposition, as heating of untreated seawater causes scale deposition on the interior surface of heat transfer tubes, which hampers heat transmission and drastically decreases plant efficiency. This section also removes oxygen dissolved in seawater to prevent corrosion of equipment. Pretreatment of seawater is described in detail in the following section.

Besides, a multistage flash distillation plant always requires venting equipment comprising an ejector and a condenser, auxiliary equipment including mainly a seawater supply pump and a brine recirculating pump, instrumentation for stable operations and control and seawater intake and discharge facilities. The intake facilities are provided for intake of clear seawater without inclusion of foreign matters such as sand and treat seawater with chlorine at the tip of intake pipe to prevent deposition of organisms inside the intake pipe. The discharge facilities discharge concentrated waste brine of a temperature higher than that of seawater with attention paid to the effect on the environment. These are discussed in detail in section 5.9.

The reducing agent tank listed in the table as a component of pretreatment facilities is a supplement to a vacuum deaerator and is used for mixing reducing agents such as sodium sulfite with brine to reduce the concentration of oxygen when the circulating brine has a high concentration of oxygen. It is auxiliary equipment and is not always required. Other auxiliary equipment includes ball cleaning device which removes sludge or scale deposited on the internal wall of heat transfer tubes without requiring a shut-down of the plant.

The steam producing unit listed in the table is required only by a single purpose plant when no external heating steam source is available. Acid pickling and alkali cleaning equipment are for cleaning and removing of scale deposited on heat transfer tubes using acid or alkali.

The flow of the multistage flash evaporation process is basically the same as that of the previously discussed single stage flash evaporation process but the temperature distribution of the former differs greatly from that of the latter as shown in Fig. 5.1-4. Let us examine the flow with Figs. 5.1-3 and 5.1-4. The seawater chlorinated at the intake enters heat transfer tubes of the heat rejection section and increases its temperature after receiving latent heat. The greater portion of seawater coming out of the heat rejection section is discharged as cooling water but the remainder is pretreated and is mixed with recirculating brine as make-up seawater. The recirculating brine enters heat transfer tubes of the heat recovery section, increases its temperature after being preheated in various stages, comes out of the heat recovery section and then enters the brine heater. Here, the recirculating brine increases its temperature after being heated further by an external heat source, passes through an orifice and reaches the flash chamber where it becomes flash brine. Since the flash brine repeats flash evaporation as it passes through various stages, it shows a stage temperature distribution as illustrated in Fig. 5.1-4. The steam produced in each stage which has a temperature higher than that of recirculating brine at the inlet of each stage, condenses on the surface of the heat transfer tube and accumulates in the tray as fresh water. The temperature of this fresh water, which is in vapor-liquid equilibrium, is generally treated as the same as the steam temperature. Like the flash brine, this fresh water also repeats flash evaporation as it passes through various stages, decreases its temperature gradually while contributing to preheating of recirculating brine and comes out from the last stage as fresh water.

In Fig. 5.1-4, the temperature of make-up seawater is the same as that of flash brine of the last stage. This is a design object to minimize energy loss by blending make-up seawater and recirculating brine at the same temperature. However, it only means that the proximity of temperature of both liquids is desirable in practice and does not necessarily mean that the temperature must be identical, as the energy loss in this case is negligible as compared with other cases.

In Fig. 5.1-4, the steam temperature is lower than flash brine temperature in each stage and this difference is considered to be equal to the sum of a boiling point rise, which is determined by brine concentration and temperature and non-equilibrium temperature difference, which is related to flash brine temperature, flash temperature range between stages, flash brine flow rate, and flash brine liquid level. Besides, the pressure loss by mist eliminator and the pressure loss by tube bundles further contribute to the decrease of steam temperature. Details of non equilibrium temperature differences, pressure loss by mist eliminator and pressure loss by tube bundles are given in section 5.3. Since the temperature difference between flash brine and steam obstructs the efficient separation of fresh water, efforts should be made to minimize this difference.

In order to understand the concept of non-effective temperature difference, let us examine the temperature distribution around the first stage of the evaporator. In Fig. 5.1-5, two states of temperature difference between brine and steam in the first stage, ΔT_I and ΔT_{II}, are assumed. If the heat value botained by recirculating brine in the first stage is taken as Q [Kcal/hr], it is expressed as follows.

$$Q = U_0 \cdot A_1 \cdot \Delta T \qquad - (14)$$

where: $\Delta T = (Ts_{max} - Ts1)/\log_e\{T_{V1} - T_{S1})/(T_{V1} - Ts_{max})\}$
U_0 : First stage overall heat transfer coefficient (Kcal/m^2·hr·°C]
A_1 : First stage heat transfer area [m^2]
ΔT : Log-mean temperature difference [°C]

Assuming that the steam temperature T_{V1} has increased from state I to state II, Q, which means the left member of the previously discussed equation (7), remains constant as far as T_{Bmax} and T_{B1} are constant because of the relation of equations (7) and (8). Since U_0 and A_1 are constant, ΔT is also constant and this relation is satisfied if the temperature distribution in tube bundle increases by $(\Delta T_{II} - \Delta T_I)$ simultaneously. However, since T_{S1} is ruled by the second stage and the subsequent stages in practice, the value of T_{B1} decreases slightly and the value of Q becomes greater with the increase of temperature of recirculating brine at the outlet of the first stage in the relation of Ts_{max} II > Ts_{max} I. While the overall flashing temperature range remains the same, the volume of heating steam decreases because of the relation of equation (12) and the amount of water production per unit volume of heating steam is increased, thereby contributing to the improvement of plant efficiency. Discussed so far is a case in which non-effective temperature difference is improved in only one stage. If improvement is made in all stages, T_{S1} increases eventually and T_{Smax} increases further, thereby contributing to the improvement of plant efficiency. For plant efficiency, the amount of water production per unit weights of heating steam is normally called the water production gain ratio (performance ratio), which is used to express the thermal efficiency. From the above, it can be said that a small value of non-effective temperature difference estimated in the plant design at a set performance ratio can reduce the required heat transfer area.

5.1.2. Process Design

The explanation in the previous section has clarified the multistage flash process in outline. If the performance ratio is to be increased, the value of Ts_{max} in Fig. 5.1-4 must be increased. On the other hand, there is a restricting condition (equation (1)) that since Ts_{max} is lower than the steam temperature in the first stage (T_{V1}). To increase Ts_{max}, it must be brough closer to T_{V1}. To do this, the required heat transfer area per stage must be increased as seen from the relation shown by equation (14), which eventually increases the cost of the plant. For process design, therefore, it is impossible to select the optimum plant design and specifications by correlating it first with economic factors. Optimization of the process, however, involves various steps and the range of application may vary from one step to another. In other words, a study of the merits and demerits of saline water conversion planned to secure water resources in a certain area in relation to the river flow and environmental conditions may be one case of optimization, in a broad sense, and the determination of plant design and specifications as discussed here may be another case of optimization. Here, optimization (of the process) of only the latter will be discussed. This optimization, of course, has a prerequisite that optimization or study of a wider range has already been completed. In the case of a multistage flash distillation plant, for example, a study of such elements as the previously mentioned throughout flow type and recirculating type, or short tube type and long tube, which are the fundamentals of the process, must be completed prior to the start of process design.

(1) Design model

Once a decision is made as to the type of process to be employed, optimizing calculations must be carried out to determine the design and specifications of the plant. For calculation purposes, a simulation model of a multistage flash distillation plant is constructed. This model consists simply of mass balance, heat balance and heat transfer formulae. Normally, two types of simulation model are provided and optimizing calculations are made in two stages.

One type is a simple model representing the plant with many assumptions, by which optimizing calculations are made for as many parameters as possible, and the other is a detailed model with the least assumptions, by which accurate optimizing calculations are made for parameters determined by the previous calculation.

(2) Simple model

For a simple model, equation (15) derived on the basis of the following assumptions is convenient.

- Assumptions used for deriving a simple model -

(1) Shape and size of the stage are the same for all stages.

(2) The volume of condensed water in each stage is negligible as compared with the flow rate of flash brine, and the flow rate of brine at the inlet is equal to that at the outlet of each stage.

(3) Physical properties of brine are same in each stage and are equal to those in the last stage.

(4) Evaporation latent heat in each stage is equal to the value at the average of the maximum and minimum temperatures of the plant.

(5) Boiling point raise and non equilibrium temperature difference in each stage are equal to the value at the concentration of discharged brine.

(6) There is no heat loss.

$$\eta = \frac{N(1-\Delta t)\{1-\exp(-N_H)\}}{1+N\cdot\Delta t\cdot\{1-\exp(-N_H)\}} \quad - \quad (15)$$

where N: Number of stages

NH: Number of heat transfer unit ($U_{av} \cdot A/W_R/C_p$)

A: Heat transfer area

C_p: Specific heat of brine

U_{av}: Average overall heat transfer coefficient

W_R: Flow rate of recirculating brine

Δt: Non effective temperature ratio

$$\left(= \frac{\text{Boiling point raise + Non equilibrium temperature differe}}{\text{max. brine temperature - seawater temperature}}\right.$$

μ: Water production gain ratio (performance ratio)
(= Amount of water production/amount of heating steam)

From the above equation, it is known that a smaller non-effective temperature ratio results in a greater performance ratio as discussed previously. It is also possible to know from the equation of the general characteristics of the multistage flash process that the performance ratio is proportional to the number of stages if the value of the non-effective temperature ratio is ignored, that an increase of the maximum brine temperature results in an increase of the performance ratio, and that the division of denominator and numerator of the left number of equation (15) by $\{1-\exp(-N_H)\}$ increaseses the heat transfer area and heat transfer coefficient, thereby contributing to the increase of the performance ratio.

For optimization, this model must be combined with economic factors, but for input data, the design conditions, which are determined separately, are used. These data include the concentration and temperature of seawater, which are determined by natural conditions at the plant site, the maximum brine temperature, which is determined by the level of pretreatment technics, and the amount of water production, which is determined by the project requirement. As seen from equation (15), a higher maximum brine temperature results in a wider flash temperature range and a greater performance ratio.
A decrease of non-equilibrium temperature difference in sections of normally high temperatures improves plant efficiency. For this

reason, a higher maximum brine temperature is generally advantageous even when such demerits as the requirement for an increase of heating steam and the resultant increase of steam cost are taken into consideration. However, the high concentration and high temperature of recirculating brine cause the deposition of hard scale on heat transfer surfaces, thereby decreasing drastically the heat transfer coefficient in pretreatment equipment which uses the pH control method - the most economical and most efficient at present. It is evident that there is an upper limit of maximum brine temperature. The upper limit of maximum brine temperature under the present technical level is 121°C when the concentration ratio of recirculating brine is about 1.7.

Strictly speaking, the concentration and temperature of seawater vary from place to place, but the average value, or standard value, or a value slightly higher than the standard value for the sake of safety, is adopted for design purposes.

After determination of the design, optimizating calculations are carried out with economic factors taken into consideration.
The economic parameters in this case include the steam cost, power cost, personnel cost, cost of chemicals for pretreatment, material and equipment cost related to the construction of the plant, cost of auxiliary equipment and facilities, cost and interest per unit volume of materials and the term of depreciation.

Fig. 5.1-6 shows an example of optimization of the number of stages, length of stage, flow rate of brine in heat transfer tubes and the seawater concentration ratio for a long tube type plant calculated by equation (15). This illustrates how the optimum performance ratio and the number of stages change depending on the steam cost and equipment cost. It shows a general tendency that the optimum value can be found in each case if the number of stages is increased to improve the performance ratio when there is an increase of the steam cost, and if the number of stages is decreased to reduce the performance ratio when there is an increase of equipment cost.

The methodology of optimization includes the gradient method, dynamic programming method and discrete maximum principle method, which are selected according to the purposes.

Steam cost has increased several times since 1973. While there has been a remarkable increase in the cost of plant, the increase of steam cost and other energy costs is far greater. Optimization of plant design, therefore, requires expenditure of a huge amount of investment to increase the number of stages and expend the performance ratio beyond the limit of the present technology of plant operation. In other words, the number of stages and the performance ratio being employed for the production plant are not determined from the result shown in Fig. 5.1-6, but are selected from the upper limits, which are close to the optimum value and which are made available by the present technical level. The upper limit with the present technical level is said to be 50 in the number of stages and about 15 in performance ratio in the case of a large plant having a capacity of more than several tens of thousand cubic meters per day.

After the main components of the plant have been determined in outline, calculation is made with a detailed model for each element to determine the optimum design of a plant by obtaining the temperature, amount of condensed water and heat transfer area of each stage.

(3) Detailed model

As mentioned previously, the detailed model uses as few assumptions as possible and obtains the final result from the accumulation of calculations made for each stage. Therefore, the model comprises mainly balance equations and heat transfer equations. The following are equations which comprise a detailed model under different classifications.

° Overall balance relation

A flow chart marked with variables is given in Fig. 5.1-7 to explain the balance relation. The meaning of symbols used in the chart is given at the end of this section. The input and output of mass and heat in the whole system have only a simple relation that seawater enters the system, that cooling seawater, product water and discharge brine come out of the system and that the brine heater has only a heat input. From these balance relations, the following equation may be derived.

$$W_s = W_c + W_b + W_w \qquad -\ (16)$$

$$W_s \cdot C_s = W_c \cdot C_s + W_b \cdot C_b \qquad -\ (17)$$

$$Q_h + (W_s - W_c) \cdot i_s = Q_c + W_b \cdot i_b + W_w \cdot i_w$$

where

$$Q_h = W_h \cdot \lambda_h$$

$$Q_c = W_c \cdot (i_c - i_s) \qquad -\ (18)$$

° Brine heater

Since the flow of mass is separated to inside and outside of the tube and there is no exchange of flow, only the balance equation relative to heat transfer from outside to inside of the tube can be derived.

$$Q_h = W_R \cdot (i_{Bmax} - i_{smax}) \qquad -\ (19)$$

To obtain the heat transfer area of the brine heater, the equation of heat transfer is used.

° Blending point of make-up seawater and recirculating brine

$$W_M = W_S - W_C \qquad -\ (20)$$

$$W_R = W_M + W_{RC} \qquad -\ (21)$$

$$W_R \cdot C_R = W_M \cdot C_S + W_{RC} \cdot C_b \qquad -\ (22)$$

$$W_R \cdot i_{RC} = W_M \cdot i_C + W_{RC} \cdot i_b \qquad -\ (23)$$

For design calculations, the cooling water discharge temperature, make-up seawater temperature and temperature of the last stage are assumed to be the same (refer to Fig. 5.1-4).

- Heat rejection section

The balance relation around the heat rejection section shown in Fig. 5.1-7 concerns raw seawater and fresh water and flash brine from the heat recovery section for input, and cooling seawater and make-up seawater and product water and flash brine from the last stage for output.

$$W_{wj} + W_{Bj} = W_w + W_{RC} + W_b \qquad -(24)$$

$$W_{Bj} \cdot C_{Bj} = (W_{RC} + W_b) \cdot C_b \qquad -(25)$$

$$W_{Bj} \cdot i_{Bj} + W_{wj} \cdot i_{wj} = Q_c + W_w \cdot i_w + (W_{RC} + W_b) \cdot i_b \qquad -(26)$$

Equations (24) through (26) are used together with the following equation for obtaining an approximate value of temperatures at the outlet of the heat recovery section.

$$(W_w - W_{wj}) \cdot \lambda_{av} = W_{Bj} \cdot i_{Bj} + W_{Bj} \cdot i_{wj}$$
$$- W_w \cdot i_w - (W_{RC} + W_b) \cdot i_b \qquad -(26)'$$

In the above, λav is the average evaporation latent heat of the heat rejection section.

- I^{th} stage

A balance flow sheet of the I^{th} stage is shown in Fig. 5.1-8.

$$W_{wI-1} + W_{BI-1} = W_{wI} + W_{BI} \qquad -(27)$$

$$W_{BI-1} \cdot C_{BI-1} = W_{BI} \cdot C_{BI} \qquad -(28)$$

$$Q_I = (W_{wI-1} \cdot i_{wI-1} + W_{BI-1} \cdot i_{BI-1}) -$$
$$(W_{wI} \cdot i_{wI} + W_{BI} \cdot i_{BI}) \qquad -(29)$$

In the above,

$$Q_I = W_R \cdot (i_{RI} - i_{RI+1})$$

$$Q_I = (W_{wI} - W_{wI-1}) \cdot \lambda_I \qquad (30)$$

Described so far are related equations of mass balance and heat balance of the entire process and of the I^{th} stage. Besides, the design calculation requires a heat transfer equation (equation of heat transfer rate), because there is a transfer of heat, equations showing the relation between the steam temperature and the flash brine temperature in the flash chamber and values of physical properties or equations of physical properties. The following is a summary of related equations required for detailed simulations.

- Heat transfer in I^{th} stage

In Fig. 5.1-8, the heat Q_I applied to the fluid in the heat transfer tube is the heat transmitted through the heat transfer tube de to the difference between the steam temperature in the flash chamber and the brine temperature in the heat transfer tube. This relation is the same as that of equation (14) used for a study of the effect of non-equilibrium temperature difference.

$$Q_I = U_I \cdot A_I \cdot \Delta T_{lm} \qquad - (31)$$

where $$\Delta T_{lm} = \frac{(T_{VI} - T_{RI+1}) - (T_{VI} - T_{RI})}{\ln\left(\dfrac{T_{VI} - T_{RI+1}}{T_{VI} - T_{RI}}\right)}$$

U_I is called the overall heat transfer coefficient, which includes the heat transfer coefficient of condensation, heat conductivity of tube materials, heat transfer coefficient inside the tube and the fouling factor. Details will be given in section 5.3. A_I represents a heat transfer area. ΔT_{lm} represents the so-called log-mean temperature difference, the average temperature difference between the fluid temperature in the heat transfer tube which varies depending on the location, and the steam temperature.

° Steam temperature of I^{th} stage

Steam temperature T_{VI} is considered to be lower than flash brine temperature by the sum of boiling point raise (ΔT_{BP}) and non equilibrium temperature difference (ΔT_{NE}) and is generally determined by the following equation.

$$T_{VI} = T_{BI} - (\Delta T_{BPI} + \Delta T_{NEI}) \qquad - (32)$$

In practice, however, there are pressure losses such as pressure loss by mist eliminator, pressure loss by heat transfer tubes and flow pressure loss and the decrease of temperature through these pressure losses must not be discounted. However, even the pressure loss by mist eliminator, which is considered to be the greatest among these losses, is so small in value that it is generally neglected as mentioned in section 5.3.3. Since the non equilibrium temperature difference will be described in detail in section 5.3.1, the related equations will not be taken up here. Steam temperature is assumed to be equal to condensed water temperature of this stage.

° Boiling point raise (ΔT_{BP}[°C])

T: Flash brine temperature (°F)

C: Flash brine concentration [weight fraction]

$$\begin{aligned}\Delta T_{BP} = (&0.1014 38\times10^{-1}+0.1021815\times10^{2}\times C-0.6386588\times10^{-4}\times T\\ &+0.6118005\times10^{-1}\times C\times T+8.809554\times C^{2}+0.2386759\times10^{-6}\times T^{2}\\ &+0.2214495\times10^{-4}\times C\times T^{2}+0.1714722\times C^{2}\times T+0.7173776\times10^{2}\\ &\times C^{3}+0.1170617\times10^{-9}\times T^{3}+0.1042795\times10^{-6}\times C\times T^{3}+0.3620461\\ &\times C^{3}\times T+0.7823281\times10^{-4}\times C^{2}\times T^{2}-0.5218751\times10^{-6}\times C^{2}\times T^{3})\\ &/1.8 \qquad - (33)\end{aligned}$$

° Specific heat of brine C_{PB} [Kcal/kg·°C]

T: Brine temperature [°C]

C: Brine concentration [weight fraction]

$$C_{PB} = 1.0022 - 1.6405\times10^{-4}\times T - 1.4185\times C + 6.0118\times10^{-3}\times T\times C + 2.109\times10^{-6}\times T^2 + 2.1753\times C^2 - 1.5937\times10^{-2}\times T\times C^2 - 2.8817\times10^{-5}\times C\times T^2 + 1.166\times10^{-4}\times T^2\times C^2 - 0.135\times10^{-2}\times C\times T \qquad - (34)$$

- Specific heat of water C_{PW} [Kcal/kg·°C]

T: Water temperature [°C]

$$C_{PW} = 1.0060 - 3.5810\times10^{-4}\times T + 4.1465\times10^{-6}\times T^2 \qquad - (35)$$

° Enthalpy of brine i_B [Kcal/kg]

Obtained from

$i_B = \int_0^T C_{PB}\,dT$ より求めた。

$$i_B = \{1.0022 - 0.82025\times10^{-4}\times T + 0.703\times10^{-6}\times T^2 - (1.4185 - 2.3309\times10^{-3}\times T + 0.96057\times10^{-5}\times T^2)\times C + (2.1753 - 0.79685\times10^{-2}\times T + 0.3887\times10^{-4}\times T^2)\times C^2\}\times T + 35.62\times C - 194.48\times C^2 + 549.872\times C^3 \qquad - (36)$$

° Enthalpy of water i_W [Kcal/kg]

Obtained from

$i_W = \int_0^T C_{PW}\,dT$ より求めた。

$$i_W = (1.0060 - 1.7905\times10^{-4}\times T + 1.38217\times10^{-6}\times T^2)\times T \qquad - (37)$$

° Heat conductivity of brine k [Kcal/m·hr·°C]

T: Brine temperature [°K]

$$k = 0.8601\times10^{-3}\times \exp\{\log(240.0 + 200.0\times C) + [2.3 - (343.5 + 370.0\times C)/T]\times[1.0 - T/(647.3 + 30.0\times C)]^{0.333}\} \qquad - (38)$$

° Density of brine p [kg/m^3]

S: Salinity [%]

T: Brine temperature [°C]

$$\rho = 0.5\times10^3\times\{0.5\times4.032219 + 0.115313\times SA + 0.000326\times FA(SA) - [0.108199 - 0.003142\times SA + 0.000846\times FA(SA)]\times TA - [0.012247 - 0.003480\times SA + 0.000018\times FA(SA)]\times FA(TA) + [0.000692 - 0.000174\times SA - 0.000106\times FA(SA)]\times(4.0\times TA^2 - 3.0)\times TA\} \qquad - (39)$$

where $SA = 2S/150 - 1$

$TA = (2T - 200)/160$

$FA(x) = 2x^2 - 1$

° Viscosity of brine μ [kg/m·hr]

S: Brine concentration, salinity [%]

$$\mu = (SA/10^V) \times 3.6 \qquad -\ (10)$$

where $SA = 1.002 + 0.001652 \times S + 0.0000083 \times S^2$

$$V = (T-20)/(T-109) \times \{1.37060 \times (1 - 0.000619 \times S + 0.000003 \times S^2) + 0.000832 \times (1 + 0.000943 \times S - 0.000015 \times S^2) \times (T-20)\}$$

° Evaporation latent heat of water λ [Kcal/kg]

T: Water temperature [°F]

$$\lambda = (1090.8 - 0.5703 \times T + 1.2819 \times 10^{-4} \times T^2 - 0.8824 \times 10^{-6} \times T^3) \times 0.55546 \qquad -\ (11)$$

° Vapor pressure of saturated water p [kg/cm^2]

T: Vapor temperature [°K]

$$P = 225.65 \times \exp\{-[7.21379 + (1.1520 \times 10^{-5} - 4.787 \times 10^{-9} \times T) \times (T - 483.16)^2] \times (647.31/T - 1.0)\} \qquad -\ (12)$$

° Meaning of symbols

A:	Heat transfer area (m^2)	
C:	Concentration of brine	(weight fraction)
C_B:	Concentration of flash brine	(")
C_b:	Concentration of discharge brine	(")
C_R:	Concentration of circulating brine	(")
i:	Enthalpy (Kcal/kg)	
Q:	Amount of heat used for preheating of brine in heat transfer tube (Kcal/kg)	
Q_c:	Amount of heat released outside by cooling seawater (Kcal/kg)	

Q_h: Amount of heat applied by heating steam (Kcal/kg)
T_B: Temperature of flash brine (°C)
T_{Bmax}: Brine temperature at the outlet of brine heater (max. brine temperature) (°C)
T_b: Temperature of discharge brine (°C)
T_C: Cooling seawater discharge temperature (°C)
T_h: Temperature of heating steam (°C)
T_R: Temperature of recirculating brine (°C)
T_{RC}: Temperature of recirculating brine at the inlet of heat recovery section (°C)
T_S: Temperature of seawater (°C)
T_{smax}; Brine temperature at the inlet of brine heater (°C)
T_V: Steam temperature (condensate temperature) in each stage (°C)
T_W: Temperature of product water in last stage (°C)
U: Overall coefficient of heat transfer (Kcal/m^2·hr·C)
W_B: Flow rate of flash brine (kg/hr)
W_b: Flow rate of discharge brine (kg/hr)
W_C: Flow rate of cooling seawater (kg/hr)
W_h: Flow rate of heating steam (kg/hr)
W_M: Flow rate of make-up seawater (kg/hr)
W_R: Flow rate of recirculating brine (kg/hr)
W_s: Flow rate of seawater (kg/hr)
W_w: Flow rate of product water (kg/hr)

- Design calculation algorism of detailed model

Of the equations mentioned so far, equations (16) through (23) which represent balance equations of the entire system and equations (27) through (31) which represent balance equations and heat transfer rate equations of each stage, play a central role in the design calculation model. Of these, equation (23) represents an energy balance for difference concentrations determined by assumption $T_b = T_c = T_{RC}$ of the design calculation, but its substance is equal to equation (21). Therefore, the core of the design calculation model is 12 equations.

On the other hand, a study of variables used in these equations shows a considerable number of dependent variables which can be obtained from other equations. Enthalpy is a dependent variable of temperature T or concentration C obtained from equations (35) and (36). Moreover, the steam temperature in the flash chamber is assumed to be equal to the temperature of condensed water and the temperature of condensed water T_W becomes a dependent variable of the temperature of flash brine T_B by equations (32) and (33) and related equations of non-equilibrium temperature difference, which will be dealt with in section 5.3.1. Evaporation latent heat λ is a dependent variable of steam temperature by equation (41) and overall heat transfer coefficient U can be obtained separately from the related equations described in section 5.3.2. The number of independent variables used in equations (16) through (22) is 16 and that used in equations (27) through (31) is 11. For design calculation, seven pieces of data - amount of water product W_w, performance ratio (same as W_h if W_w is already given), amount of seawater W_s, concentration of seawater C_s, concentration ratio (same as C_b), maximum brine temperature T_{Bmax} and seawater temperature T_s - are used as design input data.

In the calculation of each stage in equations (27) through (31), variables which are also used in equations (16) through (22) with respect to the I^{th} stage are the following four.

$$W_{BI-1} = W_R, \quad C_{BI-1} = C_R$$
$$T_{BI-1} = T_{Bmax}, \quad T_{RI} = T_{Smax}$$

Moreover, because of the conditional equation of $W_{wI-1} = 0$ and the condition that the flash temperature difference in each stage is constant and that the heat transfer area in each stage is constant, the value of all variables is determined by five equations - equations (27) through (31). If calculation is to be made for the last stage, however, variables which are also used in equations (16) through (22), except for design input data, are the following two.

$$W_{BI} = W_{RC} + Wb \ , \ T_{BI} = Tb$$

From these two variables, which are obtained from the repetition of calculation of each stage until the last stage, and seven pieces of data, all of the seven variables can be determined by seven equations - equations (16) through (22).

In the calculation of each stage, the value of T_{BI} can be obtained directly from the design calculation under the condition of constant flash temperature difference in each stage but a convergent calculation is required under the condition of constant heat transfer area in each stage. An example of the flow chart of calculation in this case is shown in Fig. 5.1-9. Details of the calculation part of each stage are shown separately from the flow of overall calculation.

The calculation program shows an example of cases where the previously mentioned seven pieces of data are provided. More specifically, step 2 reads seven pieces of design data and other necessary data, such as the number of stages in the heat recovery section and heat rejection section, and the program moves to step 3 . Step 3 calculates W_c, W_b from equations (16) and (17) and W_M from equation (20) using the read data. T_b is also calculated from equation (18) and from the relation of $T_c = T_b$. Since the previously mentioned two undertermined variables can be obtained only when the calculation of each stage comes to the last stage, this program assumes T_{smax} of first, completes calculation for the last stage, checks the agreement between the amount of water production obtained by the assumption and the amount of water production to be provided as data and repeats convergent calculation to determine T_{smax}. On the basis of assumed T_{smax}, the value for W_R, W_{RC} and C_R is calculated from equations (19), (21) and (22) and the program moves to calculation on each stage of heat recovery section. In the calculation on each stage in step 17 , such conditions as $W_{WI-1} = 0$, $W_{BI-1} = W_R$, $C_{BL-1} = C_R$, $T_{RI} = T_{smax}$, $T_{BI-1} = T_{Bmax}$ are established for the heat recovery section. Since the heat transfer area of each stage is calculated as constant, the heat transfer area is given as an assumption. For simplification of calculation, the temperature of flash brine T_{B1} in the first stage is assumed and on the basis of this assumption, the value for U is obtained first and the heat transfer area of the first stage A_1 is calculated from equation (31). In the end, convergent calculation is made for each stage so that the heat transfer area of each stage agrees with this A_1 and calculation is made until the last stage to check whether the brine temperature at the inlet of the heat transfer tube T_{RI+1} agrees with T_{RC} in the heat recovery section and T_s in the heat rejection section. If no agreement is found, the assumption of T_{B1} is amended and convergent calculation is repeated to return to step 19 .

If T_{RI+1} converges, the program returns to step 10 and the above calculation for each stage repeats only for the heat rejection section. When the calculation has covered the last stage, agreement between W_{MI} and W_w obtained in the last stage is checked. If no agreement is found, T_{smax} is amended and the program returns to step 5 and calculation is repeated until convergence is found. When convergence is found, the result is typed out with a line printer and judgement is made whether calculation should be made with new data. When calculation is required, the program returns to the start and when calculation is not required, the program comes to an end.

An example of detailed calculation programs under the condition of constant heat transfer area in each stage is shown in Fig. 5.1-10. In the figure, the result of detailed calculation of maximum brine temperature and performance ratio combined with economic data is plotted in each case. After all, details of optimization can be obtained if the maximum brine temperature is determined and a plot like the one shown in Fig. 5.1-10 is provided for each variable around the optimum point determined by the simplified model. This plot assumes that the steam cost and the cost of pretreatment remain unchanged even when there is an increase of maximum brine temperature.

For a plant design thus optimized, various flow rates, temperature requirements, and the temperature, concentration, flow rate and amount of water product in each stage are determined (automatically). From the determined result, the distribution of heat transfer area per unit quantity of condensation and of overall heat transfer coefficient against the temperature is obtained as shown in Fig. 5.1-11. The figure provides two types of fouling factor 0.0002 and 0.00015. It shows that the smaller fouling factor contributes to the increase of heat transfer coefficient and the decrease of heat transfer area per unit quantity of condensed water. And the heat transfer coefficient increases with an increase of temperature. It is also known that the heat transfer area per unit quantity of condensed water increases drastically with an decrease of temperature in the temperature range of below 60°C. From this figure, it is possible to determine an approximate heat transfer coefficient of the flash distillation plant and an approximate value of heat transfer area required for the water production capacity.

Since the figure indicates a distribution of overall heat transfer coefficient against the temperature, the existance of the optimum distribution of heat transfer area, the minimum heat transfer area for a certain amount of water production, can be estimated.
In a large multistage flash distillation plant, the cost of heat transfer tubes is said to account for about 40 percent of the total construction cost and there is a requirement for reducing the heat transfer area as much as possible. Against this, Beamer et al. have reported, on the basis of their study with

a simplified model, that a reduction of water production cost by 1 percent is possible if the heat transfer area of each stage is maintained constant. Sato et al. have reported, on the basis of their findings with a detailed model, that the saving of heat transfer area through the use of optimum distribution is very small and much cannot be expected. They further stated, after a comparison of two calculation methods - constant flash temperature difference in each stage and constant heat transfer area in each stage, that there is no difference between the two calculation methods in the required total heat transfer area but there is a big difference in the temperature of each stage and have suggested that selection of the method should be made according to the purpose of calculation.

After the completion of detailed design calculation, the result is summarized as a process flow and then the layout of each section of the plant is prepared. Fig. 5.1-12 is a process flow of a multistage flash distillation plant having a capacity of 100,000 t/day, with a performance ratio of 12, 40 stages in the heat recovery section, 3 stages in the heat rejection section, seawater temperature of 24°C, discharge brine temperature of 32°C, maximum brine temperature of 121°C and a concentration of 2.0. Flow rate and temperature of main flows are indicated in the chart. From the chart, it is known under what conditions the main components of the plant operate and what would be the appropriate capacity of these components. It also indicates that especially the seawater supply pump and brine recirculating pump should have a larger capacity as compared with other pumps. For this reason, use of a turbine driven pump is being considered, in addition to a motor driven pump for a large plant as shown in Table 5.1-1. In such a case, the steam used for the turbine is used again as part of the heating steam.

After the operating condition of the main components have been determined as shown in Fig. 5.1-12, specifications of each piece of equipment are prepared on the basis of these operating conditions. From the specifications, materials, manufacturing and fabrication costs of each piece of equipment are calculated and with the addition of operating costs, an accurate production cost per ton of fresh water is calculated. If necessary, amendment is made to Fig. 5.1-10 and the final and optimum design of the plant is established.

5.1.3. Operating Characteristics of the Process

(1) Control system

The main instrumentation system of a multistage flash distillation plant is shown in Fig. 5.1-13. Since the orifice in each stage, is fixed the brine levels in the flash chamber are maintained constant by control of the brine level of the last stage with the valve located in the path of discharge brine which is operated by the liquid level indicator controller (LIC). The maximum brine temperature is controlled at a set value through regulation of the flow rate of heating steam by control valve which is operated by the temperature recorder control (TRC). The flow rate of recirculating brine is controlled at a set value by control valve which is operated by the flow recorder controller (FRC). The location of control valve is determined in the design stage so as to prevent a two phase flow of recirculating brine in the heat transfer tube. Since recirculating brine boils at a pressure of about 2 kg/cm^2 at the maximum brine temperature of 121°C, the pressure of recirculating brine before the orifice of the first stage must be higher than 2 kg/cm^2. Otherwise, the recirculating brine boils in the tube of the brine heater, thereby causing scale deposition trouble, or the temperature will not rise to the set value, or a value which is completely different from the design value is obtained because of orifice characteristics of two-phase flow, and the operation of the plant under specific conditions becomes impossible. This pressure is provided by the discharge pressure of the brine recirculating pump.

Therefore, study must be made as to the diameter and location of the control valve and the design of the orifice of the first stage must be based on this study. If the control valve is located before the brine heater in the path of recirculating brine, care should be exercised when increasing the maximum brine temperature at start-up of the plant when there is a small flow of circulating brine.

(2) Adjustment and start-up of plant

Various adjustments are required before the start-up of a multistage flash distillation plant. Equipment which requires adjustment are mainly the orifices, vacuum system, condensed fresh water tray connecting pipe and control equipment.

When a multistage flash distillation plant is constructed, a leak test of vessels is required first of all and then the opening area of orifices, valves on venting tubes or venting orifices between stages and orifices on the condensate fresh water tray connecting tube are set to the design value, which is determined by a detailed plant design, and the plant is put into a break-in operation. During the break-in operation, the opening degree of orifices or valves in the vacuum system is adjusted and set to the design value while the temperature and flow rate in main stages and tubes are being checked. After the level of flash brine and condensed fresh water in each stage have been checked with level gauges such as glass level gauges, the plant is shut down and the opening degree of orifices is adjusted. The level of flash brine and condensed fresh water are adjusted to the specified range of design liquid level with repetition of this operation if necessary. When continuous operation for several hours becomes possible, PID action of control equipment is adjusted under a condition of steady continuous operation using an optimum adjustment method such as the Ziegler-Nichols method or the transient response method.

Upon completion of the break-in operation, the plant is ready for continuous operation. The start-up procedure of the plant in this case is as follows.

(1) Switch on power sources of all recorders and set temperature recorder-controller (TRC) and flow recorder-controller (FRC) shown in Fig. 5.1-13 to manual operation.

(2) Start the seawater supply pump, open the valve on the seawater supply pipe connected to the flash chamber and fill the flash chamber with seawater to the specified level.

(3) Supply steam to the steam ejector and vent the air in the flash chamber.

(4) Start the brine recirculating pump and supply brine little by little.

(5) After confirming that the liquid level control of the last stage is set to automatic operation, start the discharge brine pump.

(6) At the same time, start the make-up seawater pump and supply a small amount of make-up seawater to pretreatment equipment.

(7) Check the degree of vacuum in the flash chamber and if the reading is above the specified value (approximately - 600 mmHg), start supply of heating steam to the brine heater little by little.

(8) Supply sulfuric acid to pretreatment equipment and set the pH controller to automatic operation.

(9) Start the product water pump.

(10) When there is a rise of the degree of vacuum in the last stage, increase the flow of recirculating brine and heating steam gradually. Flash distillation (evaporation) begings in sequence starting from the first stage with the rise of maximum brine temperature.

(11) Increase the amount of make-up seawater to keep pace with the preceding step and set the make-up water flow controller to automatic operation.

(12) When the degree of vacuum in the last stage has reached the set value after repetition of step (10), increase the flow of recirculating brine and maximum brine temperature to a value close to the set value of steady operation and set FRC and then TRC to automatic operation.

In following the above procedure, it must always be kept in mind that the maximum brine temperature should never exceed 121°C and that the brine level in each stage should not exceed a certain standard level. If the maximum brine temperature exceeds 121°C, deposition of hard scale on the interior walls of heat transfer tubes is very likely, as mentioned in the following section. Once the deposition of hard scale occurs, it becomes a nucleus and contributes to continuous deposition of hard scale even when subsequent steady operation maintains the maximum brine temperature below 121°C. The rise of brine level causes a delay in the improvement of the purity of product water. A sharp rise of brine level causes immersion of heat transfer tubes and decreases heat recovery efficiency drastically, thereby putting an extra burden on the brine heater, which results in a continuous decrease of maximum brine temperature. As a result, the brine level keeps rising more and more in a vicious cycle and the operation may have to be suspended in some cases.

The time required from start-up to steady operation is determined by the plant capacity, ejector capacity, heating steam supply capacity and brine heater capacity. Since the capacity of the ejector generally has only a small margin over the required capacity at steady operation, the time required for attaining a specific degree of vacuum influences overall time requirement. When a concrete vessel is used, however, considerable time is allowed for the increase of maximum brine temperature to prevent cracking of the vessel by thermal stress as described in detail in section 5.5. This factor determines the time required from start-up to steady operation.

The procedure for shutting down is the reverse of the start-up procedure and no special care is required. In the case of concrete vessels, however, time constraints for decreasing the temperature will be required.

(3) Steady continuous operation

At the initial stage of plant operation, heat transfer tubes are clean and have a high heat transfer coefficient. At a temperature above 100°C, the heat transfer tube has a heat transfer coefficient of about 3,500 $Kcal/m^2 \cdot hr \cdot °C$, which is considerably high as compared with the value shown in Fig. 5.1-11. Accordingly, the plant efficiency (performance ratio) becomes high and temperature distribution in each stage is quite different from that of the design calculation which adopts fouling factors of 0.00015 and 0.0002 shown in Fig. 5.1-11. If the plant is operated continuously for several hundred hours under this condition, however, heat transfer tubes become fouled gradually with a deposition of sludge or scale and begin to show a fouling factor close to the design value. In the design calculation, however, it is common to apply a constant fouling factor for all stages of heat recovery section and heat rejection section. On the other hand, the sludge, for example, which is the actual cause of the increase of fouling factor, deposits mainly inside the heat transfer tubes of low temperature stages.
The fouling factor generally shows a distribution and even when the average fouling factor of the plant comes close to the design value, temperature distribution of each stage does not necessarily agree with the design value. Accordingly, the liquid level does not coincide with the design value. It is natural that the liquid level, in the initial stage of operation when the fouling factor is small, is different from that of a later stage when the fouling factor becomes larger. Adjustment of orifice during the break-in operation, therefore, must be performed with this fact taken into consideration.

As the fouling factor becomes greater, plant efficiency decreases accordingly and it becomes necessary to clean heat transfer tubes to improve the overall heat transfer coefficient. In a large plant equipped with ball cleaning equipment, heat transfer tubes are cleaned with sponge balls. Ikazaki et al. have reported on the optimum cleaning operation and have stated (1) that the controlling factors of sponge ball cleaning are, the ball cost for a short cleaning cycle, and the steam cost for a long cleaning cycle and that the optimum cleaning cycle is when these two factors have nearly the same share in the appraisal function; and (2) have also clarified a general characteristic that increase of the number of balls and decrease of the frequency of ball circulation produces good results.

Steady continuous operation lasts for about one week at the shortest and about 300 days at the longest. During this operation, such periodic checks and maintenance as cleaning of electrodes of the pH meter in the pretreatment equipment, inspection of control valves as well as the instrumentation air supply system and check of various instruments including controllers, check of water quality and measurement of dissolved oxygen in make-up seawater and recirculating brine are required. It is especially important to make careful inspection of electrical components which tend to develop such faults as poor contact and short circuits caused by salt, as the plant is normally constructed close to the sea.

5.1.4. Partial Load Operation and Dynamic Characteristics

(1) Effect of external disturbance and load fluctuations

In order to maintain stable and continuous operation of a multistage flash distillation plant, it is essential to acquire full knowledge of the behavior of the plant against external disturbances such as a fault of the heating steam system. This is because such knowledge is fundamental for control of the plant and is the basis for unrestricted operations of the plant including start-up, shut down or partial load operation and for coping with unforeseen developments, such as accidents in operation.

Here, the behavior of a multistage flash distillation plant against the change of maximum brine temperature and flow rate of recirculating brine will be studied qualitatively as a first step.

When a multistage flash distillation plant is bothered by some external disturbance, the element mostly affected is the liquid level of each stage, which is controlled only at the last stage as shown in Fig. 5.1-13. To facilitate study on the liquid level, a schematic diagram of energy balance relations around the orifice is given in Fig. 5.1-14. If the energy balance is taken with attention paid to the flash brine between these two stages, equation (43) holds. The flash brine in i^{th} stage is assumed to have finished evaporation and moved close to the orifice of $i+1$st stage.

$$\frac{g}{g_c} L_i + \frac{\bar{u}_i^2}{2g_c} + JE_i + \frac{P_i}{\rho_i} = \frac{g}{g_c} L_{i+1} + \frac{\bar{u}_{i+1}^2}{2g_c} + JE_{i+1} + \frac{P_{i+1}}{\rho_{i+1}} + \Delta F_{i+1} + q_{i+1} \qquad -\ (43)$$

where L: Liquid level

$\bar{u}$: Average flow rate of liquid

E: Internal energy of liquid

P: Pressure (saturated steam pressure)

g: Acceleration of gravity

gc: Conversion factor

J: Energy conversion factor

ΔF: Energy loss by orifice

ρ: Liquid density

q: Work to outside

Since the energy, equivalent of temperature difference between i^{th} stage and i + 1st stage, gives out heat outside the system as steam

$$q_{i+1} = J (E_i - E_{i+1}) \qquad - (44)$$

If $\rho_i = \rho_{i+1} = \rho$ is taken and assumption is made that the brine flows steadily and that the flow rate in each stage is constant, it can be summarized as shown in the following equation (45).

$$(L_i - L_{i+1}) + \frac{1}{2g} \{ (\frac{L_{i+1}}{L_i})^2 - 1 \} \bar{u}_{i+1}^2 + \frac{g_c}{g} (P_i - P_{i+1}) / \rho$$

$$= \frac{g_c}{g} \triangle F_{i+1} \qquad - (45)$$

In equation (45), $\bar{u}^2$ is normally less than 1 m/sec. and the term of $\bar{u}^2$ is smaller than other terms. When it is ignored, this equation provides equation (46).

$$(L_i - L_{i+1}) + \frac{g_c}{g} (P_i - P_{i+1}) / \rho = \frac{g_c}{g} \triangle F_{i+1} \qquad - (46)$$

Assuming that the pressure loss by orifice is proportional to the square of the speed of flow through opening of orifice, it may be expressed as follows.

$$g_c / g \cdot \triangle F_{i+1} = \frac{1}{K} [\frac{1}{2g} (W / A_0)^2] \qquad - (47)$$

In the above, W is the flow rate of fluid volume at the opening of the orifice, A_0 is the opening area of the orifice and K is the proportional factor.

From equations (46) and (47), the following equation (48) is derived.

$$W / A_0 = K (2g \cdot \triangle P_T)^{1/2} \qquad - (48)$$

where

$$\triangle P_T = (L_i - L_{i+1}) + \frac{g_c}{g} (P_i - P_{i+1}) / \rho$$

Equation (48) is known as a related equation of orifice pressure less, in which K represents the orifice factor.

After all, the liquid level is eventually controlled by these equations. Moreover, the pressure is fixed at both ends by maximum brine temperature and by the pressure of the last stage, and the liquid level is bound by the condition that only the liquid level of the last stage is controlled. Now, let us see how the liquid level fluctuates against the change of maximum brine temperature and flow rate of recirculating brine under these conditions.

If the amount of heating steam decreases and maximum brine temperature decreases under steady operating conditions when the flow rate of recirculating brine remains constant, the resistance of the orifice becomes constant but P_i, saturated steam pressure decreases. P_{i+1} also decreases and $(P_i - P_{i+1})$ becomes small in equation (46), as the flash temperature range of the entire system decreases with the decrease of maximum brine temperature. Therefore, $(L_i - L_{i+1})$ becomes greater and there is a rise of L_i. When maximum brine temperature increases, L_i lowers on the contrary.

Next is the case where the flow rate of recirculating brine decreases under the condition that the maximum brine temperature is maintained constant. As the pressure loss by orifice changes in proportion to the square of the speed of flow, the right number of equation (46) becomes smaller. On the other hand, the temperature difference between stages becomes greater, as the heat transfer area per unit flow rate increases with the decrease of flow rate of recirculating brine. In other words, $(P_i - P_{i+1})$ becomes greater. Accordingly, $(L_i - L_{i+1})$ must be small and L_i lowers, eventually with the decrease of flow rate of circulating brine. If the flow rate of recirculating brine increases, L_i rises on the contrary.

Though the fluctuation of liquid level is small between stages, it accumulates in each stage because of the characteristic of the process that the liquid level is controlled only at the last stage, and as a result, the fluctuation on the side of high temperature stages becomes exceedingly large. An excessive drop of liquid level causes a blow-bye of steam to the next stage through the opening of the orifice, resulting in a decrease of plant efficiency and a rise of liquid level accelerates mixing of splashes with product water resulting in deterioration of the purity of the fresh water. An excessive rise of liquid level causes submersion of heat transfer tubes, which results in a drastic decrease of heat transfer efficiency, obstruction of heat recovery and a decrease of brine temperature at the inlet of the brine heater. If the brine heater has no room in its capacity, a decrease of brine temperature at the inlet causes a decrease of maximum brine temperature and a further rise of liquid level in a vicious cycle. It may result in suspension of plant operation eventually.

It is important, therefore, to keep a close watch on the behavior of liquid level against external disturbance and load fluctuations of the multistage flash distillation plant.

(2) Partial load operation

The concept of a large multistage flash distillation plant of the 100,000 m^3/day class developed under the large development project of the Agency of Industrial Science and Technology is considered to be applied to practical use as a dual purpose plant combined with a power generating plant. In such a case, it will be necessary for the power generation side to adjust energy production according to the change of power demands and a change of the amount of heating steam supplied to the water production side must be expected. Since the operation of the water production plant is determined in relation to river flow, the amount of water production is not always constant. At any rate, stable and efficient partial load operation of the water production plant is very convenient and beneficial. It may be said that operation of the plant with changing loads is to some extent a necessity.

In this respect, research was done by Hayakawa et al. and Sato et al. Hayakawa et al. clarified the change of various state values of partial load as shown in Fig. 5.1-15 in the simulation of a 10-stage plant and reported that operation at partial load is very limited, unless adjustment of the opening area of orifices is made, because of the flow condition of brine in the chamber and restriction on liquid level. On the other hand, Sato et al. simulated a 39-stage flash distillation plant and compared the result with the result of a test plant having a capacity of 3,000 m^3/day. Fig. 5.1-16 is an example of the test result, which shows the temperature distribution of flash brine in each stage obtained from the simulation and the result of tests which were carried out under the same conditions. In the chart, the axis of ordinates shows a temperature difference as compared with the result of simulation at full load operation. Of special interest is that both results show a sudden decrease of temperature difference in low temperature stages. The temperature difference between stages at that time is about 2°C, which is equal to the temperature difference between stages at full load operation. This indicates that there is no flash occurring at this point. The difference in the location of the non-effect stage without flash between the simulation and the test result may be explained by the special characteristics of the test plant in that it has a distribution of fouling factors, while the simulation assumes it to be constant. The difference in temperature distribution between the two results is due to the fact that the simulation obtains an orifice factor of equation (48) at a full load operation and considers it as constant, but the orifice factor actually changes because of the change in the state of two-phase fluid flowing through the orifice depending on the state of flash.

The effect of a change of maximum brine temperature T_{Bmax} on liquid level distribution at partial load operation and the effect of a change of the flow rate of recirculating brine W_R on liquid level determined in the simulation are shown in Figs. 5.1-17 and 5.1-18, respectively.

Fig. 5.1-19 shows a liquid (brine) level in the 8th stage, which had the highest fluctuation, as determined by the test result.

The result of the simulation supports the behavior of liquid level against a change of maximum brine temperature and the flow rate of recirculating brine described in (b) of the test result. In other words, a decrease of maximum brine temperature or an increase of the flow rate of recirculating brine raises the liquid level as a whole and an increase of maximum brine temperature or a decrease of the flow rate of recirculating brine lowers the liquid level. It was found that in the stage where no flash occurs, the liquid level rises sharply as compared with the reference liquid level of the last stage because only the difference of liquid level causes flash brine to flow in such a stage.

The result of the simulation also shows that the maximum point of liquid level is located around the 8th stage to 12th stage and that the stages in this neighborhood have the largest fluctuation of liquid level as compared with the reference liquid level of the last stage. This coincides with the largest fluctuation of liquid level in the 8th stage revealed by the test, and it may be said that the liquid level in the stage of this neighborhood provides a limiting factor for stable normal operation of the plant.

The disagreement between (a) and (b) of Fig. 5.1-19 and the discontinuous change of liquid level in (b) are also considered to have been influenced by the orifice factor which changes with a small change of flow rate, and direction of increasing and decreasing flow rate.

Sato et al. prepared Table 5.1-2 from the results of the simulation and reported that since partial load operation allows room in the heat transfer area, it should increase the performance ratio, but in actuality plant efficiency decreases drastically as shown in the table because of the occurrence of a non-effect stage. To solve this problem, several methods, including a change of the opening area of the orifices and prevention of flash in the high temperature stages, have been suggested. Application of these methods, however, will have to depend on future development.

(3) Dynamic characteristics (Dynamics)

As discussed in the previous item (1), knowledge of the behavior of the plant against external disturbance is a basic requirement for control and operation of the plant and is indispensable for the design of control systems and stable continuous operation of the plant.

In this respect, research on the multistage flash distillation plant was made by simulation and experiments with a test plant. In simulation, Delene et al. determined the effect of a 10 second 10°F pulse increase in brine heater steam temperature on the brine level of the first stage as shown in Fig. 5.1-20 and the response of tray brine levels to a 2°F step increase in brine heater steam temperature as shown in Fig. 5.1-21. It is estimated from these charts that the brine level has a quick response and reaches a constant value in a relatively short time and that the brine level shows a large fluctuation in high temperature stages.

Sato et al., in the meantime, conducted a dynamic characteristic test with a 39-stage test plant having a capacity of 3,000 m^3/day by expanding the range to include partial load operation, and obtained the frequency transfer function on the Bode's diagram by means of Fourier transformation of test data. Fig. 5.1-22 shows responses of brine level in the 8th stage, which shows the largest fluctuation, to a step change of maximum brine temperature. Examples of Bode's diagram obtained from these responses are given in Figs. 5.1-23 and 5.1-24. Fig. 5.1-23 is the result of a step decrease of maximum brine temperature, which indicates characteristics that the gain reaches a peak at low frequency exceeding 0 dB and that the gain drops once, but it reverses its direction and reaches a peak at high frequency. The peak on the high frequency side is also shown in Fig. 5.1-24 which gives the result when the flow rate of circulating brine decreases in steps.

From these results, they pointed out as dynamic characteristics of the process toward partial load conditions, (1) that the plant characteristics differ depending on whether the input pattern is a step increase or a step decrease; (2) that the frequency transfer function has its trough and peak and cannot be expressed by a simple degree delay system; (3) that attention should be paid only to the characteristics of a frequency band of less than 1.0 - 1.5 rad/min. in this system, and (4) that either maximum brine temperature or the flow rate of recirculating brine may become an operating variable against the external disturbance of the other.

It should be noted that phenomenon (1) is also clearly indicated in Fig. 5.1-19, which is the test result of partial loading, one of the dynamic characteristics, and which is not considered in the simulation, and that there is a considerable difference in response between Fig. 5.1-21 and Fig. 5.1-22 even though the conditions are different between the two cases. Further development of research in this field, including a study of orific characteristics, is highly hoped for at a time when more accurate data on dynamic characteristics are required for practical application of the multistage flash distillation process.

5.2. Pretreatment

Scaling on the heating tubes and corrosion on the heating tubes and the vessel by seawater are problematical for stable operation of desalination equipment operating in the evaporation process.

Insoluble salts in the seawater, calcium carbonate or magnesium hydroxide for example, separate dut to heating and concentrate the seawater. Where scale sticks to the heating tube, the function of heat transfer is greatly hindered and the heat economy of the equipment largely deteriorates. Therefore, the matter of how to prevent scaling must be regarded as serious for desalination equipment. Decarbonation and a pH controlling process by acid addition are generally effective. In this case the decarbonation technique becomes necessary for pretreatment of the seawater.

Corrosion on metal materials by seawater varies according to the temperature, concentration, pH, etc., of the seawater, however, a more important factor is dissolved oxygen. A deaeration technique to eliminate dissolved oxygen from seawater is indispensable for seawater pretreatment to keep corrosion from arising.

5.2.1. Scale Prevention

(1) Composition of seawater and scale

As shown in Table 5.2-1, the total dissolved solid component is about 3.5 Wt percent for composition of normal seawater. Sodium Na^+ and chlorine $C\ell^-$ occupy about 86% of the total dissolved components, and then calcium Ca^{2+}, magnesium Mg^{2+}, sulfuric acid SO_4^{2-}, bicarbonate ions HCO_3^-, etc. are present in s minute amount. In normal evaporation process desalination equipment, the concentration ratio of seawater is 2 - 3 times, and hence sodium chloride $NaC\ell$ does not separate. The matter separating as scale is a crystalline deposit of insoluble salts like calcium carbonate, magnesium hydroxide, calcium sulfate, etc. which are dissolved in a very small amount.

The contaminants sticking to the heating tubes of the evaporation process desalination equipment are classified, as shown in Table 5.2-2, into scale (a deposit of insoluble salt) and sludge (settlings of suspended particles in the brine). What is problematical as a contaminant is scale. However, sludge is harmful in hindering heat transfer during a long operation, and its detrimental influence can be compared with scale.

Since the study on how to prevent contamination has dealt with scale only so far, a correlativity between sludge and scale in the contaminating factors has not yet been thoroughly elucidated. In view of reducing the desalination cost, it is desirable to minimize the coefficient of contamination on the heating tubes, thus operating the plant continuously for long period of time. An explanation of how sludge plays part in contamination, which has been neglected so far, and effective means for elimination will thus be necessary.

There are two kinds of scale. One is sulfuric acid scale in which calcium sulfate works as the principal component, and the other is alkaline scale consisting of magnesium hydroxide.

Sludge is settled material with various matter mixed in it, and its principal constitutuents are fine soil and sand and organic matter suspended in the seawater supplied to the plant, and corrosive product of the components of the equipment like iron and copper.

(2) Contaminant forming mechanism

(a) Sulfuric acid scale forming mechanism

Sulfuric acid scale is a deposit of anhydride $CaSO_4$, hemihydrate $CaSO_4 \cdot 0.5H_2O$ and dihydrate $CaSO_4 \cdot 2H_2O$ which are three variations of calcium sulfate. Sulfuric acid scale is sometimes called hard scale. It is insoluble with acid and is very hard to eliminate when sticking, and thus there have been many cases in the past where the heating tubes had to be replaced.

Saturation solubility of calcium sulfate changes according to the temperature and concentration of the seawater. Three kinds of crystal forming limits are shown in Fig. 5.2-1. As will be apparent from the figure, limits for operating conditions of the plant are determined by dihydrate in the case of low temperature and high concentration and by anhydride in the case of high temperature and low concentration. Operating results of the multi-effective evaporator shell and the multistage flash evaporator are also given in Fig. 5.2-1; anhydride in the high temperature zone is strong in supersaturation and hence is separated in the multi-effective evaporator shell having a recycle of brine at the same temperature in it, but not separated in the multistage flash evaporator with the temperature changing periodically in accordance with recycling. Therefore, operating conditions of the multistage flash evaporator can be restricted by the separating limit of hemihydrate.

(b) Alkaline scale forming mechanism

Alkali scale is also called soft scale and consists of calcium carbonate $CaCO_3$ and magnesium hydroxide $Mg(OH)_2$. Alkaline scale can be eliminated through pickling and hence is less troublesome than sulfuric acid scale.

The concentration of the seawater can be pointed out as a cause of alkaline scale separating. However, usually it is caused by the temperature change in the chemical equilibrium of the scale component in the seawater.

As illustrated in Fig. 5.2-2, bicarbonate HCO_3^- changes into carbon dioxide CO_2 and carbonate CO_3^{2-} from heating the seawater, and calcium carbonate $CaCO_3$ is formed beyond saturation solubility.

$$2HCO_3^- \rightarrow CO_2\uparrow + CO_3^{2-} + H_2O \tag{1}$$

$$Ca^{2+} + CO_3^{2-} \rightarrow CaCO_3\downarrow \tag{2}$$

From further heating over 80°C carbonate CO_3^{2-} changes into carbon dioxide CO_2 and hydroxyl OH^- through hydrolysis, and formation of magnesium hydroxide $Mg(OH)_2$ governs.

$$CO_3^{2-} + H_2O \rightarrow CO_2\uparrow + 2OH^- \tag{3}$$

$$Mg^{2+} + 2OH^- \rightarrow Mg(OH)_2\downarrow \tag{4}$$

(c) Sludge forming mechanism

The principal component of sludge is corrosive products of components of the equipment. The evaporator shell is made normally of steel material, therefore oxide of iron is overwhelming. The status of oxide of iron comes in ferric oxide Fe_2O_3 and triiron tetraoxide Fe_3O_4 in most cases. Then, where copper alloy is used for the heating tubes, oxide of copper is contained in a minute amount. Besides the above, fine sand and soil and organic matter flow into the plant along with the seawater supplied. Soil and sand form silica sludge SiO_2.

(3) Contamination prevention

(a) Prevention methods for sulfuric acid scale

For methods to prevent sulfuric acid scale, the softening method, solubility increasing method, supersaturation maintaining method, crystallization accelerating method, mechanical eliminating method, etc. are taken up, however, none of them has been found reliable. Since the separating limit is determined according to the temperature and concentration of the seawater, as shown in Fig. 5.2-1, for sulfuric acid scale, the present situation is such that formation of sulfuric acid scale is avoided by controlling temperature and concentration conditions of the plant in the area where sulfuric acid scale is not separated.

(b) Prevention method for alkaline scale

Various methods are applicable for prevention of alkaline scale, however, the typical methods are the following three:

(i) Decarbonation and pH control method

This is a method comprising a method to eliminate the scale components through pretreatment and to change chemically the equilibrium of scale components.

(ii) Inhibitor adding method

The method is to prevent crystallization by maintaining the supersaturation property of the scale components by chemicals, or diffusing the scale components.

(iii) Seed adding method

The method is to prevent scaling on the heating surface by securing the formation of scale components on a seed given as a nucleus.

The above decarbonation and pH controlling method and inhibitor adding method are effective against scale but not against sludge. The seed adding method is considered to be effective against both scale and sludge.

(c) Sludge eliminating method

For elimination of sludge the sponge ball washing method which works mechanically is effective. Its effectiveness has been recognized extensively of late, and the method is now being employed in practice.

The sponge ball washing method was developed for washing the heating tubes for thermal generating condensers, and the development of a ball which is serviceable even at 130°C max. has made possible application on desalination equipment.

The above contamination preventing methods can be classified as shown in Table 5.2-3. Prevention of contamination can now be expected more effectively from using the sponge ball washing method available for sludge elimination together with the pH controlling method or inhibitor adding method for scale prevention.

(4) Effect of various scale preventing methods

The following is to amplify upon the effect of various scale preventing methods in accordance with the result of a scale preventing test carried out as a link of the large project study "Desalination and By-Product Utilization" by the Ministry of International Trade and Industry.

The testing device was a 100 m^3/day recyclic 10-stage flash evaporator. Externals of the testing device are given in Fig. 5.2-3; a system diagram is shown in Fig. 5.2-4.

The flash evaporator consists of 10 stages all told: 8 stages for the heat recovery section and 2 stages for the heat rejection section. The condenser on each stage is of short tube type, and the heating tube is installed through a detachable expansion pipe system. Therefore, the heating tube can be pulled off immediately after continuous operation to measure for contamination.

The following equipment was arranged for prevention of contamination on the heating surface.

- Decarbonation and pH controlling method: Acid injection device and decarbonation device (supplied seawater system)
- Inhibitor adding method: Inhibitor injection device (supplied seawater system)
- Seed adding method: Seed recovery device (discharged brine system) and seed supply device (supplied seawater system)
- Sponge ball washing method: Ball circulation device (circulating brine system)

For comparison and evaluation of the effects of various scale preventing methods, fundamental testing conditions were specified as follows:

Desalination rate: 100 m^3/day
Performance ratio: 4.5
Brine max. temperature: 120°C
Brine min. temperature: 45°C
Circulating brine concentration ratio: 2.0
Operating hours: Short: 200 hr
Long: 2,000 hr

(a) No treatment test

As a blank test for performance evaluation of various scale preventing methods, a treatment test free from any preventing method was carried out.

Coefficients of overall heat transfer of the brine heater and the condenser deteriorate as time passes, as shown in Fig. 5.2-5, indicating severe contamination.

The result obtained through measuring the scaling rate is shown in Fig. 5.2-6. The scaling rate is at its maximum value on the condenser working at 90°C or so in brine temperature. Composition of the scale is as shown in Fig. 5.2-7. As will be estimated from the equilibrium relation of the scale components in the seawater, scale is calcium carbonate in the low temperature zone and magnesium hydroxide in the high temperature zone.

(b) Decarbonation and pH controlling method

Bicarbonate HCO_3^- changes to carbon dioxide in seawater by adding acid. (Fig. 5.2-2)

$$H^+ + HCO_3^- \rightarrow CO_2\uparrow + H_2O \quad \cdots\cdots\cdots\cdots \quad (5)$$

By eliminating the carbon dioxide in the decarbonation device and supplying the seawater with carbonate decreased in the plant, calcium carbonate will not be formed even through heating and concentration. The pH of the seawater is normally 8.3 or so, and by adjusting the pH of the brine to 7.0 or so through adding acid, a separation of magnesium hydroxide can be suppressed.

As the result of having carried out a 200-hour test on the 100 m^3/day plant, the effect of scale prevention was good in the pH range 7.0 - 7.4, and the scaling rate was below 0.01 $g/m^2 \cdot h$. Furthermore, a 2,000-hour continuous operation test was carried out at brine pH = 7.2 ± 0.2 for pH controlling conditions.

The scaling rate in the said test is shown in Fig. 5.2-8. The heating tube was contaminated on the surface due to the test extending long time. The contaminant was mainly sludge and as the result of chemical analysis and X-ray diffraction, sludge on the condenser section was found to be magnetite Fe_3O_4, which was a corrosive product of the shell steel material. On the brine heater sludge was stuck, and moreover anhydrous calcium sulfate was separated.

Decarbonation and pH controlling methods are very effective for prevention of alkali scaling. However, sludge which is a corrosive product, sticks more onto the shell by lowering pH values. Therefore, a sludge eliminating method must be used at the same time to maintain high efficiency for a long time. Furthermore, corrosion is accelerated increasingly as dissolved oxygen is highly contained in the circulating brine, therefore decarbonation of the seawater supplied to the plant must be coordinated with keeping dissolved oxygen concentration in the circulating brine below 10 ppb and also with completely preventing air from flowing into the plant.

(c) Inhibitor adding method

What maintains supersaturation properties of the scale components and changes separated scale into sludge is the inhibitor. Condensed phosphate is well known generally as an inhibitor. The inhibitor adding method is practically used in the Middle Eastern countries.

Scaling in the low temperature zone becomes about 1/100 of that arising where treatment is not applied from adding condensed phosphate at 5 ppm or so to the supplied seawater; the higher the temperature gets, the more scaling increases, and the effect is almost not expected at 90°C. Scale on the heating tube is muddy. While the scale is left in a hard plate condition and increases in thickness in proportion to the time under other preventing methods applied, scale, when highly deposited, will be washed off by the brine in the case of an inhibitor, and thus the more time passes, the slower the scaling will increase.

A shortcoming of poli-phosphate is that hydrolysis is unavoidable at more than 90°C and thus the scale preventing effect is ineffective. Research has actively been done on the inhibitor recently, and inhibitors such as the polyacrylic soda group and maleic acid group have been developed.

(d) Standard crystal adding method

The standard crystal adding method prevents scaling on the heating surface through separating the scale components in supersaturation with standard crystals working as nuclei. The standard crystal adding method has been applied against acid scale in the salt manufacturing industry in Japan for a long time, and remarkable achievements have been obtained. A standard crystal of the same scale components is better, and that with scale coming out of the plant used as a standard crystal is particularly effective.

For prevention of alkaline scale a standard crystal obtained through mixing calcium carbonate and magnesium hydroxide is effective. The test result shows that the physical and chemical effect in the growth of standard crystals is not satisfactory enough, and it is necessary to eliminate the scale on the heating tube mechanically with friction by the standard crystal, for which the standard crystal concentration must be improved up to percentage order.

Where physical and chemical effect is available jointly with mechanical effect, the scale preventing effect comes substantially near to the decarbonation and pH controlling methods and is also characterized by keeping corrosion down. However, a matter to settle before practical application is how to recover the standard crystal. As shown in the system of Fig. 5.2-4, a settling tank is used for standard crystal recovery equipment, however, it requires fairly large arrangements which increase the cost, offsets the advantage of the seed adding method. Development of miniaturized standard crystal recovery equipment which is high in efficiency must precede practicability of the standard crystal adding method.

(e) Sponge ball washing method

A ball circulating system like that of Fig. 5.2-4 will be arranged for application of the sponge ball washing method to the multistage flash evaporator. The effect of ball washing against alkali scale is not satisfactory, as the scaling rate can be retarded to barely 1/10 of that in no-treatment. However, sponge ball washing is effective for elimination of sludge and scale where the sticking rate is low, and joint use with scale prevention methods will offer an effective means.

(f) Compound method of decarbonation/pH controlling and sponge ball washing

Long operation under only the decarbonation and pH control methods will lower thermal efficiency by sludge sticking. The sponge ball washing method effective for elimination of sludge was jointly used.

The test results were very satisfactory as illustrated in Fig. 5.2-8: the value of coefficient of overall heat transfer after 2,000 hours remained almost unchanged from that of the beginning, and the method was very effective for prevention of sludge and scale from sticking.

(g) Compound method of inhibitor adding and sponge ball washing

The inhibitor adding method changes scale into sludge, therefore an effective scale preventing method is obtainable through eliminating it by sponge ball washing.

According to the test result, scaling could be decreased to 1/10 by carrying out ball washing continuously.

(5) Scaling and the fouling factor

(a) Coefficient of overall heat transmission of evaporating stage and coefficient of contamination

Heat exchange of the condenser on each stage of the multistage flash evaporation plant can be expressed generally by the following formula:

$$Q = U \cdot A \cdot \Delta tm \cdots [\text{kcal/h}] \quad \cdots\cdots\cdots\cdots (6)$$

where Q: heat exchange quantity (Kcal/h)
U: coefficient of overall heat transfer (Kcal/m^2·h·°C)
A: heating area (m^2)
Δtm: log-mean temperature difference (°C)

Coefficient of overall heat transfer U is determined by that of Uc in a clean state and the fouling factor f.

$$\frac{1}{U} = \frac{1}{Uc} + f \cdots [m^2 \cdot h \cdot {}^\circ C/\text{kcal}] \quad \cdots\cdots\cdots\cdots (7)$$

Fig. 5.2-9 represents an influence exerted by the fouling factor on the coefficient of overall heat transfer. Scale is a substance small in coefficient of overall heat transfer, and hence the fouling factor gets large even with small quantities of scaling, thus decreasing the coefficient of overall heat transfer.

(b) Coefficients of contamination with various scale preventing methods

Fig. 5.2-10 represents a correlation between scaling and coefficient in the case of the decarbonation/pH controlling method, inhibitor adding method, seed adding method, sponge ball washing method and no-treatment. The coefficient of contamination is in proportion to scaling in every case. It is conceivable that the proportion constant in each preventing method varies because of a difference in specific gravity of the scale and the coefficient of overall heat transfer which results from the difference in composition and compactness of each contaminant.

5.2.2. Pretreatment Equipment

(1) Decarbonation equipment

(a) Principle of decarbonation

The operation for decarbonation is to eliminate carbonated matter dissolved in the seawater and is intended for prevention of scale formation in joint use with the pH controlling method and also for prevention of corrosion on the shell which results from dissolved gas. To diffuse the carbonated matter beforehand as carbon dioxide gas by adding acid is popular for this method.

(i) Chemical equilibrium of carbon dioxide

It is necessary for evaluation of the performance of decarbonation to be informed of the equilibrium state of carbonated matter in seawater. The equilibrium relation of carbon dioxide in solutions inclusive of seawater is generally as follows:

$$CO_2 + H_2O \rightleftarrows H_2CO_3 \rightleftarrows H^+ + HCO_3^- \rightleftarrows 2H^+ + CO_3^{2-} \cdots\cdots\cdots\cdots \quad (8)$$

In the above formula each concentration of carbon dioxide ($C_{H_2CO_3}$), bicarbonate ion ($C_{HCO_3^-}$), carbonate ion ($C_{CO_3^{2-}}$) and total carbon dioxide ($C_{totalco_2}$) can be expressed in the following formulas (9) - (12) from solubility (α), partial pressure of carbon dioxide gas (P_{CO_2}), activity of water (a_{H_2O}), dissociation equilibrium constant (K), etc.

$$C_{CO_2} = \alpha \times P_{CO_2} \quad \cdots\cdots\cdots\cdots \quad (9)$$

$$C_{HCO_3^-} = \frac{K_1' \times P_{CO_2} \times a_0 \times a_{H_2O}}{a_{H^+}} \quad \cdots\cdots\cdots\cdots \quad (10)$$

$$C_{CO_3^{2-}} = K_1' \times K_2' \times P_{CO_2} \times a_0 \times a_{H_2O} \frac{1}{(a_{H^+})^2} \quad \cdots\cdots \quad (11)$$

$$C_{tatalco_2} = P_{CO_2} \left(\alpha_s + \frac{K_1' \times \alpha_0 \times a_{H_2O}}{a_{H^+}} \left(1 + \frac{K_2'}{a_{H^+}} \right) \right. \cdots\cdots \quad (12)$$

In the above equilibrium relation, it is necessary to know the equilibrium concentration of total carbon dioxide under the influence of each factor of chlorine ion concentration, temperature, pH and partial pressure of carbon dioxide gas. Then, the equilibrium concentration of total carbon dioxide will be measured by means of vapor-liquid equilibrium equipment through changing each factor given above, or will be obtained through the maintenance method. One example is given in Fig. 5.2-11. In the figure an equilibrium relation between total carbonic acid and pH at a temperature of 40°C and the chlorine ion concentration 19‰ is shown. From the figure it is known that pH is about 8.3 when seawater is heated, and then concentration of total carbon dioxide is 2.0×10^{-3} gmol/ℓ (CO_2 being converted in 80 ppm). This indicates that pH lowers by diffusing carbon dioxide by adding acid, and the concentration becomes 6.5×10^{-5} gmol/ℓ (CO_2: 0.3 ppm). Then in the decarbonation operation in practice, it is desirable at such a point that alkali neutralization can be omitted when the equilibrium concentration is smaller and pH higher. Now, there is a perfect neutralization method to add more acid than alkali in the seawater and a partial neutralization method to add less acid than alkali; a recovery of pH being observed in the process of decarbonation in the latter, and hence an acid adding method utilizing this phenomenon is employed in most cases.

(ii) Controlling system

For adjustment of pH by adding acid, it is necessary to be acquainted with the aforementioned equilibrium concentration and the changing property of pH. One example of results obtained through test is shown in Fig. 5.2-12. The figure illustrates pH working, immediately after addition, against the added quantity of acid and after equilibrium to the air respectively. It can thus be understood that pH after equilibrium indicates a large buffer index of hardness to control. For injection of acid, therefore, problems are the trace injection method, selection of mixing level and controlling system inclusive of the pH detecting position.

Here, a pH controlling system set up with a 100 m^3/day desalination plant is shown in Fig. 5.2-13. The controlling system is shown, both acid and alkali. In the acid injection system a constant ratio of sulfuric acid is injected into a quantity of feed seawater and put in fine adjustment with pH at the decarbonation tower inlet later. What is problematical in the system is that care should be taken of the material and construction of the injection nozzle and mixing of acid must be quickened in consideration of using concentrated sulfuric acid. From paying due attention to these points, a controlling characteristic capable of coping with a change in the rate of flow can be secured.

On the other hand, control of alkali injection is not so necessary because of the normal decarbonation operation employing the pH recovery system. However, it will be necessary, particularly in an emergency, where decarbonation equipment deteriorates in performance or brine pH is to be improved. A control stronger than acid injection is then required for it. The method comprises a feedback control whereby the rate of alkali injection is adjusted through pH detection. The pH detecting position is intended for supplied seawater, however, pH of the circulating brine can be adjusted likewise with a given alkali solution through obtaining response characteristics of the pretreatment system.

(b) Characteristics of various decarbonation equipment

Decarbonation is a diffusion of carbon dioxide gas accompanied by reaction from the seawater and can be treated in the same way as general degassing equipment. For the level to treat, total carbon dioxide concentration in the seawater, which is about 100 ppm, will be decreased to 5 ppm or so, and hence a relatively simple operation is desirable.
The equipment subjected to test and examination comes in 5 kinds, such as packed tower, spray tower, airlift system, air bubbling system and multistage cascade fall system.
The construction of each equipment is given schematically in Fig. 5.2-14. Features and characteristics are described as follows:

(i) Packed tower system

This system is intended for decarbonation for which packing is placed in the tower and the seawater is kept in contact with the air in the packed bed. The gas diffusing rate of decarbonation is expressed in the product of the material transfer coefficient, vapor-liquid contact area and concentration difference, for which quantity of the seawater, quantity of the air, shape of the packing, as an equipment factors, and temperature, quantity of acid added, as operational factors are influential. A result obtained through observing the treating concentration against the quantity of seawater out of those factors is shown in Fig. 5.2-15. In the said figure, pH after addition of acid is low, indicating that the treating concentration lowers as the total carbon dioxide equilibrium concentration gets smaller. And the quantity of air is enough when more than 10 times the quantity of seawater, and feeding it more than that increases pressure loss, thus needing more blower power. On the other hand, performance of the packed tower is evaluated at the value of $(H.T.U.)_{OL}$ to come in the range 0.4 - 0.6 m. In view of the results above, an effect satisfying the treating level fully is obtainable through this system.

(ii) Spray tower system

The spray tower system is that of dropping the seawater from a nozzle by way of a liquid distributing plate in the empty tower with packing removed from the packed tower mentioned above. The vapor-liquid contact area and residence time are kept smaller than in the case of the packed tower, and efficiency deteriorates accordingly. According to the test result shown in Fig. 5.2-16, residual total carbon dioxide concentration comes down to 30 ppm or so with the quantity of air not influencing not. These results are conceivable for an open pond system simple in construction.

(iii) Airlift system

This system is intended for decarbonation by means of an airlift pump which consists of erecting a pumping-up tube in the seawater tank, blasting air from its bottom to produce a pressure difference inside and outside the tube and thus allowing the liquid level to ascend.

In the test the pumping-up characteristic and gas diffusion characteristic were examined. The pumping-up characteristic depends on the quantity of air and is also correlated with the shape of the pumping-up tube. As for the diffusion characteristic, a value $(H.T.U.)_{OL}$ several times as large as that of the packed tower is indicated in the well water and the service water, and further, a rather small value is indicated, as shown in Fig. 5.2-17, for the capacity coefficient K_{OLa}, thus suggesting an inferior decarbonation performance in comparison with the packed tower.

(iv) Air bubbling system

Against the packed tower system, an air bubbling system has been contrived to allow air to bubble in a channel as one of the open pond systems suitable for large capacity treatment. As the system is to effect vapor-liquid contact for diffusion while the seawater flows in the channel, pumping power is not necessary, but pressure of the blower to feed the air is largely influential on the liquid depth.

The result of these tests is shown in Fig. 5.2-18. The figure indicates the residual total carbon dioxide concentration available when changing the quantity of air against the length of channel. As will be apparent from the figure, decarbonation efficiency is improved as the quantity of air increases, however, a very long channel will have to be arranged for treatment of as much as 15 ppm or so.

(v) Multistage cascade fall system

The open pond system including the air bubbling system is cheap in construction cost but inferior in decarbonation efficiency as compared with the tower system.

In view of the above, decarbonation equipment of a multistage cascade fall system has newly been contrived. The system is to improve the decarbonation effect. It consists of putting water tanks one upon another in multistages, allowing seawater to overflow from the upper water tank and thus flow downwards to involve a volume of air cells when rushing into the lower water tank, thereby obtaining better vapor-liquid contact.

A test result is given in Fig. 5.2-19. As the result, the residual total carbon dioxide concentration for each step (concentration of each stage against the inlet seawater) is indicated linearly in a single logarithmic graph. Influential factors depend on kinetic energy. Performance in decarbonation at each falling zone can be evaluated by applying the perfect mixing tank row model.

The system is superior in performance in comparison with other systems in allowing the seawater to flow at a rate of 100 times or so, and is hence suitable for treatment in big quantities. But, the system is an open pond one and therefore open to influence by weather conditions. Because of splashing of seawater, it may be necessary to prepare a cover.

(vi) Conclusion

With reference to the five kinds of decarbonation systems, a description has been given above for features and performances; it can be said that the lower the better for the decarbonation level. However, residual total carbon dioxide concentration is permissible upto as much as 15 ppm in connection with loads on the following deaerator. Satisfactory systems, both economically and in performance, are the packed tower, air bubbling and multistage cascade fall systems. A comparison among the three systems is shown in Table 5.2-4. As will be apparent from the table, the packed tower system is disadvantageous in the operating cost, and the air bubbling system has problems in how to blast the air. On the other hand, the multistage cascade fall system is effective enough to have its performance and practicability proven at a plant of large capacity.

(2) Deaerator

(a) Principle and method of deaeration

Deaerating is an operation to eliminate oxygen dissolved in seawater and intended to prevent corrosion arising from oxidation of the vessel and heating tubes of the plant. For the level of elimination, it is required to adjust 7 - 9 ppm of oxygen dissolved in the seawater to less than 10 ppb. The means to eliminate the dissolved oxygen to such low concentration is classified into heating deaerating to decrease solubility and vacuum deaerating to lower partial oxygen pressure of the atmosphere. The vacuum system can be utilized for desalination equipment operating under the evaporation process, therefore the vacuum deserating method is used in usual cases. As a method to accelerate deaeration under vacuum, it is effective to increase the surface area by stirring and stripping vapor or atomizing by flash evaporation. For deaerating, possible systems are an external deaerating system arranged on the outside of the body as in the case of decarbonation equipment, and an internal dearating system enclosed internally. As equipment to work in these systems, the external method uses a packed tower and the internal method includes a final stage method to be carried out in the evaporation chamber of the final stage, and a heat rejection stage method to be carried out in the evaporation chamber on heat rejection stages of more than one.

(Fig. 5.2-20) The following is to describe these methods, features and characteristics:

(b) Characteristics of various deaerators

(i) Packed tower system

For this system the tower is packed with ruffling internally as in the case of a decarbonation packed tower, the seawater is allowed to go down from the top with the tower decompressed or vacuumized internally, and thus deaerating is carried out simultaneously by blast stripping vapor from the bottom. The Diffusion rate of the dissolved oxygen under vacuum is expressed by the product of the capacity coefficient and concentration difference, which is the driving force. For the concentration difference, it is recommended that the vacuum be kept high, as oxygen dissolved in the seawater is subject to Henry's Law.

A test result of the concentration change in the packed tower is given in Fig. 5.2-21 with gas partial pressure in the tower working as a parameter. As will be apparent from the figure, dissolved oxygen decreases in proportion to gas partial pressure, and a deaerating effect by stripping vapor can also be observed.

Since vapor and liquid come in differential contact in counterflow as in the case of a decarbonation tower, performance of the packed tower can be evaluated in $(H.T.U.)_{OL}$ value by giving an equilibrium concentration from gas partial pressure in the tower. The value $(H.T.U.)_{OL}$ indicates an inclination to increase in accordance with the quantity of seawater, which is given in the range of 0.5 - 0.6 m at 40 $t/m^2 \cdot h$ in the rate of flow.

(ii) Final stage system

This system is intended for deaerating the supplied seawater directly in the evaporation chamber through utilizing its final stage. As this system leaves only a limited space in the chamber, an efficient method will be required. To restrain pressure loss of the vapor, layout and construction of the tray to obtain an effective vapor-liquid contact in crossflow must be taken into consideration and further, an injection nozzle to accelerate atomization through setting the supplied seawater in the flash evaporation zone will have to be arranged.

A test result is shown in Fig. 5.2-22. The figure illustrates a concentration distribution obtainable from operation in the flash evaporation zone by means of two kinds of nozzles. As will be apparent from the figure, the porous nozzle is superior in deaeration efficiency to the single nozzle, and large matter migration can be expected thereby in accordance with atomization particularly at injection through the nozzle. For the tray construction it will be good to work with one to keep pressure loss of the brine vapor below 20 mmAq, and it is recommended that a tongued tray with good efficiency for liquid diffusion and contact be set up slanted in the direction of vapor flow.

(iii) Heat rejection stage system

This system allows the supplied seawater into the evaporation chambers of more than one heat rejection stage and deaerating it in the process to evaporate together with the brine. Therefore, the number of times of evaporation and the evaporating status at interstage orifice sections will influence deaerating.

A result obtained through a characteristic test under this system is shown in Fig. 5.2-23 with reference to a correlation between the number of times for flash evaporation and the deaerating level. As will be apparent from the figure, flash temperature difference and brine level height influence deaeration correlatively with the rate of recyclic flow, and it decreases in exponential against the number of times of flash evaporation. From the result it is affirmative that where the supplied seawater system can be changed, this system is available also for effective deaeration simply by changing the construction of the evaporation chamber slightly. However, materials of the heating tube for the evaporation chamber will have to be taken into consideration.

(iv) Conclusion

For deaeration methods, a description has been given above on features and performances of the three systems. Table 5.2-5 shows a result obtained through comparison of these outlines. From the table it can be understood that the packed tower system is an external type for arrangement with freedom in design, and the tray liquid load can be minimized to keep the dissolved oxygen below 10 ppb according to the said system. On the other hand, the final stage system and heat rejection stage system are enclosed in the evaporator shell, and the tray liquid load comes up over 100 $t/m^2 \cdot h$ in accordance as the shell is made smaller, thus giving a deaerating performance at 20 ppb in the case of the final stage system and 15 ppb in the case of the heat rejection stage system. Lower dissolved oxygen is preferable for corrosion prevention. However, dissolved oxygen in the circulating brine of less than 10 ppb is satisfactory for practicability.

Therefore dissolved oxygen in supplied seawater is permissble as much as 20 ppb. The three systems involve no problems accordingly, and the final stage system is advantageous in so far as equipment cost is concerned. Furthermore, the final stage deaerating system has been integrated in a 100,000 m^3/day module plant and proven satisfactory in performance so fa.

5.3. Plant Design

5.3.1. Brine Flow

In the flash evaporator the recirculating brine comes out of the heater to flow into the flash chamber in a superheated condition and passes from the flash chamber, as repeating evaporation, which is kept low in pressure by interstage orifices sequentially arranged from high temperature to low. The interstage orifices can be arranged in various types. However, from its governing fluidized condition in the flash chamber, the characteristics are important. Then the time the brine remains in the flash chamber is limited, and hence there arises a non-equilibrium temperature difference. Having an influence upon the performance of the flash evaporator, the non-equilibrium temperature difference is one of the most important characteristics. This paragraph takes up the brine flow in the flash chamber for description, particularly in emphasizing the non-equilibrium temperature difference and thus elucidates a correlation with plant engineering.

(1) Interstage brine flow

As shown in Fig. 5.3-1, a flash evaporation chamber provided with a rectangular orifice, ho in opening height, is used. The brine passing the orifice 1 - 2 flows with the sum of interstage pressure difference $\Delta Pv = P_1 - P_2$ and level difference $\Delta h = h_1 - h_2$ as a driving force. The mass flow W per unit width of the flash chamber with the flow coefficient of the orifice taken at C is then expressed generally as follows according to symbols given in Fig. 5.3-1:

$$W = Cp\sqrt{\frac{2g(\Delta Pv/p + \Delta h)}{1-(ho/h_1)^2}} \qquad (1)$$

The flow coefficient C is 0.6 - 0.65 or so in normal single phase flow. However, it becomes smaller generally than in the case of single phase flow, as generation and growth of bubbles are involved in passing the orifice in the flash chamber. A value at 0.48 on a three-stage testing device has been reported by the AMF Company. However, a sharp change in conditions of temperature, rate of flow, etc. may also involve a change in the aspect of evaporation and flowing condition to bring about a change in flow coefficient, thus preventing estimates for every condition under the present circumstances. Therefore, the orifice area is adjusted through experience at every stage in the trial run of the equipment. The shape of the orifice is variable as shown in Fig. 5.3-2 according to the rate of flow of brine, temperature, pressure and interstage differential pressure. a, b, c and d apply where the interstage pressure difference is small but the rate of flow is large. They are shapes seen in the low-temperature case for multistage flash evaporation. g - ℓ then apply where the interstage temperature difference is large but an active flash evaporation occurs, representing shapes which are seen in the high-temperature case. Fig. 5.3-3 illustrates the arrangement of an evaporation accelerator in the flash chamber, thereby minimizing non-equilibrium temperature difference. The flow resistance increases generally in such a case. There is a report that the flow coefficient comes down by 10% or so by fitting a baffle plate. The interstage pressure difference is small at low-temperature stages, therefore an interstage throttle mechanism to minimize pressure loss and also to accelerate evaporation is desirable, particularly for high speed current equipment. However, the two above are contradictory in general, and hence there may be a case where the flash chamber is inclined or a stage difference is given so as to accelerate the flow of interstage brine.

(2) Non-equilibrium temperature difference

The brine flowing into the flash chamber by way of the orifice gets in superheated condition by as much as the pressure drops to a boiling evaporation and then loses the sensible heat to lower temperature. And finally the temperature is to drop as low as saturation temperature Ts coping with the flash chamber pressure. Actually, however, the brine flows out into the next stage as unsaturated. The value then obtainable through deducting boiling point rise BPE by dissolved component from the difference between outflow temperature To and saturation temperature Ts is called the non-equilibrium temperature difference NETD, which is expressed by the following formula:

$$NETD=To-(Ts+BPE) \qquad (2)$$

Then,

$$\beta=\frac{\triangle T}{\triangle Ts}=\frac{\triangle T}{\triangle T+NETD} \qquad (3)$$

is called equilibrium arrival extent, where ΔT = Ti - To, Ts = Ti - (Ts + BPE), Ti and To being brine temperatures at the flash chamber inlet and outlet respectively. For the causes from which such non-equilibrium temperature differences arise:

i) residence time of the flash brine in the evaporation chamber is limited,

ii) for the growth of bubbles in the liquid some degree of superheat is necessary to overwhelm the bubble surface tension, and

iii) Boiling of the liquid under the level is suppressed due to a static pressure rise by liquid depth

will be pointed out. Since generated vapor temperature drops by as much as a non-equilibrium temperature difference is present, heating driving temperature difference will also lower that much. Heat recovery decreases and also manufactured water quantity lowers in consequence. To keep the quantity of manufactured water constant, it is necessary to increase the heating area or heat vapor quantity. The heating tube cost is 30 - 50% of the equipment cost, and the heat vapor cost is more than half of the operating cost. The heating driving temperature difference is barely 5°C or so, and hence the non-equilibrium temperature difference may exert a great influence upon desalination cost. A calculation of the influence according to a simple model is as given in Fig. 5.3-4: where there is a non-equilibrium temperature difference at 1°C, for example, the desalination cost will advance by 13% at a fuel oil cost of 30 yen/t.

It is necessary to keep the non-equilibrium temperature difference as low as possible, as it exerts a vital influence upon performance of the equipment. It also represents efficiency of the evaporation chamber with many results obtained through measurement of it, and several experience formulas have been proposed as referred to later. However, for engineering of equipment with far better performance, it is necessary to grasp phenomena in the flash chamber quantitatively. Therefore, a description will be given on flash phenomena in the chamber and correlation of the non-equilibrium temperature difference with the test results obtained by Sugata and Toyama emphatically taken up, and further test results already reported and experience formulas on non-equilibrium temperature difference will be referred to as follows:

(a) Aspect of flash evaporation and non-equilibrium temperature difference

Coogan and his cooperators measured current speed distribution and temperature distribution in a flash chamber 3.45 m long and 30 cm wide and divided the flash flow into three areas. There is the inlet area in which a jet coming out of the orifice expands upwards to reach the liquid surface and then the central area in which the current speed distribution develops to allow the maximum current speed near the lower side to come out on the liquid surface and then come near to an even current speed. Finally the flow transfers to the outlet area in which the flow is accelerated to the orifice on the next stage. However, the report is limited to one condition.

According to an observation carried out by Sugata and Toyama on single-stage flash evaporation testing equipment 3 m long and 10 cm wide, with the front glazed, aspects of flash evaporation working where a simple orifice is arranged like that of Fig. 5.3-2-a have two types qualitatively, as shown in Fig. 5.3-5 and Fig. 5.3-6. In Fig. 5.3-5 a jet from the orifice loses temperature as it involves the liquid on the upper bed which is low in temperature and penetrates from the chamber bottom to the next stage. A counterflow zone with a low temperature covering the overall stage arises on the upper bed of the jet, and boiling takes place as bubbling on the surface. In Fig. 5.3-6 air cells are produced immediately after inflow, which grow to form a jet in an intense two-phase flow. The flash liquid disperses in drops and thus evaporation proceeds quickly. However, boiling scarcely arises except in this area and a part of the surface. The boiling is suppressed by static pressure rise due to the liquid depth at the bottom as in the above case. Boiling like Fig. 5.3-5 is called bubbling boiling and that like Fig. 5.3-6 splash boiling. Variance of such two types of boiling is subject to temperature, rate of liquid flow, liquid depth, flash temperature drop, stage length, etc. In liquid depth 40 - 60 cm both types are observed at 40°C, but boiling changes from bubbling boiling to splash as the liquid depth decreases and also the rate of flow is increased. A similar inclination will be given by increasing the flash temperature difference. All is splash boiling at more than 50°C.

Next, a result obtained through measuring the non-equilibrium temperature difference at the tests is given in Fig. 5.3-7. Bubbling boiling occurs on the surface of the counterflow zone with low temperature, giving an exceedingly large non-equilibrium temperature difference. A transition from bubbling boiling to splash boiling is brought about by a change of liquid depth of 10 cm or so, thereby changing the non-equilibrium temperature difference intermittently. Where W = 1,200 ton/m·hr, for example, the non-equilibrium temperature difference at bubbling boiling becomes 1.5 - 2.0 times as large as at splash boiling. This is regarded as excessively large as compared with the influence of liquid depth indicated by the experience formula already reported. Then, change of the non-equilibrium temperature difference against ΔT indicates quite a contrary inclination to that known so far at 40°C.

Meanwhile, in every type of boiling such a simple orifice is not effective for flash evaporation on the evaporation chamber bottom particularly at low temperatures, and the flow at the bottom works as a factor to increase the non-equilibrium temperature difference. Then, from the conception that performance will be improved by taking this portion away, the evaporation chamber bottom can be raised to a trapezoid. The non-equilibrium temperature difference then working is represented by the solid mark in Fig. 5.3-7. This indicates that the non-equilibrium temperature difference decreases sharply at 40°C, ensuring the effectiveness of the raised bottom at low temperatures. The inflow liquid is transferred upwards by the raised bottom and the hot liquid is delivered onto the surface boil, therefore suppression action by the liquid depth is kept small and the non-equilibrium temperature difference lowers. Miyatake and others feel it necessary for a decrease in non-equilibrium temperature difference to prevent the hot liquid from penetrating to the chamber bottom through arranging a baffle plate downstream of the orifice, according to their test on small testing equipment.

Then, a change of equilibrium arrival extent in the direction of flow is shown with reference to the aforementioned test conditions is Fig. 5.3-8. In splash boiling it is understood that evaporation is nearly completed at 1/3 - 1/2 of the former half of the stage. But this is not necessarily to indicate that the chamber length can be so short. To shorten the chamber in length is to allow a counterflow zone to develop, thus suppressing evaporation.

(b) Estimation of non-equilibrium temperature difference

Values obtained through measuring non-equilibrium temperature difference have been reported often by OSW and some other sources, and the particulars are given in Table 5.3-1. Then, literature 1), 9), 10) and 11) refer to measurement results on three-stage testing equipment by OSW, which are also arranged in literature 3). The following experience formulas are proposed against blank stages:

$$NETD = 2.19\exp(2.76H + 0.032W\times 10^{-5} - 0.0641\,Tv) \qquad (4)$$

test range being

$$L = 4.27\text{m},\ H = 0.3 \sim 0.56\text{m},\ Tv = 38 \sim 63℃$$

$$W = 7.4 \times 10^{2} \sim 14.9 \times 10^{2}\ ton/m\cdot\text{hr}$$

$$NETD = \frac{0.156H^{0.86}.Vg^{0.71}.(W\times 10^{-2})^{0.455}}{\triangle T^{0.5}} \qquad (5)$$

test range being

$$L = 3.45\text{m},\ H = 0.25 \sim 0.61\text{m},\ Tv = 24 \sim 60℃$$

$$\triangle T = 1.1 \sim 3.9\ ℃,\ W = 3.7 \times 10^{2} \sim 14.9 \times 10^{2}\ \text{ton}/\text{m}\cdot\text{hr}$$

$$NETD = \frac{0.857H^{0.344}.Vg^{0.284}.(W\times 10^{-2})^{0.162}}{\triangle P^{0.348}} \qquad (6)$$

$$L = 4.8\,m$$

$$NETD = \triangle T\ \left(\frac{2.88}{\triangle P^{.22}.Vg^{0.05} - 1}\right) \qquad (7)$$

$$L = 4.8\ m$$

$$NETD = 5.878^{-0.0128Tv} \cdot 1.522^{6.56H} \cdot 1.176^{0.108W} \qquad (8)$$

$$L = 3.05\ \text{m}$$

where L: length of chamber, W: rate of brine flow, Vg: vapor specific volume (m^3/kg), ΔP: interstage pressure difference (mmHg). Formula (4) represents the result obtained by AMF Company on 2-stage equipment; formula (5) represents the result obtained by the said company on 3-stage testing equipment. Formulas (6) and (7) represent results obtained by BLH Company; formula (8) represents the result obtained by Oak Ridge National Laboratories through modifying formula (4). These experience formulas are available for stages variant in length respectively, and comparisons resulted as shown in Fig. 5.3-9 and Fig. 5.3-10. Fig. 5.3-9 also shows the result obtained by Sugata and others through measurement. Formula (8) will give maximum value, and formula (4) minimum. Formulas (5), (6) and (7) are intermediate, showing values nearly in the same degree. As a result of comparison with the measured values already reported, formulas (5) and (6) correspond to each other in the limit of error to nearly the same degree. Formula (8) was obtained at the longest chamber length, and formula (6) the shortest; formula (5) is intermediate. Therefore, the difference among (5), (6) and (8) may represent a difference in the length of chamber. However, while the chamber lengths of formulas (5) and (6) are considerably different, calculation values of non-equilibrium temperature difference are not so different. Now, it is conceivable that the flash aspect will change sharply between 3.05 m and 3.45 m in chamber length. And the result of formula (8) shows a value approximate to the measured value at the time of splash boiling like Fig. 5.3-6. These chamber lengths are nearly the same. Formulas (5) and (6) nearly coincide with the values at the raised bottom. From these inclinations, it is understood that non-equilibrium temperature difference is not so reduced at the blank stage more than certain extent even by making the chamber length long, but can be decreased to the same extent as a blank stage with long chamber length, even with a short chamber length, by arranging an evaporation accelerator with the bottom raised. In the test results on an evaporation accelerator which is referred to in OSW's report, the difference could scarcely be decreased more favorably than in the case of a blank stage with long chamber length. From the above results it is understood that formula (5) or (6) will be good for non-equilibrium temperature differences.

(3) Flash chamber engineering

The length of flash chamber must be long enough thermodynamically to complete evaporation as much as possible. However, it will be kept at a minimum requirement for saving the evaporator shell cost. Then in the case of long-tube type equipment, the length necessary for incorporating the condenser heating area is subject to restriction. As described in the foregoing paragraph, a decrease in non-equilibrium temperature difference cannot be expected so favorably from making the chamber length long more than certain extent. Moreover there are few data obtainable for chamber length and non-equilibrium temperature difference, and hence the chamber length is determined according to the requirements for length to incorporate the necessary heating tube bundle in the majority of cases. The length normally is 3 m or so for the equipment. Particularly in high current speed equipment, the non-equilibrium temperature difference runs exceedingly high in the case of the blank stage, therefore it is necessary to contrive something in the shape of orifice or to prepare an evaporation accelerator in the flash chamber. In the test module in the large project, a satisfactory result has been obtained at low temperature operation through slanting the flash chamber entirely with its length of 2.8 m arranged as a raised bottom at 1,700 ton/m·hr in the circulating brine flow.

5.3.2. Heat Exchange

Since seawater is of high corrosiveness in comparison with fresh water, a comparatively high-priced copper alloy tube or further expensive titanium tube is used for the heating surface of the heat exchanger, and thus the cost for the heating surface is a large proportion of the equipment cost. It is therefore essential to minimize the heating surface by improving the coefficient of heat transfer or to remove factors working to decrease the coefficient of heat transfer.

In the flash evaporator main sections, where heat exchange is performed, are the brine heater and heating tube bundle for the flash evaporation chamber. The vapor is condensed outside the tube, and the brine is concentrated to about two times the seawater which flows inside the tube to the heat exchange. It has been taken generally so far that the condensation heat transfer coefficient is by far larger than the forced convection heat transfer coefficient, and thus an effect cannot be so expected from increasing the condensation heat transfer coefficient. In the case of a condenser with a single tube or few tubes, the condensation heat transfer coefficient is by far large in comparison with the film heat transfer coefficient, and hence a better effect cannot be expected from increasing the condensation heat transfer coefficient. However, in the case of a tube bundle consisting of 3,000 - 4,000 heating tubes, the condensation heat transfer coefficient is reduced by the condensate flowing down outside the tubes, and the heat transfer coefficient inside the tube and the condensation heat transfer coefficient are nearly equivalent at normal current speed. The main means to improve the heat transfer coefficient in the tube is to increase the current speed. A method to insert a turbulence promotor can be tried alternatively. However, an increase of pressure loss is involved or flow corrosion advances actively. For the current speed, 2 m/s will be the maximum limit. An attempt to improve the condensation heat transfer coefficient is made separately so as to improve the overall coefficient of heat transfer and miniaturize the condenser. On the other hand, an attempt to use a cheap heating surface or to effect direct contact heat transfer, for which a heating surface is not necessary, is made too. However, it is very hard to ensure practicability with any of them. Drop condensation, which is useful for remarkable improvement of the condensation heat transfer coefficient, has been studied in various ways. However, no way has been found to keep drop condensation cheap and stable for a long period of time. Heat transfer accelerating tubes called Colgate Tubes and Double-fruited Tubes are fairly good prospects nevertheless. However, they are not yet ensured for durability and hence are not practicable at present.

The present situation as mentioned above, engineering requirements will have to be based on achievements obtainable through piling up past improvements steadily. These comprise preventing non-condensable gas from leaking, discharging non-condensable gas efficiently so as not to allow it to stay in the heating tube bundle, paying attention that the condensate does not flow down into the lower tubes so much, preventing scale and sludge from settling, and so forth. Therefore, engineering policies taken by each company through past experience are significant indeed. OSW of the United States has gone through the present situation on engineering of the desalination condenser, leaving valuable data in this connection.

(1) Construction

The heating tube bundle in the evaporator shell comprises cross tube type and long tube. In the case of cross tube there is no change particularly from the normal condenser and no substantial change admitted on the high temperature stage. Then in the low temperature stage, the specific volume of the vapor generated on the stage rises, the evaporator shell has been decompressed, and the air leaks in to work as non-condensable gas. Therefore care should be taken for flow on the vapor side in the heating tube bundle so as not to allow the non-condensable gas to stay in the tube bundle.

Where long-tube construction is employed, the tube plate can be only one on both ends for several stages to several ten stages. Therefore, tube plate of seawater resisting material which is thick and high-priced can be saved from use. In addition, pressure loss arising from enlargement or contraction of the cooling water channel can be minimized, too. On the other hand, however, it is necessary to seal the vapor at the interstage bulkhead of every stage. There are, therefore, methods to install O-rings between interstage bulkheads and heating tubes and methods to use pincushion sealing material as shown in Fig. 5.3-11 left. Besides, a method is available for liquid sealing which keeps the clearance between bulkhead and heating tube small and allows the condensate produced in the chamber on the high-pressure side to flow into the clearance as shown in the said figure, right.

(2) Heating tube and its configuration

The heating tube is normally zigzag (equilateral triangle) as shown in Fig. 5.3-12-(a) and (c). There may be a case where the tube is square, as shown in (b) and (d) of the figure, to minimize pressure loss on the vapor side. However, this is not for application normally on a desalination plant. A relation between tube diameter and pitch in the tube configuration is standardized, through long experience, as shown in Table 5.3-2.

An aggregation of heating tubes is called a tube bundle, for which a circular type is normal, as shown in Fig. 5.3-13. Condensation is chiefly at the condensing section in I - IV, and there is an air cooling section at the center of the tube bundle in the case of a fully circular condenser shown in Fig. 5.3-13 left, where gas mixed with vapor high in air concentration through condensation at the condensing section and air is further condensed. In the condensing section, as shown in the figure, the vapor flows downward or laterally: the downward stream is caused by a parallel flow of the condensate and the vapor, and the lateral stream by a crossflow thereof. In the air cooling section, on the other hand, the condensate flows downward and the vapor flows upward in counterflow. Both flow in vapor-liquid two-phase flow in the tube bundle. There may be a case where the heating tube bundle is rectangular as shown in Fig. 5.3-14. Where the scale is comparatively small, semicircular or semirectangular configuration may be as shown in Fig. 5.3-13 and Fig. 5.3-14 right.

The method to mount the heating tube on the plant is illustrated in Fig. 5.3-15. The heating tube is mounted normally with expansion tube as shown in Fig. 5.3-15-(a), however, it is welded like (d) according to the circumstances. Other methods are uncommon.

The specific volume of the vapor on the high temperature stage of the flash evaporator shell is small and pressure loss arises less from the vapor flowing in the tube bundle, therefore a steam lane is not particularly required in the heating tubes. However, for the low temperature stage or the heat rejection section particularly, a steam lane will have to be taken into consideration so as to allow the vapor to flow easily. Fig. 5.3-16 is proposed as a tube plate arrangement for the 35th stage of the MSF preheater in a conceptional design of a 757,000 m^3/day VTE/MSF plant. To prevent heat transfer resistance, in this case, from getting large due to the condensate dropping onto the heating tube surface to flow down in a thin film, the tube bundle is short vertically and long horizontally. Furthermore, to minimize pressure loss arising from the vapor flow, the heating tubes are spaced on the circumference of the tube bundle but close near to the center as the vapor becomes less in quantity. The distance between heating tubes thus functions as a steam lane.

The thermal stress working from the difference in thermal expansion between heating tube and evaporator shell is not so influential because the temperature range wherein the flash evaporator applies is comparatively small, and hence no measures will have to be particularly taken. The thermal stress can be partly absorbed by flexibility of the heating tube.

(3) Investigations on engineering processes

The American OSW entrusted the Oak Ridge National Research Institute with investigations on how to design a condenser for a large desalination plant. The investigation was intended practically for the engineering process through selecting such information as is reliable in various problems for engineering like heat transfer, pressure loss on the vapor side, tube length, tube diameter, discharge of non-condensable gas, dropping of condensate, etc., and a program was developed which could be used generally for engineering of desalination plants of more than 190,000 m^3/day in capacity.

(a) The following literature is recommended as reliable for engineering of the condenser:

i)	Forced convection film heat transfer coefficient	1
ii)	Condensation heat transfer coefficient	2
iii)	Pressure loss on vapor side	3 , 4 , 5 , 6
iv)	Influence of non-condensable gas	7 , 8
v)	Discharging process of non-condensable gas	9

(b) Investigations on actual engineering process

Investigations were made with 5 American leading manufacturers for an engineering process actually applied to the condenser, and the features were taken into the arrangement. The engineering process on large condensers is not necessarily elucidated logically to a full extent, and that obtained through know-how acquired through years of achievement and experience is overwhelming. The investigations are therefore significant in representing valuable data and partly arranging the know-how. The following are main features:

Non-condensable gas is collected into the air cooling section at the center of the condenser bottom by the tube configuration and vapor channel and then discharged by the vapor ejector.

Maximum current speed of the vapor is to be taken at 90 - 150 m/s, and care should be taken so as to allow the condensate and the vapor to flow in parallel, thereby minimizing pressure loss on the vapor side. The vapor current speed in the tube bundle must be 12 - 15 m/s so as not to allow non-condensable gas to stay. The air cooling section is nearly 2% of the whole tube bundle, and about 5% vapor is discharged at maximum through the discharge system of non-condensable gas.

(4) Design calculation formulas

The following formulas are used for design calculation:

a. Condensation heat transfer coefficient

$$h_m = 0.725 \left(\frac{k_f^3 \, \rho_f^2 \, H_{FG} \, g_C}{\mu_f \, D_0 \, (t_{sat} - t_w) n} \right)^{0.25} \cdot C \cdot F \qquad (1)$$

where, h_m: mean condensation heat transfer coefficient (Kcal/m^2·hr·°C)

k_f: thermal conductivity of condensate (Kcal/m·hr·°C)

ρ_f: specific gravity of condensate (kg/m^3)

H_{FG}: evaporation latent heat of condensate (Kcal/kg)

f: viscosity of condensate (kg/m·hr)

Do: OD of heating tube (m)

t_{sat}: saturation temperature of vapor (°C)

t_w: outside temperature of tube wall (°C)
$t_w = t_{sat} - 0.5(t_{sat} - t_w)$

n: number of heating tubes, vertically

F: coefficient representing influence of non-condensable gas

g_c: gravitational conversion factor (m/hr^2)

The coefficient C, representing the influence of tube number, is in the following functional relation with the number of tubes n in vertical direction:

$$C = 1.310 \qquad \text{where} \quad n = 1 \qquad (2.a)$$

$$= 1.2379 + 3.5361 \times 10^{-2}n - 1.5703 \times 10^{-3}n^2$$

$$\text{where} \quad 2 \leq n \leq 16 \qquad (2.b)$$

$$= 1.4017 \qquad \text{where} \quad n > 16 \qquad (2.c)$$

Where non-condensable gas is less than 4% in weight, F is in the following functional relation with the quantity of non-condensable gas:

$$F = 1.0 - 3.4313 \times 10 x + 1.2268 \times 10^{3} x^{2} - 1.4923 \times 10^{4} x^{3} \qquad (3)$$

where is weight percentage of non-condensable gas.

b. Film heat transfer coefficient

The film heat transfer coefficient in the tube can be calculated by the following formula in the range Re = 10,000 - 120,000, Pr = 0.7 - 120, $\ell/d \geqq 60$:

$$hD_i/k = 0.027 Re^{0.8} Pr^{0.333} (\mu_{aue}/\mu_w)^{0.14} \qquad (4)$$

where h: forced convection film heat transfer coefficient (Kcal/m^2·hr·°C)

D_i: ID of heating tube (m)

k: thermal conductivity of brine (Kcal/m^2.hr.°C)

Re: Reynolds number of brine

Pr: Prandtl number of brine

μ_{aus}: brine viscosity at mean liquid temperature (kg/m·hr)

μ_w: brine viscosity at wall temperature (kg/m.hr)

For heat transfer coefficients on the vapor side and brine side, wall temperature twi on the brine side and that two on the vapor side will be obtained for repeated calculation.

c. Where non-condensable gas is much

In case non-condensable gas exceeds 4 wt %, the following formula applies:

$$1/h_{cg}=Q_g/(Q_T h_g)+1/h_m \qquad (5)$$

where h_{cg}: condensation coefficient where non-condensable gas is present

h_g: gas film coefficient

Q_g: sensible heat eliminated from gas

Q_T: total heat eliminated from gas and vapor

h_m: condensation coefficient where non-condensable gas is not present

Then, the gas film coefficient is calculated by the following formula:

$$h_g D/k= 0.33 Re^{0.6} Pr^{0.33} \qquad (6)$$

where h_g: forced convection heat transfer coefficient (Kcal/m^2·hr·°C)

Do: OD (m)

k: thermal conductivity of gas (Kcal/m^2·hr·°C)

Re: Reynolds number

Pr: Prandtl number

d. Pressure loss

A pressure loss of vapor-liquid two-phase flow arising where the vapor flows horizontally in the heating tube bundle and the condensate vertically downwards is obtained through the following formulas:

Where $x \leq 0.15$

$$(\Delta P_T/\Delta P_G)=1.0 \qquad (7\cdot a)$$

Where $0.15 < x \leq 2.0$

$$(\triangle P_{TP}/\triangle P_G) = \exp[-0.93213 - 0.74257 \ln x - 0.13174(\ln x)^2] \quad (7 \cdot b)$$

Where $x > 2.0$

$$(\triangle P_{TP}/\triangle P_G) = \exp[-0.85714 - 0.94972 \ln x - 0.015453(\ln x)^2] \quad (7 \cdot c)$$

where

ΔP_{TP}: pressure loss of two-phase flow

ΔP_G: pressure loss in the case of vapor-liquid flow

LVF: proportion occupied by liquid

A pressure loss arising where the vapor flows downwards in the heating tube bundle in parallel with the condensate is obtained through the following formulas:

Where $x_D \leq 0.001$

$$(\triangle P_{TP}/\triangle P_G) = 1.0 \quad (8 \cdot a)$$

Where $0.001 < x_D \leq 0.02$

$$(\triangle P_{TP}/\triangle P_G) = \exp[-4.0317 + 0.54538 \ln x_D + 0.39108(\ln x_D)^2 + 1.0329(\ln x_D)^3] \quad (8 \cdot b)$$

Where $x > 0.002$

$$(\triangle P_{TP}/\triangle P_G) = 0.0225\, x_D^{-1.02} \quad (8 \cdot c)$$

where

$$x_D = \{LVF/(\rho_g/\rho_L)\}/Re^{0.5}$$

e. Overall coefficient of heat transfer

The overall coefficient of heat transfer is obtained through the following formula:

$$U = \sum_{i=1}^{n} U_i A_i \Big/ \sum_{i=1}^{n} A_i \qquad (9)$$

where U: mean overall coefficient of heat transfer (Kcal/m^2·hr·°C)

U_i: overall coefficient of heat transfer at i row (Kcal/m^2·hr·°C)

A_i: heating area at i row (m^2)

n: total number of rows

f. Effective logarithmic mean temperature difference

The effective logarithmic mean temperature difference is obtained through the following formula:

$$\Delta t_m = \sum_{i=1}^{n} Q_i \Big/ U \sum_{i=1}^{n} A_i \qquad (10)$$

where Δt_m: effective logarithmic mean temperature difference (°C)

Q_i: Quantity of heat transferred at i row (Kcal/hr)

i: number of tube rows

g. Others

For leakage of non-condensable gas from the shell wall, pressure loss by the brine flowing in the heating tube, pressure loss arising from enlargement and reduction of the passage at the brine flowing in and out of the heating tube, etc., it is convenient to utilize graphs and charts given in the standards for design of the steam condenser which are specified by HEI. Since there is no extra space left here to cite them accordingly, reference will be made to them where necessary. For manufacture and execution of work, particulars are the same as in the case of a general heat exchanger. Thermal conductivity of the heating tubes used for desalination can be obtained through Table 5.3-3. Contamination factors covering various cases are given in Table 5.3-4.

(5) Programming for design

For design of small type condensers the aforementioned charts by HEI can be used as they are with something in reserve. Where the condenser is a large type, the program given in the OSW's report will be ready for servicing.

5.3.3. Demisters

(1) General

In the case of a multistage flash evaporator, drops are dispersed in the process of steam coming out of the raw seawater, causing the so-called entrainment phenomenon. The entrainment is capable of mixing in the distilled water to deteriorate its purity, for which an eliminating device will have to be prepared. Such a device is called demister (or mist eliminator more accurately). The demister has various systems such as cyclone, packed tower, wire mesh, and baffles as collision plates. The system is selected for desalination equipment according to the following restrictions. First of all, the eliminating device must be built in the flash equipment. Therefore, the device needs to have no moving parts and, further, to be compact. Secondly, the device must work efficiently for eliminating mist which is generated by with flash evaporation and then keep pressure loss low which may deteriorate thermal efficiency. Wire mesh and collision plates will be used for systems coping well with the above requirements. However, wire mesh is preferable, in comparison of both the two, in areas for installation, pressure loss and eliminating efficiency, and hence is employed for much desalination equipment. Comparison in performance of the main eliminating devices is shown in Table 5.3-5.

The wire mesh to be set up in flash desalination equipment is a simple mist eliminating device, as shown in Fig. 5.3-17, for which slender wires (250 μm or so) are knitted and then piled up with waves 5 mm or so in amplitude and 10 mm or so in length shaped in them. Since it can be sized arbitrarily, a unit in the evaporator shell or absorption tower can be easily incorporated, and its gaps exceed 90% normally, which minimizes pressure loss. It has merits of being long in life and less in maintenance cost from having no moving parts. On the other hand, the mist load increases on the wire mesh and redispersion occur easily as the mesh-passing gas current increases in speed, for which care should be taken. In other words, the phenomenon is such that the mist, once collected, joins the gas flow again, and an occurence of the phenomenon may result in insignificance of having installed the demister. It is therefore indispensable to select operating conditions within the limits of keeping redispersion phenomenon from arising and thus to allow the wire mesh to fulfill its performance satisfactorily. The following is to describe its general performance, redispersion limits and collecting efficiency with reference to the wire mesh mist eliminator:

(2) General characteristics of wire mesh

The wire mesh in water-air group shows tendencies like Fig. 5.3-18.

a) Where the gas flows at low speed, mist collected on the mesh wire aggregates, comes vertically down onto the eliminator bottom and then is eliminated in drops. The illustration is as far as point A in the figure.

b) Where the gas flow is accelerated from point A, the mist collected in the eliminator is not eliminated in drops but stays on the bottom of eliminator. This is the so-called loading point. Where the gas flow is further accelerated from the point, the mist staying in the eliminator increases in quantity, and the proportion of pressure loss becomes large in consequence.

c) When the gas flow reaches point B, the quantity of mist collected in the eliminator and pressure loss sharply increase, the redispersion phenomenon then appears and the effect of the mist eliminator is lost.

(3) Collecting efficiency

For mist collecting mechanism in the mesh mist eliminator, collection by inertia against the wire is the normal operating condition. Therefore, mist collecting efficiency is represented by the value depending on mist drop size, gas flow speed, wire diameter, physical property of the gas, etc. For dependency on the mist drop size, a collecting efficiency over 90% is obtainable for mist drops more than about 10 μm in size through experimental observation. However, for mist drops less than that in size, the collecting efficiency decreases in accordance as the drop size decreases, thus deteriorating below 10% at 2 - 3 μm. For mist drops less than 1 μm, meanwhile, elimination by the mesh eliminator becomes difficult, and no more effect can be expected from increasing the thickness of the mesh layer.

Next, a description will refer to the relation between collecting efficiency of the wire mesh eliminator and thickness of the layer and also relation between phenomenon characteristic of the wire mesh and collecting efficiency. Assuming the collecting efficiency per unit thickness is $\Delta\eta_0$ and the quantity of mist passing the eliminator E, collecting efficiency on the overall layer will be introduced to the following from the movement of mist at micro-thickness:

$$dE/dL = -E\Delta\eta_0 \qquad (1)$$

$$\Delta\eta = (Ein - Eout)/Ein \qquad (2)$$

$$\Delta\eta = 1 - 1/e^{\Delta\eta_0 L} \qquad (3)$$

Actually, however, the collected mist comes down vertically onto the eliminator bottom in drops, therefore the collecting efficiency is different from that in the interior. In taking this into consideration, let collecting efficiency on the eliminator bottom be $\Delta\eta_S$,

$$1-\Delta\eta_T=(1-\Delta\eta_S)(1-\Delta\eta) \qquad (4)$$

then, collecting efficiency $\Delta\eta_T$ of the mesh eliminator is given by the following formula:

$$\Delta\eta_T=1-(1-\Delta\eta_S)/e^{\Delta\eta_0 L} \qquad (5)$$

According to a test result, $\Delta\eta_o=0.1172$ and $\Delta\eta_S=0.7648$ in the mesh eliminator 99% in gaps, 0.25 mm in wire diameter and 100 mm in thickness. Fig. 5.3-20 is that in which a relation between eliminator thickness L and collecting efficiency $\Delta\eta_T$ has been obtained through utilizing the above values. And Fig. 5.3-19 shows collecting efficiency in case collecting effect on the eliminator bottom is not taken into consideration also at $\Delta\eta_{in}$. As will be apparent from the result, the collecting effect on the eliminator bottom plays a fairly big and important role.

(4) Pressure loss

For pressure loss on the mesh eliminator, two cases are conceivable: one is in dry condition and the other in wet condition that mists have stuck on the mesh wire. The two are expressed by the following formulas:

$$\Delta P_d=C_{Dd}\ \rho_g/2g_c\ (V_g/\varepsilon)^2\ A_d/A_0 \qquad (6)$$

$$\Delta P_w=C_{Dw}\ \rho_g/2g_c\ (V_g/\varepsilon-w)^2\ A_w/A_0 \qquad (7)$$

where the subscripts d represents dry condition and w wet condition, C_{Dd} and C_{Dw} represent resistance coefficients in each case, A_d and A_w represent projected sections including mesh wire and sticking mist quantity, A_o represents an area in the direction right-angled to the flow of mesh eliminator. Then, w represents sticking mist quantity, ε gaps of mist eliminator and V_g gas velocity.

Pressure loss ΔP_d in dry condition has been studied so far, to which the following formula applies normally:

$$\triangle P_d = C_{Dd} \frac{2\rho_g}{\pi g_c} \frac{L}{D_f} \left(\frac{V_g}{\varepsilon}\right)^2 (1-\varepsilon) \qquad (8)$$

$$C_{Dd} \doteq 0.6 + \frac{4.7}{\sqrt{R_e}} + \frac{11}{R_e}$$

where D_f represents wire diameter and R_e Reynolds number for which wire is standardized.

On the other hand, pressure losses resulting from dry condition of the mesh eliminator and wet condition where mist sticks to it are inclined to indicate two straight lines in parallel, as shown in Fig. 5.3-21, on both logarithmic sheets. Assuming $C_{Dd} = C_{Dw}$ as there is no essential difference admitted between C_{Dd} and C_{Dw}, $A_w/A_d = 1.38$ from applying the above result to formulas (6) and (7). Now, pressure loss at the time of mist sticking can be calculated through formulas (6) and (7).

(5) Loading phenomenon and redispersion phenomenon

A loading point where collected mist does not drop steadily and a redispersion point where the redispersion phenomenon comes out can be observed from increasing the gas velocity passing the mesh eliminator. Both the two are caused by a resistance working according to the gas flow, and the resistance R is expressed generally by the following formula:

$$R = \tfrac{1}{2} C_D \rho_g V_g^2 A \qquad (9)$$

where A represents an area covered by the drops coming down vertically onto the eliminator bottom. Assuming the condition of eliminator including drops is the same at an arbitrary system, C_D and A are even in values irrespective of the system, therefore these can be compared with the test result already obtained for water-air system, etc. on the loading point. As the result, the following formula is introduced from the formula (9), and thus the loading point can be calculated from the data already known:

$$\rho_a V_{al}^2 = \rho_v V_{vl}^2 \qquad (10)$$

where the subscript l represents loading point, a data already known and v estimate unknown. For redispersion point, the following formula is obtainable through similar examination, thereby estimating redispersion point:

$$\rho_a V_{ar}^2 = \rho_v V_{vr} \qquad (11)$$

where the subscript r represents redispersion point.

5.3.4. Piping and Pump

(1) General system and characteristics of fluid

The piping and pump system for desalination plants under the multistage flash evaporation process can be classified principally as follows:

A description of the system will refer to the case of the pH control system; the system diagram is given in Fig. 5.3-22.

(a) Cooling seawater system

This system covers the heat rejection section cooling seawater system from the intake tank to the drain tank through the seawater feed pump and heat rejection section of the evaporator and ejector condenser cooling seawater system to drain the tank through the ejector condenser.

The fluid is seawater up to 40°C in normal temperature, which is saturated for dissolved oxygen quantity to high corrosiveness.

(b) Supplied seawater system

This is the supplied seawater system from the evaporator heat rejection section cooling seawater system to the evaporator heat rejection section final stage through supplied water pump, decarbonating equipment and deaerator.

The fluid is seawater 30 - 40°C, which is low in pH and high corrosiveness.

(c) Circulating brine system

This system covers the low temperature circulating brine system from the evaporator heat rejection section final stage to the heat recovery section final stage still inlet through the brine circulating pump and high temperature circulating brine system from the heat recovery section first stage still outlet to the heat recovery section first stage evaporation chamber inlet through the brine beater.

The fluid is concentrated seawater, the low temperature circulating brine system being 30 - 40°C and high temperature circulating brine system 90 - 120°C.

Brine in the system is deaerated and hence is less corrosive, however, the high temperature circulating brine system is high in temperature, which is capable of accelerating corrosion.

The high temperature circulating brine system is high in temperature, for which thermal stress will have to be taken up for examination.

(d) Discharge brine system

This is the discharge brine system from the evaporator heat rejection section final stage to the drain tank through the brine discharge pump.

The fluid is concentrated seawater deaerated and 30 - 40°C in temperature, which is less corrosive.

(e) Fresh water system

This is the fresh water system from the evaporation chamber heat rejection section final stage fresh water outlet to the distilled water pump.

The fluid is distilled water 30 - 40°C, which is almost free from corrosive property.

(f) Vapor system

The system covers the low-pressure vapor system from the vapor generating equipment to the brine heater and the medium-pressure vapor system to the ejector.

The vapor system is high in temperature, for which thermal stress will have to be taken up for examination.

(g) Condensate system

This is condensate system from the brine heater to the vapor generating equipment through the condensate pump.

The fluid is condensed water 100 - 130°C in temperature and low in corrosive property. For its high temperature, thermal stress will have to be taken up for examination.

(2) General policy for piping plan

What will work as a foundation for the design of piping are kind of fluid, rate of flow, pressure and temperature, and the matters to take into consideration are as follows:

(a) The fluid can be transported at a given rate under given temperature and pressure conditions.

(b) Economical.

(c) Easy to execute work and manage.

(d) Not to arrange piping lines which will invite excessive stress and vibration.

(3) General policy for pumping plan

For determination of the type of various pumps to be used for desalination plants, the matters to take into consideration are as follows:

(a) The equipment cost can be kept low in view of the overall equipment and, further, the power cost is cheap.

(b) Any equipment handling seawater and brine must be resistive them.

(c) Easy to operate and maintain.

(d) Suction requires high degree of vacuum in most cases, which must be satisfied.

(e) The equipment will require only limited space for installation and is arranged compactly on the whole.

(4) Fundamentals for design

(a) Selection of piping materials

The kind of the piping steel tubes has been specified in Japanese Industrial Standards. And for those piping tubes, limits of the working pressure and temperature are determined by the standards, and hence such tube as is suitable must be selected. The kind of the main piping steel tubes specified in JIS is shown in Table 5.3-6.

The tube material handles seawater or brine which is a corrosive fluid, so various anticorrosive measures must be taken into consideration.

(b) Selection of tube diameter

Design of the piping is commenced by determining tube diameter, for which the most economical diameter must be selected in consideration of kind of fluid, portion of use, pressure loss, etc.

Namely, to make the velocity in the tube high and the tube diameter small to save cost for piping arrangements is to involve increase in pressure loss and design pressure rising and thus to increase the power cost of the pump and so forth.

It is therefore necessary to determine the tube diameter by obtaining an economical velocity in the tube to minimize various expenses, including the plant construction cost and operating cost.

Velocity in the tube varies more or less according to the dimensions of the tube diameter and pressure loss. However, Table 5.3-7 shows velocities in the tube given separately by the fluid used generally.

In view of the tendency that the tube diameter increases to cope with larger equipment, which may involve a high construction cost, there may be a case for a velocity higher than those of the standard above to keep pace with the relative equipment for adjustment.

Pressure loss in the piping system can generally be obtained through the Fanning formula, which comprises friction inside the tube, knees, outlet/inlets of the water chamber, and resistance arising from valves and fittings.

Therefore, the tube diameter will be calculated according to proper velocity in the tube in consideration of what has been given above, and further the most economical tube diameter will have to be selected from among those specified in JIS.

(c) Selection of thickess

The necessary thikness of the tube will be obtained in consideration of influence of pressure, mechanical strength of materials, corrosion allowance, etc.

The necessary thickness of the steel tube subjected to internal pressure is obtainable generally through the following formula:

$$t = \frac{P \times D}{2\sigma + 0.8P} + C$$

where t represents thickness in cm, P internal pressure in kg/cm^2, D inside diameter in cm of the tube, σ allowable stress in kg/cm^2 and C corrosion allowance in cm.

Standard thickness of the specified piping will be selected with the value obtained through the above calculation as reference.

(d) Examination of thermal stress

The length of the tube changes in accordance proportionally as the tube temperature changes. Therefore where the tube is fixed at both ends, a compressive force is exerted by the length expanding thermally, and a compressive stress will be unavoidable on the tube. Compressive stress growing excessively large is capable of causing failure on the piping line, therefore a piping system heavy in temperature change must be designed with thermal stress taken fully into consideration.

For approximation to determine requirements of piping channel planning and detailed stress calculation for piping fixed only at both ends, the formula given in ASA. B. 30 Code for Pressure Piping (American standard for pressure piping) is generally used.

The standard specifies that further detailed stress calculation will be necessary in the area where the following formula is not satisfied:

$$\frac{DY}{(L-U)^2} \leq 8.2$$

where D represents nominal dia. in inches of the tube, Y total displacement in mm which is absorbed by the piping, L expansion length in m about the piping axle and U rectilinear distance in m between fixed points.

Where the above formula is not satisfied, a detailed calculation will be made further, and, as a result, necessary measures may be taken for changes in the channel, arrangement of expansion joints, etc.

Allowable stress range against the stress arising from thermal expandion, in this case, is given in the following formula by ASA standard:

$$S_A = f\,(1.25\,S_C + 0.25\,S_h)$$

where S_A represents allowable stress range against thermal stress, S_C basic permissible stress value of the material at minimum temperature expected at shutdown, S_h basic permissible stress value of the material at miximum temperature expected at normal operation or shutdown, f stress range decrease coefficient against the sum of temperature cycles.

(e) Selection of pump

The kind of pumps used for desalination plants under the multistage flash evaporation process may be considerable when including chemical injection pumps. However, the following main pumps are taken up here for description with the characters of specifications given first of all:

a) Seawater supply pump

This is a pump to supply raw seawater from the intake tank to the plant, the working fluid of which is seawater of normal temperature. It is necessary to select such a type of pump as is advantageous in net positive suction head (NPSH) against fluctuation of the suction level according to tide fluctuation and others.

For the material, corrosion resistance against seawater will have to be taken into consideration.

b) Brine circulating pump

This is a pump to draw the brine out of the evaporator heat rejection section final stage and introduce it to the heat recovery section first stage evaporation chamber by way of the evaporator heat recovery still and brine heater; the working fluid is concentrated seawater at 30 - 40°C.

Both discharge and total head are most important for all the pumps and further, degree of vacuum is very high as a suction condition, thus giving the severest conditions to this type of pump.

For the material, corrosion resistance against deaerated brine will have to be taken into consideration.

c) Brine discharge pump

This is a pump to draw the concentrated brine out of the evaporator heat rejection section final stage and introduce it to the drain tank; the working fluid is concentrated seawater at 30 - 40°C.

Suction conditions are severe as in the case of the brine circulating pump. However, the pump works with very little discharge.

For the material, corrosion resistance against deaerated brine must be taken into consideration as in the case of the brine cilculating pump.

d) Distilled water pump

The pump works for drawing the manufactured fresh water out of the evaporator heat rejection section final stage and taking it outside the system; the working fluid is distilled water at 30 - 40°C.

Discharge, total head and suction conditions are severe as in the case of the brine discharge pump. However, it is advantageous to have the suction level of distilled water higher than the brine level in the evaporation chamber.

The fluid being distilled water, anticorrosive measures are not necessary for the material.

e) Feed-water pump

This is a pump to feed seawater from evaporator heat rejection section cooling seawater system to the de-carbonator equipment; the working fluid is seawater at 30 - 40°C.

Positive pressure works on the suction side and hence the conditions are not severe.

For the material, corrosion resistance against seawater will have to be taken into consideration.

f) Condensate pump

This is a pump to pull the condensate out of the brine heater; the working fluid is condensed water at 100 - 130°C.

While the water-saturated vapor pressure works on the suction side at 100 - 130°C, there is almost no cooling exerted between the brine heater and pump, and hence NPSH is nearly equal to the height up to the condensing surface, giving severe suction conditions.

The fluid being condensed water, anticorrosive measures are not necessary. However, heat resistance must be taken into consideration.

The above has referred to characters of the main pumps, and the items necessary for consideration in selection of pumps are, kind of fluid, rate of flow, total head and suction conditions.

Total head is a pressure loss on the whole piping line, which can be obtained by determination of piping, relative equipment, fittings, etc.

In the case of desalination plants under the multistage flash evaporation process, the total head of each pump is below 100 m, therefore a single suction centrifugal pump for discharge below 30 m^3/m, a double suction centrifugal pump for 30 - 300 m^3/m and a diagonal pump for more than 300 m^3/m are generally used.

In the desalination plant, those pumps for which suction conditions are severe are the condensate pump, brine circulating pump to pull the brine out of evaporation chamber final stage, brine discharge pump and distilled water pump to pull out the manufactured fresh water. The degree of vacuum at the evaporation chamber final stage is maintained high, normally at 700 - 720 mmHg Vac., which is capable of giving rise to a cavitation at the runner inlet port, and hence an examination will be necessary beforehand.

Cavitation refers to a phenomenon where a portion coming below saturated vapor pressure locally in the fluid and then the liquid is vaporized to produce a cavity. The bubbles thus arising move in the stream and then crumble away at the high pressure zone. A repeat of the said phenomenon deteriorates performance of the pump, invites vibrations and noises, and may finally result in causing corrosion on the materials.

In planning the pump, therefore, attention must be paid particularly for determination of the suction conditions. A conception on NPSH is generally employed to examine whether or not the pump suction conditions are safe against occurrence of cavitation.

NPSH ava (NPSH that can be utilized by the pump)

This is a value representing how much the pressure immediate-before the pump runner inlet is higher than the liquid-saturated vapor pressure, which is determined according to the pump installation conditions.

NPSH req (NPSH that the pump requires)

This is minimum pressure (pressure drop at the pump runner inlet + water-saturated vapor pressure) required for the pump not to give rise to cavitation, which is inherent with the pump.

For safe operation of the pump, a relation NPSH ava > NPSH req must be maintained at all times accordingly.

It goes without saying that a pump with severe suction conditions will be installed in a low position underground. However, this is not necessarily easy for operation and maintenance in the case of large type pumps, and hence a pit barrel structure which allows only the runner part exerting an influence on the suction performance to be underground with the driving gear, inlet port and discharge port on the ground is popular.

In so far as the suction performance is concerned, an improvement of performance can be expected somewhat from keeping the pump running speed low. However, the equipment will have to be larger thereby, including the driving gear on the whole.

In consideration of every respect given above, it is necessary to select the most ideal pump construction.

(f) Selection of pump driving gear

For the pump driving gear, electric motors, steam turbines, gas turbines and internal combustion engines will be taken up. However, electric motors are generally used for desalination plants and steam turbines where the pump is of large type.

In the case of plants operating for a double purpose with power generation, the steam extracted from the generating steam turbine can be utilized on the pump driving steam turbine and further for a heat source of the brine heater, thus keeping the steam turbine power cost low. And the equipment cost of steam turbines for a small pump is high as compared with an electric motor, but the steam turbine will be low priced where the working pump is of large type. A steam turbine will therefore be profitable for a large pump.

For determination of the pump driving gear, it is therefore necessary to select an optimal pump driving gear through comparison of equipment cost and power cost and also in consideration of technical evaluation and environment for installation.

5.3.5. Steam Extraction Equipment

The steam extraction equipment for desalination plants under the multistage flash evaporation process must be designed so as to conform with various conditions given by the overall design of the plant. Multistage flash evaporation is carried out normally in a range covering about 120°C in the brine maximum temperature (where pretreatment is made through pH control process by sulfuric acid injection) to about 34°C. Namely, flash evaporation is available at pressures higher than atmospheric and at more than 100°C but in vacuum conditions (below atmospheric pressure) in the range below 100°C. Therefore it will be apparent that a major part of the evaporation range is kept in vacuum condition. The steam extraction equipment functions to keep the vacuum condition, which can be understood to play an exceedingly important part. Before giving a description of the design of the steam extraction equipment, an outline and character of vacuum conditions of a desalination plant under the multistage flash evaporation process are taken up first and then an outline will be given of selection of suitable steam extraction methods and design of the steam extraction equipment.

(1) Outline and character of vacuum conditions of a desalination plant under the multistage flash evaporation process

The flow of a desalination plant is as shown in paragraph 5.1. However, the main parts for which extraction is necessary are the evaporator, brine heater and deaerator.

(a) Gas extraction from brine heater

The brine heater is normally a shell-and-tube type heat exchanger with the heating steam coming on the shell side and the brine on the tube side. Gas extraction is carried out so as to eliminate non-condensable gas of the heating steam on the shell side, and allow the heating steam to come in contact efficiently with the heating surface at the same time, thereby improving heat transfer effect. The gas extracted from the brine heater is a mixture of non-condensable gas and steam and has a fairly high temperature, which is introduced to the evaporator high temperature stage or directly to the gas extraction equipment as the occasion demands. The former improves the heat recovery factor but on the other hand, deteriorates the heating efficiency of the evaporator, as non-condensable gas is introduced to the evaporator. The latter has characteristics contrary to the former, therefore it is important to determine a suitable extraction line through examination of how to make the best of the merits of both.

(b) Gas extraction from deaerator

The deaerator is classified into the external deaerating device and evaporator final stage deaerating device, particulars of which are as described in paragraph 5.2.2. A description will be given here on gas extraction from the deaerator for design of the gas extraction equipment. In the case of the external deairing device, the gas generated on the deaerator is sucked and discharged directly by the gas extraction equipment. However, in the case of the evaporator final stage deairing device, the generated gas has the steam partly condensed by the condenser on top of the evaporation chamber and then sucked and discharged by the gas extraction equipment. There may be a case where dissolved oxygen in the circulating brine is reduced by injecting a reducing agent in the supplied water. This treatment is called chemical injection deaerating, which is carried out often in combination with an external deairing device or evaporator final stage deaerating device. In this case, the supplied water, once deaerated, joins the brine. Therefore dissolved oxygen to be treated in the brine falls and thus the quantity of the chemical for reduction can be saved. Out of the non-condensable gas generated at evaporation of the circulating brine in the evaporator, carbonic acid gas will be prevented from being generated in consequence. However, other non-condensable gases (N_2, CO_2, etc.) are generated, for which extraction will be necessary.

(c) Gas extraction from evaporator

The gas generated in the evaporator hinders heat transfer of the condenser, therefore it is necessary to extract the non-condensable gas to improve the heating efficiency. A incorrect method of extraction may exert a detrimental influence on heating efficiency and a normal capacity for desalination cannot be secured consequently. Therefore a suitable method for extraction must be employed through examination. The following refers to the kinds and characteristics of the extraction methods from the evaporator.

(i) Cascade extraction method

As shown in Fig. 5.3-23, the gas extracted at the foregoing stage is introduced to the next stage. Therefore, non-condensable gas generated at the high temperature stage flows through each stage in sequence, exerting an influence gradually stronger on heat transfer. However, the gas being introduced from the foregoing stage, the heat loss is minimized and thus heat is recovered in satisfactory condition.

(ii) Barrel extraction method

As shown in Fig. 5.3-24, the gas extracted in each stage is sucked directly by the gas extraction device, therefore the non-condensable gas generated at each stage is free from influence on the other stages, and thus a decrease in heating efficiency is reduced from non-condensable gas. However, since the gas generated at each stage in the high temperature zone is sucked directly by the gas extraction device, the heat recovery factor becomes inferior, the gas to be sucked by the extraction device increases in quantity to accelerate loading on the extraction device, and thus the gas extraction device and the piping are to be large and complicated and high costing.

(iii) Combination of cascade and barrel methods

This is to make the best of both merits by combining the cascade and barrel methods, which is illustrated in Fig. 5.3-25 for example. Various ways of combination can be considered. However, what is important is to employ that most suitable to each case. In the case of the illustration, for example, the reason why the gas is extracted at the first stage is that the supplied water mixed with circulating brine contains non-condensable gas which cannot be extracted by the deaerator, which generates a considerable amount of gas at flash evaporation in the first stage. Therefore a detrimental influence on heat transfer by non-condensable gas on the latter stage side is to be avoided by extracting the non-condensable gas. With the similar object in view, the gas will be extracted from a suitable place, and further it is necessary to minimize the heat loss arising from sucking hot vapor on the high temperature stage into the gas extraction device. As mentioned, it is important to plot out in the most profitable way through employing an optimal method for gas extraction with the influence of non-condensable gas and heat loss taken into consideration.

(2) Selection of gas extraction device

Vacuumizing equipment for the gas extraction device is of various types, which must be carefully selected for the scope of application. Main types and characteristics are given below.

For desalination plants operating under the multistage flash evaporation process, design is normally made so as to have the final stage temperature at 34°C or so against the intake seawater temperature working at 25°C. To secure a degree of vacuum to work at saturated vapor pressure, that of 1 - 40 mmHg abs or so will be necessary with the piping pressure loss as far as the gas extraction device taken into consideration, for which a steam ejector is used generally. The reason why the steam ejector is so popular is that the available degree of vacuum is coincidental and the steam ejector is economical as it has less trouble because it has no rotating part to suck the mixture of non-condensable gas and vapor.

(3) Steam ejector

There is a case where a mechanical vacuum pump is used partially other than a steam ejector for the gas extraction device of a desalination plant under the multistage flash evaporation process. However, the situation is such that the steam ejector is serviceable in most cases, and hence a description will be given here on outline of the steam ejector, calculation of the quantity of sucked gas and design of the steam ejector.

(i) Outline of steam ejector

As shown in Fig. 5.3-26, the main part of the steam ejector comprises a nozzle, diffuser and suction chamber. In addition the part through which to introduce the working steam to the nozzle is called the steam chamber. The working steam is injected through the nozzle, which is of divergent type, and the pressure energy is translated into velocity energy here to jet at high speed. In this case the suction chamber is vacuumized to suck in the gas. The sucked gas (being a mixture of non-condensable gas and vapor) is mixed with the high speed steam current to come into the diffuser, where it is decelerated gradually and retranslated into pressure energy, thus coming out of the diffuser. Vacuum is generated by sucking the gas through such a process. However, there are left many points unsettled logically, and thus the real circumstances are such that the design is performed on test data secured individually by each company. In amplifying upon the function of each part further, the nozzle functions to convert the working steam pressure into velocity energy, and a divergent nozzle is used. The nozzle aperture is determined according to the necessary quantity of steam obtained through the suction rate and required degree of vacuum. Upon determination of the nozzle aperture, steam consumption depends on the working steam pressure regardless of the degree of vacuum in the suction chamber. The divergent nozzle is used for translating the pressure into velocity energy thoroughly, thereby jetting the working steam through the nozzle at supersonic speed. Divergence of the nozzle normally is 10 - 30°. However, the nozzle for a high vacuum ejector is large in the rate of expansion and has a large section at its end. The suction chamber mixes the working steam and the sucked gas in it, and its construction will be such that the sucked gas can be introduced to the diffuser inlet without disturbing its flow. The diffuser is a part to compress the mixed fluid, comprising a taper zone, parallel zone and divergent zone; determination of the aperture for the throat zone (parallel zone) will largely influence the performance of ejector. Where the working steam pressure is constant, the ratio of nozzle throat aperture to diffuser throat aperture is one of the factors to determine whether the ejector is used at high vacuum or low. In the case of an ejector to be used at low vacuum of more than about 100 mmHg abs which is used for atmospheric release, the diffuser aperture is 2 - 5 times as large as the nozzle aperture. However, in the case of a booster of high vacuum, the said aperture increases 10 - 20 times. The value must be determined according to the working conditions of steam pressure and others. Where the final stage for the desalination plant has a temperature of 34°C, it is necessary to secure the ejector-generated vacuum at 20 - 40 mmHg abs in consideration of the piping pressure loss covering the final stage to the ejector. The degree of vacuum generated at one stage of the steam ejector is limited to 60 - 80 mmHg abs for atmospheric release by means of steam at 7 kg/cm^{2}G or so, but will be practical at 100 mmHg abs in consideration of economical steam consumption. It is therefore necessary to arrange the ejector for desalination plants in more than 2 stages. A flow for arranging the desalination plant with a 2-stage ejector is given below. The gaseous body extracted from the final stage contains vapor, non-condensable gas arising at evaporation of the brine and air leaking into the evaporator from outside. If the vapor content of the gaseous body can be condensed by condenser to be eliminated, then the ejector will be loaded less. In case much vapor is contained in the extracted gaseous body or the condenser cooling water is easily utilized, the condenser is used in the form of a vent condenser. A flow using the vent condenser is shown in Fig. 5.3-27.

Here the condenser is shown in a shell-and-tube type heat exchanger. However, a barometric condenser may be used. In the case of a barometric condenser, extracted gas or ejector outlet mixed gas is subjected to direct contact heat exchange with cooling water, therefore the condensed drainage cannot be recovered. In the flow sheet given above, the gaseous body extracted from the final stage is condensed for vapor content only by the condenser, and the saturated vapor corresponding to partial pressure in the vent condenser and non-condensable gas are sucked by the first stage ejector to be discharges. After condensing for vapor content only by the condenser, the discharged mixed gas (working steam and sucked mixed gas) is sucked by the second stage ejector and discharged to the second stage condenser. As in the case of the first stage, the mixed gas is condensed for vapor content only by the second stage condenser, and the non-condensable gas is released in the atmosphere. The vapor condensed by each condenser to drainage is recovered in the distilled water reservoir, as illustrated, by way of the vent condenser. In a plant actually in operation, the drainage may containing impurities and hence it may be taken out separately from the distilled water manufactured. As referred to before, the point for extraction being kept under various pressure conditions, the design of an optimum gas extraction device will have to be performed by coupling the point to the most suitable place in the flow.

(4) Composition of extracted gas

(i) Composition of gas extracted from brine heater

Where heating steam is condensed on the drum side of the brine heater, non-condensable gas in the heating steam is accumulated little by little to exert a detrimental influence upon heat transfer. Therefore, the gas is extracted to have non-condensable gas eliminated. The extracted gas is a hot gas in which non-condensable gas and heating steam are mixed.

(ii) Composition of gas extracted from deaerator

A considerable amount of non-condensable gas is generated in the deaerator through being eliminated from seawater. In case the gas is extracted through evaporating the seawater by steam, a large amount of vapor and non-condensable gas are generated, which must be extracted, and hence the ejector load will have to be minimized by arranging a vent condenser.

(iii) Composition of gas extracted from evaporator

Non-condensable gas dissolved in the brine is generated in the evaporator according to evaporation of the brine. Where the evaporation chamber is kept below atmospheric pressure, air leaks into the chamber from outside. These non-condensable gases and vapor will have to be extracted.

As mentioned above, gas extraction will vary sharply in quantity through process design of the plant. It is therefore necessary to optimize the situation thoroughly at the stage of basic design of the overall plant. Then, non-condensable gas is calculated for quantity according to individual conditions for generation in the deaerator, brine heater and evaporator, as described before.

(5) Design of steam ejector

The overall gas extraction device will be optimized in consideration of the ratio of non-condensable gas quantity calculated in (4) to generated vapor. While cooling water temperature, extracted gas composition, temperature, etc. are influential, employment of the vent condenser will finally be determined through cost comparison. The degree of vacuum generated in the ejector will be determined at the evaporator final stage in consideration of pressure loss in the piping. For the working steam pressure, the minimum pressure immediately before the ejector will be set as design pressure; superheated vapor, saturated vapor and wet vapor will be taken up for vaporized conditions. However, wet vapor must be avoided as the ejector performance is subject largely to a detrimental influence by drainage. The superheated vapor is somewhat inferior in performance to the saturated vapor. However, there is no big difference between the two. Therefore at the working steam supply point, vapor superheated somewhat will be supplied in consideration of heat diffusion in the piping. Temperature and composition of the extracted gas are essential factors for heat exchange design on the vent condenser and each condenser and must be clarified for its necessary weight for determination of the ejector extraction capacity. A description has been given as above for guiding principles of ejector system engineering; an optimum design will have to be performed in making the best of characteristics of the ejector with running cost, construction cost and performance taken up thoroughly for examination.

5.4. Metallic Materials and Corrosion Prevention

5.4.1. Corrosive Environments and Materials

As described, there are various methods available for desalination. of which the multistage flush evaporation method has the characteristic of accelerating metallic corrosion most actively through treating the seawater at high temperatures up to 120°C with a high flow rate. The equipment then comprises, as shown in the flow sheet, a heater, an evaporator vessel for the heat recovery section and an evaporator vessel for the heat rejection section of the main part, a seawater intake, a decarbonator, a deaerator, pumps and piping. The evaporator vessel consists of a flash chamber, a condenser enclosing a heating tube bundle, a demister and a fresh water trough. Materials for each device and part are standardized to some extent through usual know-how on seawater resisting materials, corrosion tests in laboratory and test plants for desalination device materials and through practice with operating plants. However, a selection of materials including new materials to be employed is still important in consideration of life and maintenance required for the plant, difference in the components of seawater at the site, engineering peculiarities and differences in the operating conditions. For example, a combination of materials which was successful at one plant is not necessarily going to lead to success at another plant different in site or in design, thus bringing about corrosion troubles in many plants. Some designers insists, therefore, that a corrosion test should be kept running for one year at least in a test plant at a site before establishing a large plant.

The environment for each part of the equipment is given in Table 5.4-1. What causes problems with reference to metallic corrosion in the said environment will be brine, seawater and vapor containing noncondensable gases, of which brine is particularly influential on corrosion with respect to dissolved oxygen concentration (DO) and temperature, and seawater with respect to sand, mud, organisms and contaminants of S^{-2}, NH_4^+, etc. which are mixed in it. Metallic materials to be used are mainly mild steel, copper alloys, stainless steel and titanium, of which mild steel is very popular for evaporator vessels, heater vessels, piping, etc. The recent tendency, however, is that only such areas as are in contact with clean vapor and fresh water are left exposed without any corrosion preventive measures; other areas are applied with metal lining of copper alloy, stainless steel, etc. and mortar lining for high temperature zones and with mortar lining, organic coating and lining or with cathodic protection for other zones below medium temperature. The performance of organic coating and lining or mortar lining is subject largely to how to work is executed. However, the mortar lining is advantageous to facilitate repair work.

5.4.2. Heat Transfer Tubes, Tube Plants and Water Chamber

The tube bundle consisting of heat transfer tubes and tube plates functions like the heart of the plant, and its cost is 10 through 30%, for a plant of large capacity. The matter of selecting materials for the tube bundles heat transfer tubes particularly is therefore very important, and thus a study on the materials to select has been carried out in each country.

A.D. Little, Inc. reports in a survey as of 1971 that heat transfer tubing materials for MSF plant are copper alloys for 99.2% as shown in Table 5.4-2, and titanium is used a little.

(1) Heat transfer tubing materials

The object of project research done by the Industrial Science & Technology Agency on heat transfer tubing materials was to reconsider the adaptability of all the main tubing materials, such as the iron group, aluminum alloys, titanium, copper alloys, etc. under the same conditions, thereby recommending which are optimum for work in a large MSF plant. The Heat Transfer Tubing Materials Working Group (Prof. Go Okamoto as chief investigator) was set up as a central body for execution of researchs; then Japan Gasoline and Mitsubishi Heavy Industries were entrusted with the research to carry out the loop test (3 loops coming in A, B and C arranged; small test pieces in plate done chiefly) for screening all the test materials under the deaerated brine environment of MSF plants and field tests to check practicability of the materials screened by the loop test through equipping a small MSF plant (24 m^3/day, heat recovery section in 4 states, heat rejection section in 2 stages) with the tubes made of the selected materials at Chigasaki Test Facility. In addition, basic research to examine brine environment and conditions under which cheap heat transfer tubing materials of the iron group (carbon steel and low alloy steel) can be used was carried out by the National Chemical Laboratory. In the light of those fruitful results, heat transfer tubing materials were selected (titanium 0.4 mm tube for the brine heater, aluminum brass for the heat recovery section, titanium 0.4 mm tube for the heat rejection section) for the Oita Test Module Plant and while carrying out an operation test, the behavior of the heat transfer tubings was taken into consideration, and then after the module test various analyses and tests were made for hydrogen absorption of titanium tubes.

The following is to introduce, for every material, results obtained through the National R & D Project and literature and information available at home and abroad:

(a) Materials of the iron group

There are several reports made on low alloy steels (including carbon steel), stainless steel and coated steel materials to work for heat transfer tubing of an MSF plant. However, a test plant or equipment for desalination using these heat transfer tubing materials has never been manufactured so far.

(i) Low alloy steels

What is attractive with low alloy steels including carbon steel, is its cheap cost as a material.

However, a survey made since then clarifies that while the regular length tube below 6 m is cheap, the longer its length, the higher its cost, thus the price factor is not good as far as the long-length tube is concerned. However, the situation is due to commercial practice for the most part, and the evaluation is hence not absolute for the future.

The basic research at the National Chemical Laboratory looked at mild steel and $2\frac{1}{4}$ Cr-1 Mo steel, and the brine environment and conditions (DO, pH, temperature, flow rate, etc.) to improve corrosion resistance were taken into consideration. Fig. 5.4-1 illustrates the influence of DO and pH at 120°C by means of a 6% NaCℓ solution, where it is found that the corrosion resistance of steel can be improved remarkably by decreasing DO below 1 ppb and increasing pH up to 9.5. However, the improvement in corrosion resistance of steel is not to be expected so much by decreasing DO extremely around pH 7.5 of brine for an MSF plant employing pH control. There is no difference observed in the corrosion of mild steel and $2\frac{1}{4}$ Cr-1 Mo steel in a test using an NaCℓ solution. However, corrosion resistance of low Cr steel will be increased more than mild steel from its protective film having been improved in property when using an NaCℓ solution mixed with Mg^{2+} or seawater as shown in Fig. 5.4-2. The figure is an arrhenius plot illustrating the influence of liquid temperature, wherein mild steel remains straight but low Cr steel bends at 100°C, thus changing the corrosion reaction mechanism.

The result obtained through the 1,000-hour loop test at Chigasaki shows that the corrosion rate of low alloy steel becomes low generally at 100°C or below and DO < 10 ppb and corrosion is less as the steels contain more Cr.

However, local corrosion starts to increase at Cr > 5% and hence 2 - 3% Cr steel will be best. Since the corrosion rate of low alloy steel tube is subject to influence by DO as mentioned, the use is limited to the heat recovery section only in the case of an MSF plant.

The test result on heat transfer tubes made of 2 - 3Cr steel screened through the above loop test shows that 3Cr-1Aℓ steel, 3Cr-1Ni steel, 2Cr-0.1Mo-1.2Mn-Cu steel, etc. are comparatively good. However, the degree of corrosion by temperature is not clear. Fig. 5.4-3 shows a 5,000-hour field test result on 3 Cr-1 Aℓ steel (employed on the heat recovery section). The figure illustrates the corrosion rate at every 1,000 hours by the stage of the heat recovery section, wherein the corrosion rate of the second stage is somewhat large and decreases with time showing a tendency of approach to a critical value. The critical value is 0.1 - 0.15 mm/y or so, which can be said to be considerably large. In the field test moreover, sludge sediments in the inside of heating tubes, thus the overall heat transfer coefficient gradually decreases. Among those materials which were subjected to test in the heat recovery section, low alloy steels were largest in the rate of sludge sedimentation. This has something to do with its large corrosion rate. The heat transfer tube is exposed to vapor outside, and hence the outside of low alloy steel tubes was subjected to corrosion of about 1/10 of the inside in the field test.

As compared with the results of the loop test and basic researches, the corrosion rate of low alloy steels was considerably large in the field test. That sludge containing much iron dissolving out of the steel vessel sticks to the inside of tubes to exert a detrimental influence on the protective coat of low alloy steels (magnetite) may be a cause. For prevention of the phenomenon it is necessary to wash the inside of steel tubes by the sponge ball method during operation and in fact, the effect was manifested at another plant at Chigasaki.

On the other hand, air comes into the interior of the equipment during shutdown of the plant and in this case, corrosion grows on both the inside and outside of the low alloy steel tubes. To prevent corrosion from growing, the tubes will have to be charged with deaerated seawater, nitrogen gas, etc., which may be pointed out as a large shortcoming unavoidable with steel tubes.

(ii) Stainless steels

What is good about stainless steel is that it is corrosion-resistant and less subject to sludge sticking. However, the corrosion-resistance of stainless steel against seawater and brine is effective for general corrosion but problematical in pitting corrosion or crevice corrosion.

Screening was hence carried out at the loop test for pitting resistant stainless steels under development by each manufacturer. The result was evaluated by amount and maximum depth of pitting. For heat transfer tubes the maximum depth is more important practically, and temperature and DO of the brine environment are particularly influential.

The test pieces were screened about half way through the loop test; a field test was then carried out for stainless steels, which is generally dearer than copper alloy, on both the heat rejection section and the brine heater for which copper alloys are capable of causing trouble and heat transfer tubes are less used. The result was that local corrosion was negligible as compared with the loop test on the whole and that only micropits (below 20μ in depth, 20 - 30μ in diameter) were produced on that of good corrosion resistance. For example, 22Cr-26Ni-5Mo-Ti steel and 25Cr-5Ni-2Mo steel were as comparatively acceptable as austenite steel and two-phase steel respectively. However, stainless steel entirely free from micropits could not be found.

There is a tendency in the United States to increasingly use stainless steel heat transfer tubes for a condenser using seawater at thermal power plants, and AL-6X steel (24.5Ni-20.25Cr-6.25Mo-0.025C), for example, is recommended recently. In case 316 steel cheaper than that is used, the use of sponge balls is regarded as effective for prevention of pitting corrosion, and for this reason effectiveness will have to be checked on a combination of stainless steel heat transfer tubes and the sponge ball method for an MSF plant.

(iii) Coated steel materials

Steel coated with chromium, aluminum, titanium, etc. and steel coated with the epoxy group and phenol group in suitable compounds were checked in the loop test Then, a field test was carried out on heat transfer tubes for which chromized steel and aluminized steel as metal-coated steel and Meton, Säkapfen and Corsite as resin-coated steel were selected. Phenomena were observed in which the ground iron dissolved out of pinholes and the chromium layer came off in the case of chromized steel tube (200 - 300μ coating) and aluminum dissolved out severely at the high temperature stage to leave the ground iron exposed partly in the case of aluminized steel tube (100 - 150μ coating). Meanwhile, the resin-coated steel tubes (coated 100 - 200μ) were satisfactory in corrosion resistance but lowest in overall heat transfer coefficient of all the materials tested. Thus all the coated tubes subjected to test were found impractical after all.

(b) Aluminum alloys

There are many reports given on adaptability of aluminum alloy heat transfer tubes to MSF plants with reference particularly to OSW. Some small test plants using aluminum alloys for tube bundles, water chambers and piping were once put in operation. However, no case has been found for application on a practical plant.

At the National R & D Project, 1,000 series of pure aluminum standardized, corrosion resistant aluminum alloys 3,000 series of Aℓ-Mn, 5,000 series of Aℓ-Mg, 6,000 series of Aℓ-Mg-Si and alclad were subjected to testing.

Difference in corrosion was scarecely observed in the loop test according to difference in environment and conditions and kinds of aluminum alloy, and the corrosion rate of each alloy was 0.06 mm/y or so. However, the mean value of the maximum dapth of pitting corrosion reached 0.7 mm/y, and crevice corrosion was observed in the clearance with the test piece holder.

The field test was carried out with heat transfer tubes of each material mentioned above in the brine heater, heat recovery section and heat rejection section (tube plates being 5052 and 6061). The local corrosion characteristic of the heat transfer tubes was evaluated in surface roughness after the test: the heat rejection section was in the best condition and the third stage of the heat recovery unit was worst, and #3003 was found most stable of all as the kind of alloy.

In parallel almost with the field test, a 5,000-hour test on the aluminum heat transfer tubes was carried out at the joint reasearch of the aluminum alloy-made MSF plant at Chigasaki. The said plant was a 20 m^3/day short-tube type one which was manufactured by the Japan Light Metals Association, using aluminum alloys for tube bundles, water chambers and piping, and all the aluminum materials were taken up for research. In evaluation of local corrosion on the heat transfer tubes at the test in maximum depth of pitting corrosion, that of #3004 of the 3,000 series was most stable in corrosion resistance as in the case of the field test, but #5050 was found defective in producing a through hole in the brine heater. Then, in the evaluation by stage, the final stage of the heat rejection section was rather good and an appreciable difference was not observed among other stages.

In both cases of the field test and the joint research, the rate of sludge deposition on the heat transfer tubes of the highest temperature zone in the brine heater was large, and could hardly be removed, maybe due to a binding condition with corrosion products of aluminum at high temperatures.

(c) Titanium

Titanium usually maintains a corrosion resistance through its strong passive film in neutral seawater and brine but is capable of increasing crevice corrosion and pitting corrosion in accordance with rising temperature and salt density. For example, Bohlmann, etc. reported that crevice corrosion grew easily in titanium in seawater (static state particularly) of more than 100°C, this phenomenon was not influenced by DO of the seawater. The result given in

Fig. 5.4-4 was also reported on local corrosion characteristics of titanium according to density and temperature of the NaCℓ solution; the circulating brine in the water chamber and the heat transfer tubes corresponding to NaCℓ 5.4%, which may cause crevice corrosion at temperatures greater than 105°C. Fig. 5.4-5 represents a test result by N.G. Feige, wherein a tendency is shown that resistance to pitting corrosion and crevice corrosion increases in the order of Ti-2Ni > Ti-0.2Pd > titanium.

In the loope test of this project plate test pieces of three kinds of pure titanium, welded titanium and Ti-0.15Pd were tested. However, corrosion loss was scarcely observed. In the field test a thin-walled titanium tube 0.3 mm thick was fitted by expansion on a tube plate of titanium and naval brass, and a 2,000-hour test (at each stage from the brine heater to the heat rejection section) and a 5,000-hour test (brine heater only) were carried out: the titanium tube and tube plate were nearly satisfactory in corrosion resistance, and the naval brass tube plate was also free from trace of corrosion. This might be due to the crevice of titanium tube and titanium tube plate or the naval brass tube plate having been subjected to cathodic protection by steel piping.

There is a case where a crevice corrosion was produced on a combination of titanium tube and titanium tube plate at brine of more than 80°C at the MSF plant in operation. For prevention, it is most effective to remove the clearance by seal welding of the tube end and the tube plate. In the combination of the titanium tube and the copper alloy tube plate, cathodic protection against galvanic corrosion of copper alloy is to be executed from the beginning. However, overprotection must be avoided in this case by potential controlling or by employing a sacrifical electrode lest titanium should be polarized below -650 mV which is capable of giving rise to the brittleness described below due to hydrogen absorption.

The thermal conductivity of titanium is extremely low, (14.0 Kcal/m·hr·°C) 1/6 of aluminum brass, and therefore the tube must be as thin-walled as possible not only for reduction of cost but also for a large overall heat transfer coefficient and at the same time, a decrease in sludge deposition can be expected. The titanium tubes employed for the conventional power plant and MSF plant have a wall thickness of 0.5 - 0.7 mm for reasonable strength. However, thin-walled welded tubes of 0.3 and 0.4 mm were subjected to test for the National R & D Project to cope with the above requirements. The overall heat transfer coefficient U (averaged from the brine heater to the heat rejection section) of the titanium tubes 0.3 mm thick in the field test was regarded as coming on nearly the same level with U of aluminum brass tube 1 mm thick (there may be a case where titanium is inferior at the brine heater) even under conditions where the tubes were kept clean immediately after commencing the test or after sludge accumulated through using the tubes for more than 2,000 hours.

Meanwhile, thin-walled titanium tubes 0.3 mm thick or so being low in rigidity, trouble occurs with reference to tube oscillations due to steam flow outside the tube. Therefore the oscillation characteristic of 0.3 mm titanium tubes was elucidated by means of an oscillation testing device which was similar in shape to the condenser working at the MSF plant of 10,000 m^3/day scale in this project, thus giving an effective prevention measure for design.

It is well known that titanium easily absorbs hydrogen and produces hydride (TiH_2) of black needle crystals in reaction with hydrogen. The solubility of hydrogen in pure titanium at normal temperatures is 20 - 50 ppm and where hydrogen is solidified more than that, a hydride will be separated at room temperature under conditions of equilibrium. JIS for pure titanium material specifies that hydrogen be less than

130 ppm. However, hydrogen of normal products is less than 50 ppm in the majority of cases. Separation of hydride increases generally as hydrogen density increases, and thus the titanium material becomes brittle. A problem in this case is limiting the quantity of hydrogen which causes brittleness. For example, a result has been obtained through test by Lenning, etc. that tensile strength and hardness gradually increase in accordance with hydrogen increase but elongation and contraction of area gradually decrease. However, the threshold value of hydrogen capable of causing brittleness is not clarified.

Hydrogen absorption of the titanium heat transfer tubes was no problem in this project at first. However, titanium tubes in the brine heater of a small MSF plant for deaeration and decarbonation study got brittle after repeating the test (8,800 hours all told) through pickling the heat transfer tubes many times, and hydrogen analysis conducted showed that the titanium tubes on higher stages had a lot of hydrogen with a maximum of 870 ppm at the brine heater. Then, various laboratory researches, including hydrogen analysis of titanium test pieces after loop and field tests, were made through the cooperation of the nonferrous subgroup (The Furukawa Electric Co., Ltd., Sumitomo Light Metal Industries Ltd. and Kobe Steel Ltd.), and the following phenomena were clarified: 1) Hydrogen absorption was not observed in the loop test. 2) Hydrogen absorption (264 ppm max.) took place clearly on the titanium tubes of the brine heater but was not observed on the heat rejection section at 2,000-hour field test. Then as the result of a 5,000-hour field test, hydrogen quantity was not so related with the length of test time (1,000 - 5,000 hours). 3) In a test with 120°C deaerated natural seawater in an autoclave, hydrogen was almost not absorbed by mere dipping tests of titanium, however, hydrogen absorption took place by holding titanium potential below -750 mV. 4) As shown in Fig. 5.4-6, titanium sharply decreases in elongation at more than 650 ppm in hydrogen absorption.

In short, hydrogen absorption phenomena were distinctly observed with the titanium tubes fitted on the high temperature stage or particularly on the brine heater of the MSF test plant. Thin-walled titanium tubes (0.4 mm mostly, 0.3 mm partly) were fitted on the brine heater and heat rejection section later at the Oita Module Test Plant and after termination of tests in low temperature operation for 3,340 hours and high temperature for 3,280 hours, hydrogen analysis was conducted on the titanium tubes, thereby detecting 404 ppm max. on the brine heater (at high temperature operation, 115 - 121°C) but no increase in hydrogen quantity on the heat rejection section (at high temperature operation, 84 - 93°C).

Meanwhile, it is informed that favorable operation has been maintained with the well known MSF plant bearing on the tube bundles made of titanium (designed output 2,840 m^3/day, max. output 5,680 m^3/day) of the Harvey Aluminum Corp. at Caribbean St. Croix Island since it was constructed in 1965, and nothing has been reported on troubles with the titanium heat transfer tubes (0.7 mm thick) except sticking of organisms to the seawater inlet port. Maximum temperatures of the brine for the said plant are reported at two levels of 85°C at designed output and 121°C at max. output, and if no problems occurred, like hydrogen brittleness on titanium tubes for the brine heater, it might be because of having operated normally at 85°C max.

In analysing the above tests and achievements, it may be considered that hydrogen absorption phenomena of titanium tubes is capable of arising generally at high temperatures, but not below 90°C in brine temperature unless particularly conditioned. And from the fact that the brittleness phenomenon has taken place apparently at 870 ppm in hydrogen, the limiting concentration of hydrogen will be lower than that, probably around 600 ppm. These limiting temperatures and concentrations will have to be clarified and, at the same time, the causes should be traced accurately.

For intake of polluted seawater including S^{2-} and NH_4^+ at the MSF plant, what is most problematical is the heat transfer tubes for the heat rejection section, and, where possible, titanium tubes must be employed.

(d) Copper alloy

Copper alloys have been regarded as ideal materials for the heat transfer tube from their popularity in the past and superior heat transfer properties in addition to corrosion resistance. Usually the heat transfer tube materials for desalination plants are copper alloys, and for selection of heat transfer tube materials at present, other than titanium aluminum brass and cupronickel (70/30Cu-Ni, 90/10Cu-Ni) are taken up. In fact, various test results obtained by OSW for heat transfer tube materials of the copper group endorse the particular corrosion resisting properties of the above three kinds of alloys, and the order is nearly 70/30Cu-Ni(715) > 90/10Cu-Ni(706) > aluminum brass (687) with little difference between 706 and 687 generally. The newest test result of all, which was obtained through the CDA-OSW corrosion test loop of the OSW Material Test Center and desalination plant, is introduced below.

The corrosion test loop reproduces the environment and conditions in the interior of heat transfer tubes for the brine heater at a multiple-effect multistage flash desalination plant, exposing copper alloy to salt water at 121°C, pH 7.4 and 72 ppb in DO for 697 days at the longest. The test result shows the corrosion rate is inclined to drop after 365 days at 4 ft/s rate of flow, to balance after 365 days at 8 ft/s and then inclined to increase after 696 days at 16 ft/s. The order, according to the kind of alloy, is, 715 at the top, and then 706 and 687 following with little difference between each other.

Next, the CDA-OSW desalination plant (22.7 m^3/day) was constructed in the same place after the above loop was dismentled and was designed so as to reproduce all the environment of desalination plants working under the evaporation method.

Table 5.4-3 gives the order of various copper alloys for corrosion resistance at each part of the plant and the scope of corrosion rate. The corrosion rate is particularly high at the heat recovery section where DO is comparatively high, and a difference between alloys is revealed most definitely; the rate falls, on the other hand, at the brine heater outlet with low DO and the vertical tube evaporator, and a difference between alloys becomes less noticeable. Corrosion resistance was highest with 715 and 613 on the whole. Fig. 5.4-7 and Fig. 5.4-8 illustrate how temperature and DO are influential by means of a testing tube bundle letting flow discharged seawater from the plant. The corrosion resistance of 687 sharply deteriorates as temperature and DO concentration rise, but 715 and 706 keep superior corrosion resistance. The tendency was such that the corrosion resistance of those copper alloys was improved and moreover, the difference disappeared between alloys where DO concentration or temperature fell.

On the other hand, since the scope of test research on the copper alloy heat transfer tubes in this project was mainly a test of new copper alloys which keep corrosion resistance even with or superior to the copper alloys in use at present and are cheap in addition, testing on dear cupronickel was limited to a loop test, and the alloys manufactured for trial were taken up for comparative examination with reference to aluminum brass in the field test.

Firstly, at the loop test, the corrosion resistance of copper alloys at 100°C were superior on the whole but corrosion increased considerably at 120°C. The copper alloy 70/30Cu-Ni was best of all, then comes a group of 90/10Cu-Ni and aluminum brass. A better result was shown with 90/10Cu-Ni than aluminum brass as DO got rather high at 120°C, but on the other hand, there was nearly the same corrosion resistance between aluminum brass and 90/10Cu-Ni even at high temperatures under a low oxygen environment DO < 10 ppb. Among those of trial manufacture, the alloy 35Zn-1Al-0.04As-Cu was found to be a good prospect.

Secondly, at the 2,000-hour field test, various copper alloys, manufactured for trial, and pollution-proof copper alloy (B-KCB) were subjected to tests with reference to aluminum brass and admiralty. The result is as given in Fig. 5.4-9, wherein two kinds of aluminum brass brought the best result. At the 5,000-hour field test only aluminum brass was subjected to test at each stage from the heat rejection section to the brine heater. As shown in Fig. 5.4-10, corrosion was severest at the brine heater but a tendency to saturation was indicated at 5,000 hours. That is, an appreciation of the aptitude of aluminum brass for the brine heater at 121°C max. may vary according to the way of emphasizing severe corrosion or the tendency of corrosion saturation at 5,000 hours. As a precaution, it may be wise to avoid use of aluminum brass at 121°C At the 5,000-hour test, sludge deposited most severely at the second stage and the first stage but less at the sides higher and lower in temperature. As for influence of the test duration, sludge deposition was inclined to saturate at 5,000 hours at any stage and the heat transfer characteristic did not descend so much thereafter. However, sludge deposition might be subjected to an influence by seasonal and local characteristics of the seawater and equipment characteristics.

Then, where bottom sediment rich in nutrients for bateria is fed into heat transfer tubes for the heat recovery section, the inside of the tubes at the low temperature stages, which are filled with deaerated brine, becomes optimal for sulfate reducing bacteria to multiply, thus generating hydrogen sulfide. The copper alloy heat transfer tubes are then subject to pitting corrosion macroscopically and intergranular corrosion microscopically. There are contradictory reports referring to the comparative merits of cupronickel and aluminum brass for polluted seawater, which must be due to varying circumstances of the polluted seawater. For example, cupronickel is superior to ammonia but aluminum brass is superior, on the other hand, to hydrogen sulfide, and thus the matter of how to select either against polluted seawater is not easy. It is recommended after all that in case water is taken from a polluted sea area normally titanium or Sn-Aℓ-Cu group (pollution-proof copper alloy) is employed, and in case seawater is polluted mainly by medium or slight hydrogen sulfide aluminum brass is used with the condition of ferrous sulfate addition to the seawater which forms corrosion resistant film containing iron. Care should be taken, at any rate, that exposure of the a surface of ordinary copper alloy to polluted seawater is capable of leading to the worst results.

Copper alloy easily incurs sand erosion from seawater including much sand, and 70/30Cu-Ni containing 2% of Fe will be optimal in this case.

(2) Tube plate and water chamber

According to the material survey by A.D. Little, Inc. in 1971, tube plate is copper alloy for 95.4% to cope with heat transfer tubes being copper alloy for the most part. Then, 90/10Cu-Ni and naval brass are predominant for 81% in this case. Next come bronze or aluminum bronze, aluminum brass and 70/30Cu-Ni in that order. However, these total only 13%. To curtail the cost there may be a case where 90/10Cu-Ni clad steel or Monel clad steel is used instead of solid copper alloy. This is expected to increase in the future.

Meanwhile, the material survey on the water chamber refers to using solid 90/10Cu-Ni and 70/30Cu-Ni in part. However, it is made of mild steel for 76% and protected with nonmetallic lining or metal clad mostly, the part left bare being only 14%. Epoxy, rubber, FRP, etc. are used for organic lining. However, these are not durable at the high temperature stage. Mortar is used for inorganic lining, which is durable even at the high temperature stage and further advantageous to facilitate servicing work. Metallic clad is suitable for use on medium and high temperature stages, for which 90/10Cu-Ni, Monel, stainless steels and Aℓ-Ni bronze are used. The said survey further refers to the water chamber having been applied with some cathodic protection at 43 plants. Cathodic protection by galvanic coupling using sacrificial anodes like zinc or iron is used for the most part; iron anode are suitable for protection of the tube end and tube plate of copper alloy from corrosion. On the other hand, cathodic protection by impressed current was employed only at 5 plants with platinum-plated titanium electrodes working as anodes. It is then reported that the cathodic protection of mild steel in deaerated brine at high temperature is not only difficult but also less effective. However, further study will be necessary under existing circumstances.

Materials must be selected carefully on the presumption that these heat transfer tubes, tube plates and water chambers are generally in contact with each other electrically, unless the water chamber is lined internally with nonmetallic material. Fig. 5.4-11 represents corrosion potentials of metals and alloys in flowing seawater at ordinary temperatures, which is applicable straight to the heat rejection section. In the case of the heat recovery section and the brine heater, the deaerated brine is objective and temperature is generally high in addition, and hence the absolute value for potential in the figure cannot be utilized. However, it has been experienced practically that it is useful for the judgement of potential with reference to some combinations of metals. It may be ideal to use the same material or that of approximation in potential for the above three parts. Where impossible, then the most suitable material will be selected for the heat transfer tubes, and the tube plates and the water chamber will be applied with cathodic protection.

5.4.3. Steel Vessels

Materials for the vessel are selected with engineering conditions, corrosion resistance of the material and technical level for operation taken thoroughly into consideration. Steel plate is preferable as the most popular vessel material for its cheap cost, easy manufacture and high strength; where operation at the high technical level is possible under minutely-examined engineering conditions, it will be satisfactorily serviceable for many years by making suitable allowance for corrosion. The water may cost, in the majority of cases, less than employing other methods of corrosion prevention, too.

However, unless the plant is operated at brine pH and DO levels which are designated in the specifications, corrosion on the steel plate will sharply increase, thus shortening the plant life. There may be a case where brine pH is low and DO is high according to the engineering conditions, and, if so, corrosion on the steel plate becomes severe.

For example, Mr. Togano and some others report, in their research, that corrosion on steel in a 6% NaC solution at 120°C varies as shown in Fig. 5.4-1 according to brine pH and DO level. And OSW report points out the corrosion rate on steel in deaerated seawater as shown in Fig. 5.4-12, and thus similar behavior is expected in brine.

As described, corrosion on the steel plate varies in quantity according to engineering conditions and the technical level for operation, and where corrosion is expected to exceed the allowable limit, proper measures for corrosion prevention will have to be taken into consideration. The portions of the interior of the vessel for which corrosion prevention measures are required is where brine is in contact, and particularly portions with which brine collides are important. In case concentraion of DO and carbon dioxide in the brine are high, even such portions as are in contact only with vapor will allow them to migrate to the vapor side to accelerate corrosion, and hence proper measures for corrosion prevention must be taken, too. However, it is general to apply corrosion prevention to the portions lower than the demister, i.e. the overall range with which brine comes in contact. In case, however, several portions only are affected concentrically by erosion - corrosion, or only the welded zones are affected, the corrosion prevention measures may be taken locally.

As will be apparent from researches by Mr. Togano and others and from the OSW report, the necessity of corrosion prevention measures depends on temperature, pH, DO level, flow rate, etc. of the brine in contact with the vessel. For exmple, corrosion prevention measures must be taken in such portions as are low in pH but high in DO level regardless of temperature and rate of flow. Where both pH and DO level are high or low on the other hand, the necessity will be determined according to each degree and of brine temperature and flow rate. Then, there may be a case where corrosion prevention measures must be taken in such portions as are high in temperature or in rate of flow regardless of being high in pH and low in DO level.

Furthermore, a case may happen where corrosion prevention measures are unavoidable on one hand, but no necessary on the other hand, according to properties of the raw seawater even under the same conditions. Thus, as much information will be necessary as possible before decisions on taking corrosion prevention measures. However, the circumstances are such at present that the type, place, intensity, etc. of corrosion to arise cannot be anticipated thoroughly beforehand, and experience is still important for the decision in most cases, thus making it difficult to find suitable corrosion prevention measures. To design a plant with corrosion prevention measures perfect at every phases is extraordinarily high in cost. To try to lower the cost is naturally to increase the danger of corrosion. The designer must therefore plan everything so that the water cost can be kept lowest by balancing the two factors above.

The corrosion prevention method which is popular for application on steel vessels is cladding and lining with corrosion resistant metal. or lining with nonmetallic materials like organic resin. A method by inhibitor, cathodic protection, etc. is not employed at present. The reason is that inhibitors, cathodic protection, etc. are expected to have a high running cost and have no experience as yet.

Lining with stainless steels like SUS304, 316, etc. as a corrosion resistant metallic material is most popular. However, there is a case where lining with 90;10Cu-Ni or 90/10Cu-Ni clad steel is used. These are used, in most cases, where the brine temperature works at more than 100°C. Of course, they can be used on lower temperature zones, but cases where they have actually been used are not many so far, because the corrosion rate of steel is retarded at the low temperature zone as compared with the high temperature zone, and non-metal lining which is very low in cost can be employed at temperatures below a certain degree. Stainless steel is acceptable generally as a material having good corrosion resistance to seawater at a high flow rate but is capable of allowing pitting corrosion where brine at high temperature and low pH comes strongly in collosion with it. A buffer plate would have to be arranged, in this case, to absorb shocks. And pitting corrosion occurs easily when shutdown of the plant from brine occurs. Drainage must be carried out thoroughly after shutdown of the plant.

Such a problem will not arise in the case of 90/10Cu-Ni. However, since materials of the copper group are generally weak against ammonia, care should be taken where ammonia is contained in seawater.

FRP, epoxy resin, phenol resin, chlorinated rubber, neoprene, etc. are used as organic resins. There are resins fully resistant to high temperatures. However, bonding strength of resin and steel plate becomes problematical as temperature rises high, therefore it is preferable to avoid using it at high temperatures. However, organic resin is appreciably cheap in comparison with corrosion-resistant metal, and for this reason development is expected in the future with regard to its lining and coating method.

As an alternative interior lining material for the steel vessel, materials of the iorganic group including, for example, mortar lining which will be described in the next chapter, are regarded as hopeful recently.

5.4.4. Piping

(1) Seawater line

From the viewpoint of corrosion, the seawater line consists of the following three systems:

- The seawater line covering the outlet of the seawater intake pump to the inlet of the heat rejection section and the line covering the outlet of the heat rejection section to the sulfuric acid mixer (raw seawater line)
- The line covering the outlet of the sulfuric acid mixer to the decarbonation tower and the outlet of the decarbonation tower to the deaeration tower (low pH line)

- The line covering the outlet of the deaeration tower to the final stage (deaerated seawater line)

The above way of classification applies to the case of a plant to work according to the pH control process by acid infusion, and the low pH line is to be omitted in the case of the chemical dosing process by infusion of a descaling agent.

Since raw seawater flows straight into the raw seawater line, steel pipes of STPG, STPY, etc. cannot be free from corrosion. However, the corrosion rate is not so large due to the temperature being comparatively low (below 40°C generally), and hence bare steel tube can be used by taking corrosion allowances properly and carrying out correct maintenance. Generally, however, corrosion product of steel are capable of choking the spray nozzle fitted on the decarbonation tower or deaeration tower. Therefore a suitable method of preventing corrosion will have to be used. It is then preferable to employ cheap organic resin coating like tar epoxy or epoxy resin for low temperatures. However, these resin coatings are weak in eddy currents, so care should be taken in use. As a corrosion preventive measure which is more trustworthy, the use of chlorinated rubber or polyethylene resin for lining, or mortar steel pipe will be considered. The mortar steel pipe being thick-walled, a means to have a short pipe becomes necessary for mounting the butterfly valve. To ensure safety furthermore, corrosion resistant metal like stainless steel or 90/10Cu-Ni can be used.

Next, the pH value of the seawater in the low pH line normally is 4 - 5 from the sulfuric acid mixer to the decarbonation tower and 5.5 - 6.5 from the outlet of the decarbonation tower to the deaeration tower. It is therefore wrong to use bare steel pipe in the low pH environment, and some suitable method for corrosion prevention must be considered. It is dangerous to use materials of the copper group like 90/10Cu-Ni because of low pH seawater containing air. Care should be taken in using stainless steel in the said environment, as stability of the passive state film deteriorates to crevice corrosion, stress corrosion cracking or pitting corrosion on shutdown of the plant. For prevention of the said corrosion attack, in the case of crevice corrosion, for instance, it is preferable to use the rubber group or Teflon group for packing, instead of asbestos, because crevice corrosion is grows easily, particularly on the flange surface. For prevention the stress corrosion cracking, it is necessary to prevent excess piping load and also to remove residual thermal stress thoroughly after welding. As for pitting corrosion, seawater in the piping must be discarded on shutdown of the plant and stainless steels of the extra low carbon group such as SUS304L, 316L, etc. must be selected so as to relieve considerably the detrimental influence of pitting and intergranular corrosion by welding.

A measure more popular for corrosion prevention is employment of organic resin lining. With the low temperature working, polyethylene resin will be satisfactory, which is cheaper than FRP, neoprene rubber, etc.

In the case of the deaerated seawater line, at a plant working through infusion of a descaling agent, seawater normally comes in pH 8 and moreover the seawater has already been deaerated and is low in temperature allowing the steel tube to remain as it is. In the case of a plant employing a pH control process by acid infusion, the pH value has recovered to 7 or so here, and hence steel pipe can be used satisfactorily by making corrosion allowance. In case, however, iron ions are to be kept from flowing into the vessel, the use of corrosion resistant metal or employment of organic resin lining must be considered.

(2) Brine line

The discharge brine line is omitted here for description, because corrosion on the piping does not exert any influence on its processes and further, the water chamber connecting pipe is treated in the same way as the water chamber and the brine connecting pipe is also subjected to the same treatment as the vessel. For the brine line, therefore, the following two must be taken up for consideration:

- The line from the final stage to the inlet of the heat recovery section by way of the brine circulating pump (low-temperature line)
- The line from the heat recovery section to the evaporation chamber on the first stage by way of the brine heater (high-temperature line)

Since the brine temperature working in the interior of the low-temperature line is normally below 40°C, even steel pipe like STPG or STPY will be serviceable so far as life is concerned. However, brine flows directly into the heat transfer tubes in the heat recovery section from this line, and hence to minimize deterioration in performance of the heat recovery section due to corrosion products of iron in the heat transfer tubes, the line will have to be applied preferably with a lining of corrosion resistant metal or organic resin. Then, a copper alloy such as naval brass is used for the tube plate, which is capable of giving rise to galvanic corrosion on the steel pipe side, and a corrosion preventive measure for the pipe will have to be taken accordingly. For prevention it is not necessary to use corrosion resistant metal like stainless steel or 90/10Cu-Ni but a lining with FRP, neoprene rubber or cheaper polyethylene resin is enough.

Steel pipe left as it is will invite severe corrosion in the high-temperature line, and hence a suitable measure for corrosion prevention must be taken. For prevention, piping of corrosion resistant metal or mortar are employed. The line being high in temperature, resin lining must not be used. Stainless steel may involve the problem of corrosion at welded zones subjected to a thermal influence where the brine temperature is high. Therefore, it is necessary to use SUS316L or so at least, for example, where the brine temperature exceeds 100°C. For safety, moreover, 90/10Cu-Ni or its lining tube will have to be used, too. In the case of plants where the maximum temperature is appreciably low, stainless steel 304 and 316 are preferably used.

(3) Distilled water line (Condensate line)

Since this line allows deaerated fresh water or that with less impurities having a conductivity less than 10 μ℧/cm to flow, a steel pipe like STPG will normally do, even at high temperature. However, in case water of high purity is required where distilled water is intended for the boiler, stainless steel is preferable. Then, the corrosion rate will be accelerated after fresh water is exposed to the air in the fresh water tank or when air leaks into it from the pump, and in such a case a corrosion preventive measure will become necessary by using stainless steel or applying resin coating.

(4) Chemical agent infusion line

A steel pipe can be used as it is for the concentrated sulfuric acid line, caustic soda line, sulfurous acid line, etc. The corrosion rate of the antifoaming agent line is high, because oxygen concentration rises from the agent solution diluted with fresh water or seawater. The chemical agent infusion line is small in aperture, and where corrosion grows on the piping, the corrosion product is capable of choking up the pipe, therefore 316 stainless steel, for example, is used for the antifoaming line in most cases. The material for the descaling agent infusion line in a plant employing the chemical dosing process varies according to the kind of descaling agent to be used. However, in case iron ions interfere to deteriorate the performance of the descaling agent or the corrosion product is capable of choking the pipe, use of stainless steel will be preferable. Resin piping such as vinyl chloride and polypropylene can also be used for the antifoaming line and the descaling agent line instead of stainless steel.

5.4.5. Pump

(1) Corrosion environment of pumps

The large-sized pump used in multistage flash-type desalination plants principally handles the following fluids:

Raw seawater, deaerated and decarbonated seawater, brine and distilled water.

Raw seawater: Marine pollution has become severe recently, and the corrosion environment metallic materials becomes worse and worse due to ammonia and hydrogen sulfide dissolved in the seawater.

Seawater deaerated and decarbonated: The corrosion environment is improved through deaeration and extraction of corrosive gas by this treatment.

Brine deaerated and decarbonated: Concentration of chlorine ions of circulating brine is 33,000 - 38,000 ppm or so. However, the corrosion environment can be said to be rather preferable as compared with raw seawater because the oxygen is less dissolved. Where dissolved oxygen is constant in quantity, the corrosion rate of metal rises in accordance with temperature rise. However, liquid temperature of the pump for circulating brine is actually below 40°C.

As described above, the corrosion environment for the pump is raw seawater and deaerated brine which is somewhat better than raw seawater. However, it must be noted that air may leak into the pump from outside at shutdown. Therefore, seawater resisting materials will be selected for the pump.

(2) Materials used for pump

Materials used against seawater cover various kinds from popular and cheap materials like cast iron to that of high cost which contain much nickel, such as Monel metal and Hastelloy.

The following is to describe each material:

(a) Plain cast iron; spheroidal graphite cast iron

Fig. 5.4-13 and Fig. 5.4-14 show corrosion test results on cast iron, of each kind, which was carried out in seawater at Takeyama Shore, Kanagawa Pref. As will be apparent, corrosion loss sharply advances at more than 20 m/s in the flow rate for FC20 and at more than 15 m/s for FCD45. Thus FCD45 is stronger at less than 15 m/s, but FC20 becomes stronger, at high-speed. This may be due to the graphitized corrosion layer of spheroidal graphite in FCD45 being inferior in strength to the corrosion product layer (graphitized layer and rust layer) of FC20.

The suction bell, swing pipe, discharge casing, etc. of the vertical shaft pump are usually exposed to a flow rate less than 15 m/s, and hence there is a case where these cast irons are used for seawater not subjected to pollution in consideration of the corrosion allowance. For the desalination plant, however, these are not used at present. At the desalination module test plant in Oita City, in consideration of the short duration of testing, plain cast iron was used on the low flow rate zone of the vertical shaft pump casing in the seawater intake pump, and no corrosion involving a problem took place because the seawater was clean.

(b) Low chromium cast iron

The addition of chromium to cast iron makes the rust layer on the surface of cast iron compact and also gives better adhesion. For example, cast iron containing a little over 1% of chromium can be obtained cheaply and is more corrosion resistant than plain cast iron. As will be apparent from Fig. 5.4-13, low chromium cast iron of spheriodal graphite structure has good corrosion resistance in a flow rate of seawater more than 15 m/s as compared with other cast iron.

(c) High nickel cast iron

It is well known generally as Niresist cast iron and has already been standardized in ASTMA436. With 13.5 - 36% nickel and a minute amount of chromium and copper contained, this high nickel cast iron is specified as TYPE I to V in ASTM. In metallographic structure this cast iron is that for which graphite in pieces and 5 - 20% in volume disperses in an austenite matrix.

In seawater where rusting goes on slightly, there is no serious problem. As in the case of plain cast iron, moreover, the type of corrosion is general corrosion and not pitting corrosion or crevice corrosion, thus serving well enough.

As illustrated in Fig. 5.4-15, it can be found that the results obtained through corrosion tests on Niresist cast iron of the three types gives a variant rate of flow as the corrosion rate becomes high. That corrosion decreases at high-speed is due to passivation going on, and this kind of material is expected to be resistant even in such parts with severe flowing conditions. Despite the material cost being very high, the material is used entirely for the casing and say nothing of high flow rate part of the pump casing (guide casing for the vertical shaft pump, spiral casing for the horizontal shaft pump).

The Spheroidal graphite group (Niresist ductile cast iron, ASTM 439-71) is possessed of a strength about double that of the flaky graphite group and is stiff in addition, and hence it is useful to cope strength requirements.

(d) Stainless steel of the austenite group

This is a material fairly superior in corrosion resistance against flowing seawater but gives rise to local corrosion on portions slow in the flow rate, therefore care should be taken on the portions where seawater is stagnant in the pump casing or faucet joint.

However, the pump impeller or the shaft running at high speed is conditioned favorably in this respect, and it is employed for almost all seawater pumps at present with a satisfactory result in durability.

An environment where the temperature of seawater exceeds about 80°C is capable of causing stress corrosion cracking and in this case, working stress must be designed accordingly low. SUS304 and 316 steel are predominant as materials. However, there may be a case where the costly Carpenter 20 (20Cr-30Ni-2Mo-Cu) with high crevice corrosion resistance is used.

(e) Other stainless steels

2-phase (ferrite and austenite) stainless steel and high chromium low carbon stainless steel have been developed as new stainless steel materials which are trustworthy in resistance against local corrosion (pitting corrosion, crevice corrosion, stress corrosion) in seawater. For example, SUS329J1 (25Cr-5Ni-2Mo) for the former and SHOMAC30-2 (30Cr-2Mo-0.003C-0.008N) for the latter can be used. However, they are very expensive materials, and use on the pump is extremely limited.

(f) Higher nickel alloy

Materials of superior corrosion resistance against seawater are Monel metal, Hastelloy C, etc.

Monel metal: Monel is a nickel alloy containing copper at 30 - 37%, and its passivating characteristic based on high nickel content indicates an extremely superior corrosion resistance against the air-saturated high-speed seawater. On the other hand, however, pitting corrosion in static seawater occurs easily, and care should be taken. Monel metal is a costly material and for this reason, it is used on impellers or shafts only in special cases.

Hastelloy C: This is corrosion resistant material of the Ni-Cr-Mo-Fe-W group which contains nickel at about 70% and is not only stable against the seawater but also high enough in strength. But its high cost prevents use on the seawater pump in most cases.

(3) Cathodic protection for the pump

As described, corrosion resistance is usually handled for each component of the pump by means of seawater resisting materials. However, it is not economical, in the case of pumps of large capacity, to manufacture the large casing with costly seawater resisting materials. For an alternative method, cathodic protection is available for manufacturing the casing with cheap plain cast iron. Since a pump for desalination of seawater generally works on a large capacity, protective current must be kept somewhat large. Therefore an external power supply system is usually employed for cathodic protection.

It goes without saying that care should be taken of applying cathodic protection on the pump so as not to deteriorate performance of the pump; perfection in corrosion prevention must be secured through taking care of areas to mount electrodes, protective current density, electrode materials, wiring, etc.

(a) Areas to mount electrodes

A design for areas to mount electrodes and also for the number of them is determined so that the necessary protective current density may be obtained to lower the potential of the object to standard protection potential. An area where the flow rate varies partially in the pump is capable of producing a local cell. Where foreign matter sticks to the surface of stainless steel in seawater, unevenness will occur in supplying oxygen, thus giving rise to pitting corrosion or crevice corrosion from activation of such parts as are less supplied with oxygen. It is necessary to give consideration so that corrosion protective current will reach effectively to such parts.

The current value required in seawater varies according to the temperature and deaeration of the seawater, flow rate and metallic materials. However, the effect will be obtained usually at 0.1 - 1 A/m^2.

(b) Electrode material

What is preferable with electrode (anode) materials is that current density is large, shape is obtainable freely and consumption is minimized.

For use on the pump, wear resistance and shock resistance will be the most important factors, as the electrode is exposed in the stream.

In consideration of what has been described above, high silicon cast iron and that containing chromium at 4 - 5% will be optimal for the seawater pump under existing circumstances.

(c) Wiring

Wire doubly covered with polyethylene and vinyl chloride is used for the wiring connecting output terminals of the DC power unit and each electrode mounted on the pump casing. The wire is enclosed in a protective tube so as to be safe from being damaged underwater.

Where an electrical contact between the rotating body and the casing is insulated, corrosion protective current will be prevented from flowing fully into the rotating body, therefore the top coupling is provided with a carbon brush to keep the electrical contact.

(d) Embodiment

Figs. 5.4-16 and 5.4-17 represent a brine circulating pump delivered to the seawater desalination test module plant of the Industrial Science & Technology Agency at Oita and its cathodic protection device.

The suction bell, guide casing and discharge casing use plain cast iron, the impeller uses austenite stainless steel (SCS13), the shaft uses austenite stainless steel (SUS304), and the barrel is manufactured with mild steel plate (SS41).

High silicon cast iron is used for the electrodes. Electrodes mounted on both inside and outside of the casing are 14 pairs all told.

Corrosion protective current was supplied at 1.0 A/m^2 inside the pump with a high flow rate and 0.5 A/m^2 to cover the interval with the barrel outside the pump with a low flow rate.

After operation for about 3,400 hours at 41°C and for about 3,400 hours at 91°C, the pump was drawn up to check for corrosion. It was then found that the corrosion prevention was perfect and almost no corrosion arose except a slight electrocoating (precipitate comprised mainly of calcium) on the suction bell, etc. Fig. 5.4-18 is a photo showing electrodes before and after use.

5.5. Concrete Evaporator Shell and Mortar Lining

5.5.1. Development of Concrete Evaporator Shell

(1) U.S.A.

To cope with the shortage of water to be expected in the future, development of a seawater desalination technique has been strongly taken up in the U.S.A. early, and researches on seawater desalination have been carried out in every field concerned through having established the Office of Saline Water (OSW) in the Department of the Interior in 1952. Particularly in the 1960's, many personnel and much research expense were invested as a national project in parallel with space development, and the national budget expended for it up to 1973 had in fact reached 275 million dollars.

As for development of the concrete evaporator shell, its economical advantage was perceived early and a series of studies has been conducted since 1964 with the technological cooperation of the United States Bureau of Reclamation (USBR), the Oak Ridge National Laboratory (ORNL), the Brookhaven National Laboratory (BNL), etc., where the Office of Saline Water and the Atomic Energy Commission assumed a leading position in aiming at decreasing the cost of desalination.

Various tests were carried out, first of all, for basic properties of concrete materials at the USBR, including a test on the durability of concrete against hot seawater and hot distilled water, a test on change in properties of concrete such as strength, modulus of elasticity, creep, etc. under high temperature conditions, and tests on air-tightness of the shell using a concrete model and on temperature distribution in the concrete wall; those tests have resulted in clarifying that concrete material is satisfactorily serviceable as a material for the shell. Particularly in a loop test on durability of concrete which was carried out for long periods by changing the mix proportions of concrete and kinds of cement, it has been concluded that concrete with proper mix proportions is durable enough against high temperature brine, while the portion directly exposed to the flow of high temperature distilled water should be coated on the surface against the possibility of being eroded.

In parallel with a series of those basic studies, the OSW carried out a conceptional design of a seawater desalination plant of 50 MGD (about 200,000 m^3/day) in 1965. The conceptional design was realized by trusting private enterprises with certain specifications. 8 companies, including Aqua-chem, Badger and Lockheed participated. These three companies submitted conceptional design of the plant using concrete shells. The concrete shells adopted were variant in sectional shape. They were rectangular, circular and elliptical according to the company, and the structural types of the shells were reinforced concrete, prestressed concrete or a combination of the two. The designs were to characterize each strong point, and the feasibility of the concrete shell was demonstrated. What is to be emphasized here is that, in every case, the construction expenses were minimized and the desalination cost could be sharply decreased by utilizing the concrete shell.

Furthermore, a module shell had been constructed in reinforced concrete at the San Diego Test Facility of the OSW in parallel with indoor tests since 1967. The shell was 3 m in height, 2.4 m in width and 6 m in length with much reinforcement (Photo 5.5-1), and was intended for obtaining data on deformation and stress arising in the shell at various temperatures and pressures, and on long-term changes in properties of the shell, by supplying vapor and hot seawater from adjacent test plants. However, the test was in trouble in continuing further studies due to delay in execution of the construction work and operation of the plant at an early stage, thus ending in an unsatisfactory result.

On the other hand, the development of resin concrete can be taken up as one of the remarkable achievements of the studies of concrete shells in the U.S.A. The research development of resin concrete was commended in 1965 through joint operation by the Brookhaven National Laboratory and the Bureau of Reclamation under the supervision of the Atomic Energy Commission and the OSW joined in the cooperative study and showed that the material had superior performance for use on the desalination equipment. The resin concrete includes polymer concrete (PC) obtainable through mixing aggregate and resin instead of cement and polymer impregnated concrete (PIC) obtainable through impregnating hardened conventional concrete with resin in its pores. Since PIC was found to be superior to PC in strength and durability through study at the beginning, PIC was mainly used for the series of tests. The conclusion was obtained that the wall thickness could be decreased to 1/3 - 1/5 by using PIC for the concrete shell, compared with conventional concrete. However, PIC still leaves problematical points in construction and its economy, thus keeping its practical use from realization at present.

Meanwhile, the OSW has carried out a study on utilization of cement mortar for the inner surface lining of steel shells, noticing the superior durability of concrete material. As for mortar lining, a long-term performance test has been carried out at the test plant for desalination at the Fontainvalley Test Facilities.

As described above, a full-scale study for development of the concrete shell had been carried out in the U.S.A. during the period 1964 through 1971, which led to the conclusion that the concrete material could be satisfactorily utilized for evaporator shells. Since then, however, the scale of the study on concrete shells has been reduced due to an overall change in the R & D project on seawater desalination, and the succeeding study with a module plant has not been carried out.

(2) France

Development of research on seawater desalination equipment was commenced by the Commissariat L'energie Atomique (CEA) in 1966 and studies have been carried out at the seawater desalination test facility of the Toulon Naval Arsenal. As for concrete shells, basic study on concrete materials and study with a pilot plant were carried out by the CEA through cooperation with manufacturers concerned. Fundamental tests were carried out on durability of concrete with various kinds of cement and aggregate, or thermal conduction in concrete members and on air-tightness of concrete. Construction and operation of the pilot plant using concrete materials was commenced in 1973; the concrete shell used there was a prestressed concrete structure with a rectangular section, the inner size of which was 3.6 m high, 6 m wide and 12 m long. The shell was partitioned into 3 stages internally and the wall was 40 cm thick. A seawater resistant cement was used according to the French Standard, the aggregate used silica, and 80 kg/cm^2 prestress was employed for the shell wall. Externals of the pilot plant are shown in Photo 5.5-2; the pilot plant was put in operation for 18 months covering November 1973 through May 1975 with temperature conditions changed from 40°C to 120°C and then in continuous operation at 120°C during the period May to June 1976. As the result of a minute observation on changes in properties of the inner surface of the concrete shell after termination of those operations, such changes in properties as would exert a detrimental influence on durability or air-tightness of the shell were not detected at all, and the concrete shell was evaluated as extremely durable. A series of these tests were over in 1976, and then the test has proceeded with examination of the economy of the plant using a concrete shell in view of the positive results of these tests.

(3) West Germany

Research on development of a concrete shell for seawater desalination equipment was commenced in West Germany from 1968, and particularly since 1973 a full-scale study has proceeded as an operation within the structure of COST53 of the European Community through cooperation with Spain and Yugoslavia and with funds supplied by the German Ministry of Development. As for the allotment of work among the three countries, West Germany takes charge of the process engineering of the equipment and design and construction of the shell, Spain takes charge of the study of chemical properties of concrete material and Yugoslavia the study of physical properties of concrete and a miniature model test; total amount of the research fund is about 5 million Deutsche Marks (about 6 hundred million Yen).

With reference to the study by West Germany, a study of materials in part, a trial design of the vertical evaporation equipment and a structural analysis of various loads are being carried out by the Technische Universität Berlin through cooperation with Krupp, Düsseldorf Cement Laboratory and Battel Laboratory (Frankfurt). Particulars of this research are as follows:

(a) Strength characteristics of concrete

(b) Influence exerted on modulus of elasticity of concrete by temperature and water content

(c) Influence of water-cement ratio, type of cement, temperature, humidity, etc. on creep of concrete

(d) Drying shrinkage of concrete

(e) Influence of temperature and humidity on heat conductivity of concrete

These researches are prearranged to go on until 1978 and pending upon the result, manufacture of a practical plant using a concrete shell with 800 m^3/day capacity is expected at Helgoland in 1978.

(4) Japan

Research for development of a concrete shell in Japan were commenced simultaneously with the study on seawater desalination by the Agency of Industrial Science & Technology (AIST), the Ministry of International Trade & Industry. That is, the AIST took up "Seawater Desalination and By-Product Recovery" as a research theme for the development of large industrial technology (Big Project), which was effective from 1969. And as one of the research items, development of a concrete shell considered as important for the reduction of desalination costs was adopted. The research was pushed forward by the National Chemical Laboratory for Industry, Tokyo (NCL) of the AIST as the central figure under a research system through the joint efforts of government and people. Progress of the research is as shown in Table 5.5-1.

First of all, a basic study on concrete materials had been commenced in 1969 and in the same year, "Concrete Shell Material Committee" was established within the Society of Seawater Science, Japan, entrusted by the AIST. Feasibility of materialization of a concrete shell was made possible through examination of test results obtained till then by the Office of Saline Water, U.S. Department of the Interior, fundamental tests on material properties including resistance of concrete against hot seawater and resistance against hot water of various kinds of lining materials, foundamental tests on thermal stress of the concrete structure and extensive feasibility covering conceptional design on concrete shells for a 100,000 m^3/day seawater desalination plant. Based upon the results obtained through those tests, durability tests on concrete materials against hot seawater and hot distilled water, tests for changes in mechanical and thermal properties of concrete under high temperature conditions, structural tests on thermal stress by means of a concrete shell model and tests on air-tightness of concrete shells, etc. were carried out at the Marine Test Facilities of the NCL, Chigasaki (City), Kanagawa (Pref.) from 1970. Fundamental properties of concrete shells under various conditions could then be clarified.

In parallel with these tests, "Study with a Concrete Evaporator Shell Module" was entrusted to private laboratories since 1971 as a link in the chain of the Big Project. The Study was aimed at finding hot the concrete shell would behave under the operating conditions of an existing desalination plant. Firstly in 1971, a test module plant using a prestressed concrete shell (PC shell) which was taken to be most reliable in air-tightness, etc. was constructed and put in operation. Then in 1972, a reinforced concrete shell module (RC shell), taken to be more economical than the prestressed concrete shell, was constructed and operated, which led to the conclusion that reinforced concrete was serviceable enough as an evaporator shell. A series of those studies brought a lot of valuable data on the design and construction of a practical evaporator shell, and on behavior of a concrete shell under operating conditions of seawater desalination plants, thus stimulating materialization of the practical concrete shell. Photo 5.5-3 shows a bird's eye view of the concrete shell module plant.

On the other hand, "Study with 100,000 m^3/day Test Module" had been planned since 1972, including construction and operation of a demonstration plant in which the results of engineering researches on the multistage flash evaporation process and material tests concentrically carried out at the Marine Test Facilities of the NCL till then were incorporated. A basic design of a 100,000 m^3/day seawater desalination plant was decided on in 1972 for the first stage of the Study, and a concrete shell was employed for the evaporator shell in the basic design. Design and construction of the Test Module Plant were commenced in 1973, and a concrete shell was employed as the heat rejection section.

A bird's eye view of the concrete shell for the heat rejection section is as shown in Photo 5.5-4, which will be particularlized later. The Test Module Plant was kept running during the period September 1975 to March 1977 for about 6,600 hours (275 days), and an operation study was carried out under various temperature conditions. Upon termination of the whole operation, close observation was performed on the inner surface of the shell and, at the same time, the concrete was subjected to various analytical tests: any changes in properties to exert a detrimental influence on durability of the shell were not, and it was ensured that concrete was durable enough to work under operating conditions of desalination plants. It was proven through the study that the concrete shell could satisfactorily cope with requirements for performance of the evaporator shell for seawater desalination plants and at the same time, structural behavior of the concrete shell under operating conditions and a method of how to operate the practical plant using a concrete shell could be clarified. Data available for design, construction and operation of a practical concrete shell could thus be obtained.

As described above, Japan has succeeded in the development of concrete shells leading internationally and is now at the stage of how to put them in service.

5.5.2. Construction of Concrete Evaporator Shells

(1) Planning of Concrete Evaporator Shells

The concrete shell has proven to have excellent performance not only in economy but also in durability in comparison with a steel one. However, since concrete is a material having different properties from steel, its properties must be well understood and every requirement for having its qualities thoroughly understood must be incorporated in the planning for utilization.

The concrete shell comprises the shell body and its foundations, and particularly the conditions to which the shell body is subjected are rather special with respect to the desalination plant, and hence several points for attention that are different from those for conventional structures are to be pointed out for manufacture. That is, the interior of the shell is exposed to various conditions such as a maximum temperature of 120°C, internal pressure of 0.04 - 2 kg/cm^2 (absolute pressure), flashing flow of the concentrated brine and, furthermore, dropping of the condensate on the upper surface of the inner wall during operation, and thus the concrete shell must be resistant to these conditions. Requirements for the concrete shell are pointed out as follow:

(a) Must be thoroughly durable against high temperature brine and condensate at 120°C.

(b) Must be thoroughly safe structurally against thermal stress and pressure working on the shell.

(c) Must be such a structure as will function satisfactorily as a part of the desalination equipment.

The following is a fundamental conception of sectional shape and structural type for planning such a concrete shell as will satisfy those conditions fully:

Various sectional shapes can be considered for the concrete shell, however, the typical shape may be a rectangle or circle and, as an intermediate, an ellipse or semicircle. Each shape has merits and demerits: the circular section, for example, is structurally advantageous against loads like pressure, thermal stress, etc. compared with the rectangular section, but causes difficulty in construction work and a high construction cost. Therefore, the shape to be employed depends on which is a priority among structural characteristics, easiness of construction and economy at the time of planning the shell.

Prestressed concrete construction and reinforced concrete construction is used for the type of structure, each having features as described later: the prestressed concrete structure is suitable for the high temperature stage module in which thermal stress is large, and for the module coming in a lower temperature range than that, reinforced concrete structure which is superior in economy is suitable.

As described above, in planning concrete shells, sectional shape and structural type must be selected through grasping working conditions of the shell and also in consideration of easiness of construction work and economy.

Given next is a description of the shell foundation. The shell foundation is a structure to transmit loads of the concrete shell to the ground and also to keep the shell stable; once construction of the shell is over, the foundation cannot be repaired in most cases, and hence planning based on a great store of technical know-how and experience will be required for design and construction of the foundation.

Items required for the shell foundation are as follows:

(a) The shell body must be kept positioned correctly during emergency and normal times.

(b) Such a structure as will restrain a volume change of the shell body according to its temperature change must be avoided.

(c) The structure must be such that loads working on the shell foundation can be directly transmitted to the ground.

To design and construct the shell foundation satisfying the above conditions, the optimum foundation will have to be planned through various investigations to ensure properties of the ground and study of the foundation of an existing structure in the neighborhood of the prearranged area for construction of the plant, thereby reviewing structural and economical matters.

(2) Prestressed Concrete Evaporator Shell

(a) Features of the prestressed concrete structure

For its many merits, concrete material has been used extensively as a construction material, however, what is worst is that resistibility is limited against tensile force.

As described above, high air-tightness is required for evaporator shells of desalination plants, and at the same time, the shell is itself subject to extremely severe conditions. The structural type that is regarded as most trustworthy against these conditions is that of prestressed concrete structure. Prestressed concrete is designed so as to eliminate tensile stress arising on a concrete member subjected to various loads, by giving compressive force beforehand with steel materials of high strength. Therefore, the prestressed concrete structure allows no cracking to develop because it is put under compressive stress at all times. Prestressed concrete is classified into concrete designed so that the resultant stress by the introduced prestressed stress and that due to the design load will not work as tensile stress (full-prestressed) and that of being designed so that the resultant stress will not exceed the allowable tensile stress (partial-prestressed).

The prestressed concrete structure is most reliable. However, it is disadvantageous economically in comparison with the reinforced concrete structure described later, and hence it is taken to be best that the shell be used for the high temperature stage module putting the shell under the severest conditions.

(b) Design and construction of prestressed concrete shells

1) Materials to use

The prestressed concrete consists of concrete, prestressing steel bars and reinforcement and is required to have high strength.

Concrete generally decreases in compressive strength and modulus of elasticity as temperature rises, and the ratio of decrease varies according to temperature conditions under which the concrete works. Therefore, it is necessary to plot out concrete with its strength and modulus of elasticity decreased according to the working temperature conditions.

The tendency of drying shrinkage of concrete and creep by its plastic property also increases in accordance as temperature rises, which is an important factor for determination of how much to prestress, particularly in the prestressed structure and will exert a big influence on economy of the shell. For design of the shell, therefore, these properties are to be taken thoroughly into consideration. Also, prestressing steel bars enlarge under high temperature conditions, which must thoroughly be taken into consideration for use.

2) Thermal stress

Pressure, temperature, etc. as loads, will be taken into consideration for design of the concrete shells. However, thermal stress is the most important of all. Since flash evaporation is kept going under high temperature conditions in the interior of the shell but the exterior is in contact with the open air, there arises a difference in temperature between the inside and outside of the shell wall, and thus a thermal stress is induced. Thermal stress is an important factor for design and construction of the shell and at the same time, cracks occurring in the concrete shell due to thermal stress may decrease air-tightness, which is an important function of the shell or seawater coming through the cracks may work to corrode the steel material within the concrete or decrease the durability of it.

The prestressed concrete shell is so arranged as to cope with thermal stress by introducing compressive prestress, and minimum thickness of the concrete member is determined according to requirements for construction works or design against pressure load. The thickness of the shell wall may bring about a large thermal stress at the high temperature stage module, thus making thorough prestressing difficult. Therefore, the arrangement is such that the outer surface of the shell is covered beforehand with a heat insulating material to limit temperature difference in the members within a specified value. It is then important and economical as well to introduce prestress matching the limited temperature difference.

Now, an optimal sectional thickness will have to be determined when plotting out thermal force by calculating temperature distribution arising on the concrete member beforehand and examining heat insulating material thickness and prestressing value synthetically. Then, the matter of which to take, "full-prestressed" or "partial-prestressed", for thermal stress is to be determined according to the state wherein the shell is placed and the method of operation, thereby ensuring the safety and function of the shell to the fullest extent.

3) Air-tightness

High air-tightness is required for the shell so as to keep the interior of shell working at less than 100°C below atmospheric pressure during operation.

It is known through tests that air-tightness of concrete is subject to the influence of mix proportions of concrete, configuration or drying rate of the concrete member. As is popular for use in structures like radial shelters or large water tanks, concrete prepared compactly has proven experimentally to be a material having high air-tightness. Moreover, the prestressed concrete structure is essentially free from cracking, thus involving no problem regarding air-tightness. However, the construction joints of concrete shells and interfaces with metals embedded in the concrete are portions that may decrease air-tightness of the shell. Therefore, measures to prevent air leakage must be taken by carrying out construction joining carefully on these portions of the shell construction, and countermeasures will have to be considered at the time of design.

4) Construction practice of concrete shells

The concrete shell is fitted with various hardwares, for which high precision is required in mounting; special treatment for air-tightness is necessary on the construction joints or the interfaces of hardwares and concrete, thereby improving air-tightness of the shell. Furthermore, precision in arrangement of prestressing steel bars is required along with density, durability and high strength for the concrete itself, and hence a high standard of technique will be necessary for the execution of work on the shell as compared with general concrete work. Through investigation and examination carefully done beforehand, work on the shell must be executed with scrupulous care actually based on a close working plan.

Considerations pointed out as above are then applicable in nearly the same way to reinforced concrete shells described later.

(c) Designs

There is no case realized so far in which a prestressed concrete shell has been used in actual plants, however, a prestressed concrete shell of workable scale was once constructed for a test plant for various operations. The shell was constructed in the Marine Test Facilities of the NCL at

Chigasaki as a link of a large project by the Agency of Industrial Science & Technology mentioned before, and its schematic shape and dimensions are as given in Fig. 5.5-1 and Table 5.5-2 (see Photo 5.5-3). Since the shell was constructed for a test plant, mounting of machinery and appliances was minimized only to the flash evaporation in the function of the plant. A rectangular section was employed for the shell in consideration of economy and ease of construction. The sectional thickness was 50 cm in view of safety, easiness of construction work and air-tightness of the member against loads working on the shell. In careful examination of the load which would work on the shell, the prestressing quantity introduced was determined according to the fundamental conception described before. There was a slidable shoe mounted between the foundations and the shell so as not to restrain a change in volume caused by temperature rises on the shell, and a fixed support coupling the foundation and the body through an anchor bolt was properly arranged so as to resist lateral loads working during earthquakes, etc.

(3) Reinforced Concrete Evaporator Shells

(a) Features of reinforced concrete structure

What is a remarkable feature with a reinforced concrete shell is that it is extremely economical in comparison to a prestressed concrete shell, and there is a big difference in construction between the two. Namely, while the prestressed concrete is of such construction as will prevent cracks from occurring by having the concrete given compressive force beforehand, the reinforced concrete is designed so that concrete resists compressive force only and the reinforcement in the concrete resists against all tensile force. Therefore, the reinforced concrete is designed on the premise that cracks will arise in the area where tensile force is caused in the concrete section by the load working on the shell.

On the other hand, high air-tightness is required for the evaporator shell, and hence evaluation of the cracks will be problematical in plotting out the reinforced concrete shell. Actually, however, the load working on the shell used for desalination plants comes mostly from temperature and pressure to cause no penetrating cracks through the shell wall section, and moreover the concrete itself has an appreciable tensile strength, which does not always allow cracks directly from the working tensile force. The width, depth, etc. of cracks arising in concrete vary according to the dimensions of the working load, type and amount of the reinforcement, and dimensions of cracks expected to arise can be somewhat controlled at design.

(b) Design and execution of works on reinforced concrete shells

The conception of load conditions working on the reinforced concrete shell and design and construction thereof is fundamentally the same as in the case of prestressed concrete shells. However, the prestressed concrete shell is used for the high temperature stage. The reinforced concrete shell will be compared with the prestressed concrete shell as follows for considerations necessary in the design of the reinforced concrete shell:

1) Materials to use

Reinforced concrete comprises concrete and reinforcement; high strength is not so necessary for the concrete itself as in the case of prestressed concrete shells, however, it still must be sufficiently durable for use. Properties of the concrete at high temperature can be regarded in the same way as in the case of prestressed concrete.

2) Thermal stress

Thermal stress is also most important in the case of reinforced concrete shells. However, the conception of reinforced concrete shells for thermal stress varies from that of prestressed concrete shells. Thermal stress is that arising from changes in volume of the concrete member according to temperature changes being restrained by the concrete section or an adjacent member. It is, in other words, internal stress. Therefore, where cracks arise in a concrete member in the reinforced concrete structure, the restraint is released by movements of the member due to the crack, thus thermal stress decreases. In the case of reinforced concrete shells, a crack develops in the concrete member upon thermal stress reaching a certain amount, and thermal stress decreases by so much accordingly, therefore its value is regarded as coming below a given value at all times. It is therefore reasonable to plot out thermal stress on the reinforced concrete shells by taking the above-mentioned property into consideration.

3) Air-tightness

In the entire evaporator shells a module using a reinforced concrete structure is used in such portion as is comparatively low in temperature, and the fact that the shell is covered with heat insulating material externally is to minimize thermal stress arising in the shell, and to keep the crack width smaller. Tests done so far clarifying the above, it is not likely that cracks arising in the shell will exert particularly detrimental influence on air-tightness. Therefore, nothing will need to be done for air-tightness in the case of reinforced concrete shells by applying similar treatment to prestressed concrete shells for air-tightness.

(c) Designs

As an example for the design of reinforced concrete shells, those of "Study with Concrete Shell Module" and "Study with 100,000 m^3/day Test Module" which were constructed as testing equipment can be taken up. The latter shell for the 100,000 m^3/day module is that which was designed to have a scale working as practical equipment, and its sectional shape and dimensions and specifications are as shown in Fig. 5.5-2 and 5.5-3 (see Photo 5.5-4). To serve as a testing plant, the shell has been designed to resist load conditions for both the heat rejection section and the high temperature stage below 100°C, adjusting to operating conditions of the module plant. As in the case of prestressed concrete shells, the said shell is mounted with a given thickness of heat insulating material on its outer surface so as to minimize the temperature gradient coming out through the shell wall, and a slidable shoe is arranged between the foundation and the shell.

5.5.3. Characteristics of Concrete Shells

(1) Durability of Concrete Shells

Since the interior of the concrete shell is exposed to hot concentrated brine and condensate, the concrete should be durable against these conditions. As the result of having carried out a series of tests to examine the influence to be exerted on concrete by hot brine and condensate, a conclusion has been obtained that the concrete material is exceedingly durable against these conditions. The testing device is as shown in Photo 5.5-5.

A result obtained through testing mortar in hot seawater for change in the compressive strength ratio and the bending strength ratio conducted to see the durability of concrete against hot seawater is as illustrated in Fig. 5.5-3. The result shows that where the water-cement ratio of the concrete comes below 45%, a deterioration is almost not indicated in both compressive strength and bending strength, however, where the ratio comes over 55%, the concrete changes in property by substitution of Mg^{++} ion in the seawater for Ca^{++} ion. From the results above it has been made clear that a high quality of concrete with proper mix proportions is satisfactorily durable against hot seawater.

A test for corrosion resistance against hot distilled water has resulted as shown in Fig. 5.5-4. The amount of dissolved matter from the concrete varies according to the kind of materials and the temperature conditions. However, a relation between amount of dissolved matter and the leaching period is expressed by the following equation:

$$m = k_0 (1 - e^{-\alpha v}) t^n \quad \cdots\cdots\cdots\cdots\cdots\cdots \quad (1)$$

here, m: total amount of dissolved matter (kg)/
specimen surface area (m^2)

v: vapor condensation rate (ℓ)/
specimen surface area (m^2) • number of days (day)

t: leaching period (day)

From the relation above the corrosion depth of the concrete shell and time are correlated as given in the following equation:

$$\left.\begin{aligned} D &= \frac{k_0}{C} (1 - e^{-\alpha v}) t^n \\ V &= \frac{24 K \Delta T}{q_0 L} \end{aligned}\right\} \quad \cdots\cdots\cdots\cdots\cdots\cdots \quad (2)$$

here, D: leaching depth of concrete wall (m)

C: unit cement rate of concrete (kg/m^3)

K: heat conductivity of concrete (Kcal/m•hr•°C)

q^o: vapor condensation latent heat (Kcal/kg)

L: concrete wall thickness (m)

T: temperature difference between inside and outside concrete wall (°C)

n, α, k_o: coefficients given in Table 5.5-4

In case there is a difference in temperature of 15°C through the concrete wall 0.5 m thick for which ordinary Portland cement 400 kg/cm^3 in unit cement rate is used, the depth of dissolution loss for 15 years can be obtained at 0.95 mm from the above equation, leading to a conclusion that the same may scarcely involve a problem in view of practical equipment.

On the other hand, with reference to the durability of concrete shells, a result obtained through observing closely the change in properties on the inner surface of the shell upon closing all the operation tests for "Study with Concrete Shell Module" and "Study with 100,000 m^3/day Test Module" has ensured that the concrete shell is exceedingly superior in corrosion resistance.

According to the corrosion tests carried out by the U.S. Bureau of Reclamation on concrete in hot seawater (concentration ratio being 2), the denatured layer on the concrete surface gets deeper in accordance with the exposure period, and the thickness obtained after exposure for a long period according to the test data covering about 4 years comes in as shown in Fig. 5.5-5. It is then estimated that each becomes 10 mm and 20 mm at the brine temperatures 100°C and 120°C on an average, respectively.

(2) Mechanical Properties of Concrete Shells under High Temperature Conditions

(a) Change in physical properties of concrete at high temperature

It is ensured through various tests that concrete deteriorates in its mechanical properties, particularly in strength and modulus of elasticity under high temperature conditions, and thus creep increases. Increase in creep decreases the amount prestressed, thus exerting a considerable influence on economy of the shell.

An example of the test results on strength and modulus of elasticity of concrete at high temperature is as given in Fig. 5.5-6. As will be apparent from the result, compressive strength deteriorates as the temperature rises. However, the percentage of deterioration is 15% or so even at 100°C; the temperature conditions then influence the deterioration and the number of days to keep the temperature is not so serious in this case. A similar result has been obtained for modulus of elasticity (Fig. 5.5-7). On the other hand, test results on concrete creeping at high temperature are as shown in Fig. 5.5-8, from which it is found that concrete creeping is accelerated as the temperature rises but not so much up to 80°C and then goes up sharply beyond 100°C.

To design the concrete shell so as to cope with the requirements against the above concrete properties at high temperature, it is recommended to use the modulus of elasticity at a reduced value according to the service temperature conditions and adjust the creep coefficient to cope with the temperature conditions for creeping, thereby obtaining a reasonable amount of prestress.

(b) Behavior of concrete shell subject to thermal stress

The thermal stress to which the concrete shell is subjected under operating conditions is an important factor for the structural design of the shell, and its action is very influential to cause cracking in the concrete shell which may reduce air-tightness of the shell or allow the seawater to come into the inside of the concrete, thus corroding steel material inside or losing durability.

Various tests have been conducted for thermal stress. Figs. 5.5-9 and 5.5-10 give an example of temperature distribution and thermal stress distribution at the shell section when heating the interior of the prestressed concrete shell. The test circumstances are as shown in Photo 5.5-6.

The result shows that the temperature distribution in the shell well matches the values obtained through a non-steady two-dimensional heat transfer analysis process, and the thermal stress arising on the shell also well matches the values obtained through a conventional elastic analysis process with the shell regarded as a rigid box frame structure. A similar result has been obtained as well regarding the reinforced concrete shell. Temperature distribution of the shell and dimensions of thermal stress can now be obtained through analytical process under a given boundary condition. That thermal stress can be expected correctly as above is to ensure reasonable design of the shell and means, at the same time, that a stage for heat-up and cool-down at the plant can be determined at the shortest period of time for the operation method of the desalination plant using a concrete shell.

On the other hand, behavior of a shell kept under conditions where thermal stress works for a long period of time undergoes a change entirely on the shrinkage side due to creep and drying shrinkage of the concrete, and particularly creep may bring about a deformation in the case of prestressed concrete construction as illustrated in Fig. 5.5-11, which has been discovered experimentally. These must be evaluated carefully for design of a prestressed concrete shell.

(3) Air-tightness of Concrete Shells

A high air-tightness is required for a shell used for an evaporator, because air leakage into the shell reduces heat conductivity of the heat exchanging tube to the detrimental effect of thermal efficiency and is also unacceptable in view of corrosion on the material. Air permeability of the concrete shell is closely related to water permeability, and the coefficient of air permeability comes 1,000 - 10,000 times as large as that of water permeability; as for air-tightness of the entire shell, it is conceivable that air leakage is large, particularly from construction joints of the concrete and also from mating zones of the hardwares mounted on the shell, and various tests were carried out to check how they were influential. The result shows that coefficient of air permeability of the concrete itself is very small at 10^{-7} cm/sec. and that the air leakage is severest from mating zones of the steel materials around through holes of the pipe embedded in the concrete and nextly from the concrete construction joints as shown in Fig. 5.5-12. These portions can reasonably be improved by providing air stoppers, which has been proven through tests. On the other hand, any reduction of air-tightness which may cause a detrimental influence on the plant operation has not been observed with the concrete shell used for "Study with Concrete Shell Module" and "Study with 100,000 m^3/day Test Module" during long operation. Judging from the result, such a shell as is satisfactory in air-tightness and can be manufactured through employing a suitable structural type for the concrete shell and carrying out close treatment for air-tightness on construction.

(4) Economical Advantage of Concrete Shells

The concrete shell is very economical as compared with the steel shell, as it is provided with superior characteristics in easiness of construction work, durability, maintenance, etc. in addition to its raw material being cheaper by far in comparison with steel. What is characteristic of economy of the concrete shell is described as follows:

i. Materials used

For the concrete shell its raw materials are cheap and obtainable with ease-almost all in the neighborhood of the plant construction site. In addition, the concrete shell is built on a principle to make it at the site by using materials supplied at the site, thus curtailing transportation expense.

ii. Ease of construction work

Shape and size of the shell can be selected fairly freely, and the necessity of skilled workers, like welders in the case of steel shells, is limited only to a part, and the labor cost can be minimized accordingly. The shell can be prefabricated then, according to conditions.

iii. Maintenance and management of equipment

Since concrete is superior in durability against hot seawater, maintenance and management of the shell can be facilitated during long operation. Maintenance-free in other words.

iv. Merits of larger-sized shells

Generally, the larger the concrete structure, the better is the conversion efficiency of molds, workers, etc. Furthermore, increase of the repeated work itself improves the labor technique, and thus a larger-sized shell may bring about more merit.

There were several cases in which a concrete shell was compared with a steel shell for construction cost in the past, but in every case a result was obtained that the larger-sized the equipment, the more advantageous was the concrete shell, and the smaller-sized the shell on the contrary, the cheaper the steel shell.

Various factors are involved for the boundaries to exist, however, the concrete shell will be more advantageous economically in the case of a scale over several ten thousand cubic meters per day, as a guideline from the past tests. In comparison for construction cost of the evaporator shell only, there is a case where the concrete shell costs less by more than 20 - 30% with reference to the scale 100,000 m^3/day.

In a conceptional design carried out by the OSW in the 1960's, the concrete shell came about 30 - 40% cheaper on an average as compared with the steel shell. The values given in the said conceptional design are not necessarily related directly with the comparison in cost in Japan, but they are expected to serve as a comparison in costs of the two.

In the case of concrete shells, there is room to reduce its cost further through improving the method of construction work in the furture. For example, availability for prefabrication of the concrete shell, rationalization of how to mount internal equipment for the shell, etc. - these will be feasible in the future, and thus the concrete shell is expected to be more and more economical as compared with the steel shell.

5.5.4. Mortar Lining on Steel Shells

(1) Performance of Mortar Lining

The seawater desalination equipment working practically at present all uses steel evaporator shells. However, what is most problematical for the steel evaporator shell is corrosion on steel materials. Various anticorrosive lining processes have been developed to prevent the steel material from corrosion. The following is to classify them by material for use:

(a) Using organic materials ..... Epoxy resin, vinyl ester resin, etc.

(b) Using metallic materials .... Titanium, cupronickel, stainless
steel, etc.

(c) Using inorganic materials ... Cement mortar

For organic materials, tests have been carried out by each country including the U.S. Bureau of Reclamation. However, material which is durable for a long period of time under such severe conditions as seawater desalination equipment has not yet been developed. As for metallic materials, those having the necessary durability and performance have already been developed for practical use, however, a high lining cost is inevitable in each case, leaving economical problems. Such being the situation, it can hardly be said that such organic or metallic materials as will satisfy both performance and economy have been developed at present.

Meanwhile, lining using cement mortar as inorganic material is regarded as promising in economy, durability and performance despite having not been used on a practical shell over a long time. As described in the paragraph on concrete evaporator shells, cement materials are not only economical but also exceedingly durable against hot concentrated brine due to using proper material and executing work with scrupulous care. This has been proven through various tests. Furthermore, the cement mortar is itself a material of very high alkalinity, and hence a steel surface covered with mortar is kept under high alkaline conditions and thus put in an condition to prevent corrosion. As described above, the cement mortar lining has been used experimentally on the testing plant at the Fontainvalley Test Facilities, U.S.A. In this case, mortar using alumina cement was applied on the inner surface of the evaporator shell and also on high temperature piping at 25 mm in thickness.

On the other hand, a mortar lining was applied experimentally on the high temperature stage of the equipment for the 100,000 m^3/day Test Module in Japan, which was put in operation for about one year and a half. Upon closing the operation, the mortar was removed to see how the steel surface was corroded, but it was almost free from corrosion, which demonstrated the superior performance of the mortar lining. Judging from the result, it is conceivable that the cement mortar lining is, in its performance, serviceable as a lining material for a steel sheel. The mortar lining applied on the Test Module Plant is as shown in Photo 5.5-7.

(2) Applying Method and Features of Mortar Lining

(a) Applying method of mortar lining

The applying method of mortar lining is schematically as follows:

(i) Pretreatment: This is an indispensable process whatever material may be used for lining; the method for treatment varies according to materials used and necessary expenses are naturally variable. Not such a high standard of pretreatment as in the case of resin material will be necessary for the mortar lining, and that of removing rust or oil stains on the steel surface will be enough.

(ii) Steel wire fabric mounting: Wire fabric is used for the purpose of increasing the strength of the mortar lining itself, preventing mortar from peeling off on the execution of work and improving the adhesion of mortar and lining face mechanically through studs, and is mounted on the steel surface when finishing it with pretreatment. The wire fabric is made of steel or stainless steel, which will be selected for use according to the conditions working on the lining.

(iii) Mortar finish: The conditions working on the lining are very severe in comparison with general concrete structures, and hence a high durability is required for the mortar itself. Therefore, the mortar must be adaptable to execution of the work and compact in addition. A spraying process is optimized for mortar and the final touches on the surface must be performed by trowelling.

(iv) Curing: The mortar lining forms a stable lining through the reaction of cement in mortar with water and thus gives full play to lining performance. Therefore, curing by watering or wetting will be necessary until strength in the mortar is fully generated.

(b) Features of mortar lining

Mortar lining has may features, which are pointed out as follows:

a. Economical.
b. Superior in performance for corrosion prevention.
c. Materials for use can be obtained easily near the site.
d. Superior in working property.

The features given as above will be described briefly as follows:

(i) Economy: What is featured with the mortar lining above all is its economy. Since the main material is cement and sand, which are cheap compared with those of the resin or metal groups, the lining cost is extremely low as compared with other materials. There is no case antecedent ever before where the lining cost was integrated under the same conditions, therefore an accurate comparison in cost is not available right away. However, in calculating the estimations tried by plant manufacturers, it is conceivable that the resin group lining is about 1.5 - 2.0 times, the stainless steel lining 4.0 - 5.0 times and the cupronickel lining 6.0 - 7.0 times, cost of the mortar lining being taken at 1.

(ii) Performance: The mortar lining is provided, as already described, with full performance working on the steel shell. It goes without saying that it is trustworthy in performance. Also, it can be so affirmed on the premise that the working has been done under strict supervision.

(iii) Working property: The materials for use can easily be obtained near the construction site, and its handling is relatively free from being influenced by quantity.
In addition, special machinery and skilled workers are almost not necessary for the work, besides mortar spraying equipment, and thus the work property can be regarded as superior in comparison with other lining materials.
As for the working method, it can be worked in parallel with the manufacture of the steel shell at the shop beforehand as well as on the site.

5.6. Operation and Maintenance

5.6.1. Process Control

A seawater desalination plant of the brine recirculating multistage flash evaporation type is ensured of fairly stable operation.

The largest possible disturbance for the multistage flash evaporator is a change in heat input, (which may safely be regarded as constituting the whole disturbance,) particularly in the case of a plant working for the double purpse of power generation and desalination. A change is temperature and property of the raw seawater may be pointed out as another possible disturbance.

To keep the equipment stable in operation at all times against these disturbances, a process control system shown in Fig. 5.6-1 will be necessary. The system comprises those factors which are given below.

1. Heating steam temperature control system
2. Circulating brine temperature and flow control system
3. Seawater temperature control system
4. Make-up seawater flow control system (circulating brine density control)
5. Control system for make-up seawater level in decarbonator
6. Control system for brine level in final stage
7. Control system for distilled water level in final stage
8. Control system for condensate level in brine heater
9. pH control system

(1) Control of each system

(a) Heating steam temperature control system

The heating steam to the brine heater consists of vented and exhaust steam of the power generating turbine, exhaust steam of the auxiliary turbine for the driving pump and steam of the exclusive boiler, and is superheated in every case.

The brine heater outlet brine temperature is limited below 120°C so as to avoid a deposition of calcium sulfate. However, an excessively high temperature of the heating steam may allow the temperature of the heating tube wall to exceed 120°C, and hence the heating steam temperature will preferably be kept below 130°C. A heating steam temperature control combined with a desuperheater will be necessary in consequence.

For the control system, two types of desuperheating devices work: one is the steam atomizing type desuperheating device as shown in Fig. 5.6-2 and the other is the spray type as shown in Fig. 5.6-3. In comparison of both, the steam atomizing type is superior in every respect such as minimum pressure difference required for desuperheating water ($\Delta P = P' - P$), treated quantity of desuperheating water (W), range ability against load fluctuation, remaining drainage and necessary length of straight pipe.

(b) Circulating brine temperature and flow control system

The circulating brine temperature is maximized in value at the brine heater outlet and minimized at the final stage flash chamber outlet. The difference between the maximum value and minimum value of the circulating brine temperature is called the flash temperature range, which is a factor to improve thermal efficiency. Since desalinated water quantity is proportional to the product of the flash temperature range and circulating brine quantity, to make the flash temperature range large is to decrease circulating brine quantity. It is therefore desirable to raise the brine heater outlet temperature. However, the temperature is limited below 120°C due to a deposition of calcium sulfate scale. An accurate temperature control is required to keep the scale from depositing.

There is a close relation between the circulating brine flow rate and the circulating brine temperature. That is, in the multistage flash evaporator, the brine fluidizes through evaporation stages according to interstage differential pressure which is determined by the saturated steam pressure coping with the brine temperature. The interstage differential pressure therefore varies as the brine temperature changes. However, since the interstage orifice is fixed, the change in differential pressure is compensated for by the head of the brine level. For example, the interstage differential pressure falls because of the brine temperature falling. Therefore, the brine level in the high temperature stage rises to allow the brine to flow. The saturated steam pressure changes sharply with temperatures of 100 to 120°C, therefore the brine level at the high temperature stage or the initial stage particularly rises, even when the brine temperature falls slightly, and thus carryover seriously deteriorates the purity of the desalinated water. To cope with such fluctuations of the brine level, the circulating brine flow rate will have to be adjusted.

Fig. 5.6-4 illustrates the control system. The control system is characterized by adjusting the rate of circulating brine flow in accordance with the circulating brine temperature. Setting of this control system will be a factor influencing stability of operating the plant in harmony with the construction of the evaporator.

(c) Seawater temperature control system

The cooling seawater flowing in the condenser of the heat rejection section functions to receive the major part of heat coming into the system from the brine heater and reject it outside. Therefore, a change of heat input to the process and a change of cooling seawater inlet temperature disturb the cooling seawater system.

The control system includes an outlet seawater temperature control shown in Fig. 5.6-5 and an inlet seawater temperature control shown in Fig. 5.6-6.

In the inlet seawater temperature system, change of the raw seawater temperature doesn't matter, but disturbances of the heat input are influential. On the other hand, the outlet seawater temperature system is free from interference through change of the raw seawater temperature and disturbance of the heat input. Therefore, the outlet seawater temperature control system will be preferable in the case of a plant working for the double purpose of power generation and desalination where a big fluctuation is expected in the heat input.

(d) Make-up seawater flow control system (circulating brine density control)

In the case seawater desalination plants employing the evaporation process, it is necessary to keep the density of circulating brine in the system below allowable limits at all times, thereby preventing deposition of calcium sulfate scale. In addition, brine density must be determined in consideration of the boiling point elevation of the brine and pretreating cost of the make-up seawater. Control of the make-up seawater flow is important in this respect.

The control system comprises an independent control of Fig. 5.6-7 and a proportional control of Fig. 5.6-8.

In the case of such a system to control the make-up seawater flow independently, the quantity of the make-up seawater is constant when the quantity of desalted water decreases, and hence no change will be made to pretreating cost. In the case of a system to control the make-up seawater flow in proportion to desalinated water quantity, the concentration ratio of the brine is always constant, no scale deposition must be expected, and the make-up seawater decreases in quantity upon decreasing desalinated water quantity, thus cutting the pretreating cost. The proportional control system will be preferable accordingly.

(e) Control system for make-up seawater level in the decarbonator

Where the pH control method is employed for scale prevention, the make-up seawater is decarbonated by adding acid to change bicarbonate ions into carbonic acid gas and is then introduced into the deaerator to have the dissolved oxygen removed and supplied to the evaporator.

Now, while the decarbonator is open to the air, the deaerator is high vacuumized in this case, therefore seawater must be absolutely prevented from bringing the air into the deaerator. It is therefore desirable that seawater will be supplied under air-tight conditions with both the decarbonator and deaerator in this control system.

The control system controls the level as shown in Fig. 5.6-9. The illustration represents the case of a final stage deaerating system where the deaerator is enclosed in the final stage evaporating chamber.

(f) Control system for brine level in final stage

Where the final stage brine level drops, the brine circulating pump will give rise to cavitation, thus keeping the plant in extremely unstable operation. This status left for a long time is capable of damaging the brine circulating pump. Carryover increases as the level rises, and the desalinated water deteriorates in purity in consequence. Therefore, the brine level control in the final stage is important for operation.

The control system controls the level as shown in Fig. 5.6-10. In the evaporator, seawater which is surplus to the quantity of desalinated water is supplied to control the brine density to a given value. It is then necessary to discharge the brine which is the difference between the quantity of the supplied seawater and that of the desalted water, and the brine thus discharged is extracted out of the evaporator final stage. This control system is to adjust the quantity of discharged brine and control the evaporator final stage brine level.

In small and simple equipment, a pump for exclusive use on the discharged brine is not fitted and can be substituted for by the brine circulating pump as illustrated in Fig. 5.6-11.

(g) Control system for distilled water level in final stage

This control system is intended to extract the distilled water as desalinated water, while air-tight, out of the evaporator final stage which is kept vacuumized, as shown in Fig. 5.6-12.

(h) Control system for condensate level in brine heater

Intended to discharge the condensate in the brine heater with constant level against any load fluctuation. The control system is as shown in Fig. 5.6-13.

(i) pH control system

The pH control system is necessary where the decarbonation pH control method is employed for prevention of alkali scale.

The system for decarbonation is of pH self restoration: the quantity of sulfuric acid to infuse will be controlled so that self-restoration may be secured with pH at decarbonator outlet.

The control system is as shown in Fig. 5.6-14. The arrangement is such that pH of the circulating brine is detected and caustic soda is added, which is for emergency coping with an abnormal drop in the brine pH caused by erroneous operation or failure of the plant.

Where fluctuation occurs frequently in the quantity of make-up seawater, it is preferable to adjust the quantity of sulfuric acid to infuse first in proportion to the quantity of make-up seawater and then take the pH of the seawater at the decarbonator outlet for fine adjustment, as illustrated in Fig. 5.6-15.

(2) Instrumentation for plant control

Which type of instrumentation to take, electric or pneumatic, will be determined in consideration of precision, plant capacity, number of instruments and combination with the computers which are required for the plant.

Electric type instrumentation has merits of being easy in connection with computer control and cheap in construction costs.

The pneumatic type ensures stoutness and simple operation of the machinery and instruments. A system diagram of the pneumatic equipment is shown in Fig. 5.6-16.

(3) Detecting elements for control

(a) Temperature detecting element

The temperature detecting element is represented by a thermocouple and thermometric resistance. The resistance bulb is superior in measuring precision; the thermocouple is superior in stoutness. The thermocouple is suitable for the flash evaporator, as oscillation is inevitable at the time of flash evaporation and brine fluidization.

(b) Flow detecting element

The flow detecting element comes as a differential pressure type and an electromagnetic type. The differential pressure type is preferable in price and precision. An orifice plate, Venturi tube and nozzle are used as a throttle mechanism for the differential pressure type. Notwithstanding the large pressure loss involved the orifice plate is suitable because of its cheapness, short length of straight pipe required, easiness of replacement and stability of performance.

An electromagnetic flow meter is used for measurement in the ball circulating system.

(c) Level detecting element

The level detecting element is a displacer type utilizing buoyancy and differential pressure types. Both are serviceable.

(d) Density detecting element

The density measurement can be converted by measuring the liquid for conductivity. The conductivity detecting element is electrode type and electromagnetic type. For fresh water the electrode type is preferable because it is superior in precision; for seawater and brine the electromagnetic type is preferable because it is comparatively free from staining on the cell.

(e) pH detecting element

For the pH detecting element, the glass electrode type is recommedable generally because it has stable performance.

(4) Automation

It is desirable that the plant should be automated for operation under the best possible conditions to prevent start-up, shutdown and fluctuation of heat input. Such a process of unattended operation has been employed particularly for plants working for double purposes.

A start-up process of a multistage flash evaporation plant is schematically given as follows:

(a) Seawater pump start-up process

Before starting up the seawater pump, ensure that each valve in the seawater line is in the correct state.

(b) Water filling process

Open the seawater pump outlet valve and lead the seawater into the final stage of the evaporator by way of the condenser of the heat rejection section and decarbonator.

(c) Vacuum increasing process

Start-up the ejector and step up the vacuum so as to obtain the necessary vacuum for flash evaporation of the brine.

(d) Brine circulating pump start-up process

Start-up the brine circulating pump normally.

(e) Process for raising brine temperature up to 60°C through the program setter

Commence the decarbonator by starting up the pump for infusing sulfuric acid, start-up the brine heater and raise the maximum temperature up to 60°C by the program setter.

At first flash evaporation doesn't occur, and the brine flows through each stage only by its head difference. A steam pressure difference a rises as the temperature rises gradually to flash evaporation, thus allowing the brine to flow easily. Start-up the desalted water line automatically and then the condensate line.

(f) Process for raising brine temperature up to maximum through the program setter

Raise the brine maximum temperature, as controlling pH value, up to the rated value (104°C in this case) as designed by the program setter. How to set the relation of brine temperature and flow of circulating brine is important in this process. The start-up operation is completely over when the program setter shuts down, each pump has completed start-up and each valve is in a normal state.

A principal block diagram to control start-up and shutdown of the desalination plant is shwon in Fig. 5.6-17. In the case of start-up, the arrangement is such that the central program setter is actuated by signals from the sequencer, the brine maximum temperature is set against the time, and the rate of circulating brine flow is set against the brine temperature.

In the above conception, automatic operation of the multistage flash evaporator can be carried out.

(5) Emergency measures

Emergency measures for the plant can be classified into the following according to the extent of influence:

(a) Those for which operation of the plant must be stopped immediately after the occurrence of abnormities

1. Total power failure
2. Breaking of power supply for instrumentation
3. Breaking of air supply for instrumentation
4. Stop of circulating brine pump or abnormal drop in pressure
5. Stop of seawater pump

(b) Those for which the plant operation is to be stopped after occurrence of abnormities according to the duration of the abnormal situation and circumstances for restoration

1. Drop in steam pressure
2. Drop in delivery pressure of each pump
3. Drop in rate of make-up seawater flow
4. Heater outlet brine temperature being high or low
5. Stop caused by overload on auxiliary equipment driving machines

(c) Those for which the plant cycle is altered partly by occurrence of abnormities

1. Deterioration in purity of fresh water

   The fresh water is drained into a rejection line due to deterioration in purity.

2. Deterioration in purity of condensate

   The condensate is drained into a rejection line due to deterioration in purity.

3. Drop in rate of make-up seawater flow

   The sulfuric acid infusing pump is stopped because of a drop in the rate of supplied seawater flow.

(d) Those for which operators are acquainted with the abnormality by alarm lamp and alarm buzzer

1. Each abnormality indicated in paragraphs (a) to (c)
2. Brine pH being high or low
3. First stage brine level being high
4. Final stage brine level being high or low
5. Final stage distilled water level being high or low
6. Final stage pressure being high (degree of vacuum being low)
7. Decarbonator water level being high or low
8. Drop in air supply pressure for instrumentation
9. Condensate level being high or low

5.6.2. Measurement

(1) Epitome

Measurement of temperature, rate of flow, pressure, etc. is necessary for smooth operation and control of the desalination plant. It is desirable to control properties of the seawater, i.e. dissolved oxygen, total carbon and pH above all for prevention of scale and corrosion. It is further necessary to be acquainted with each value of COD and BOD for the circumstances in which the seawater gets polluted. On the other hand, it is also important to analyze the components of brine in operation and the components and status of scale and sludge sticking to the heating tubes and vessel to be checked after operation.
To carry out instrumentation and analysis accurately in keeping pace with operation of the plant is indispensable for performance and maintenance with the plant. The following refers to item of measurement and analysis and methods.

(2) Measurement

(a) Temperature

Thermometry is an important item for a firm grasp of desalination efficiency. Items and portions to measure vary according to the system of the plant and the purpose of operation. However, thermometry is conducted generally to obtain overall heat transfer coefficients of the brine heater and condenser at each stage, brine evaporation characteristics and performance of the decarbonator and deaerator.

Thermometry is then conducted practically at 10 - 150°C in range and 0.05 - 0.1°C in accuracy by means of thermocouples or thermometric resistance. The output corresponding to temperatures measured is transmitted as an input to the controlling equipment and is converted to indicate temperatures simulataneously.

(b) Pressure

In the case of a flash evaporator, the evaporating chamber and the deaerator are decompressed with vacuumizing equipment like the ejector and at the same time, steam at 2 - 5 $kg/cm^2 \cdot G$ is used for heating of the brine heater. It is then necessary to detect the delivery pressure of the pump in accordance with the make-up and recirculation of the seawater.

Therefore, heating steam pressure, ejector-supplied steam pressure, degree of vacuum of the ejector, degree of vacuum of the initial stage, final stage and deaerator, and delivery pressure of the seawater pump, make-up seawater pump, brine circulating pump and fresh water pump are measured. The measuring method depends on required precision. However, the measurement is made generally by indication on Bourdon tubes at the site or by detection with diaphragms and transmission through conversion into the quantity of air or electricity. For delivery pressure of the pump, operation of the pump is controlled by interlocking with the pressure switch.

(c) Rate of flow

The rate of flow is an important measuring item as in the case of temperature. Accurate measurement is desired particularly at a plant of large capacity. For measuring items at normal plants working according to the evaporation process, the quantity of seawater, quantity of the make-up seawater, quantity of circulating brine, quantity blown down and quantity of desalinated water are taken up.

For measurement an orifice is used, and the generated differential pressure is transmitted by means of an oscilator. For flow control the deviation arising between the detected rate and a set point is adjusted by control valve.

(d) Density

Density of seawater is proportional to electric conductivity, and hence a means to detect the conductivity is used for measurement. Particularly in case the concentration ratio is to be obtained simply, the measurement is made according to the conductivity of raw seawater and circulating brine. This method applies to the purity of fresh water produced, too.

(e) pH

Supervision of pH is necessary at all times for suppression of scale formation and retardation of corrosion. Measurement it conducted with a circulating electrode incorporated with a glass electrode, comparison electrode and temperature compensating electrode.

The measuring objects include raw seawater, make-up seawater, circulating brine and fresh water produced. Since the electrode works for pH measurement, a stain on the surface of the electrode involves a problem of precision, and calibration by a standard liquid will be necessary once a day.

(3) Analysis

(a) Total carbonic acid

The carbonic acid radical in the seawater varies in dissociation according to conditions, and hence there taken is the method used is to measure the density of total carbonic acid as carbonic acid, bicarbonate ions and carbonate ions. The measuring method comprises the chemical analysis method employed by Japanese Industrial Standards and the continuous measuring method. Outline and problematical points for each method are referred to as follows:

(i) Chemical analysis method

The method specified in JIS turns the dissociated ions into carbonic acid gas by adding acid to the seawater and diffusing free carbonic acid by boiling and heating. The diffused gas has sulfurous acid gas absorbed and removed by a sulfuric acid solution saturated with potassium dichromate and is then absorbed in a barium hydroxide solution. The diffused gas is circulated for absorption, and then the total carbonic acid is converted to carbonic acid gas to calculate through acid neutralization to titration. A warning for this operation is that fluctuation in pressure of the circulating gas due to boiling should be minimized and that the sample should be kept from diffusing outside, where it is over saturated, at the time of infusion into the boiling tube. For improvement of this method, barium chloride should be added to decrease dissolution of barium carbonate. There is a further report that since a reaction of potassium dichromate in the cleaning agent with carbonic acid gas is conceivable, it is desirable to use sulfuric acid only for suppressing the reaction.

(ii) Equipment analysis method

The method of analyzing carbonic acid on equipment uses infrared absorption. A system of the device construction is shown in Fig. 5.6-18. First, nitrogen gas is extracted by separating carbonic acid radical with acid. The extracted gas is then detected in a nondispersive infrared absorption area covering 2.8μ and 4.3μ which are the absorption wavelengths for carbonic acid gas. No problem will particularly occur for the sensitivity of detecting carbonic acid gas. However, a method must be contrived for improving efficiency of the extractor.

(b) Dissolved oxygen

Values of less than several ten ppb must be assured for the performance of the deaerator for the desalination plant, and hence care should thoroughly be taken for analysis of dissolved oxygen with such a low density not only in the analyzing process, but also in the sampling process. The analyzing method includes chemical analysis and equipment analysis as in the case of total carbonic acid. For the former the Winkler method, the sodium azide method, and the Miller method are used. However, the Winkler method employed in JIS is most popular. In the case of equipment analysis, equipment to detect oxygen obtained by membrane extraction by means of polarograph or galvanic cells has been developed.

Operations and warnings for the Winkler method and galvanic cell system are given as follows:

(i) Winkler method

This method is to fix oxygen as manganese hydroxide by adding manganese sulfate and sodium hydroxide to the seawater taken in the sampling tube. Where dissolved by adding acid under the presence of iodine ions, the precipitate liberates iodine according to the oxygen quantity, which is titrated to determination by reducing agents like thiosulfuric soda. For correct judgement of the titration, potentiometric titration is used. In sampling, care should be taken for rapid fixation so as not to involve bubbles from outside. For possible influence by interfering ions, it is necessary to investigate coexistent components beforehand. Iodine added beforehand is to cope with the reducing matter like sulfite ions. However, the allowance is limited to 0.05 ppm or so for iron ions or bivalent ions particularly.

(ii) Equipment analysis method

A process to extract oxygen by means of a membrane and a process to extract it directly with carrier gas are used for the galvanic cell system. In the former membrane system, a stain on the extracting membrane is capable of deteriorating efficiency; in the latter direct system bubbling is problematical. For these analysis methods a system of British-made EIL-Cambridge dissolved oxygen meters is shown in Fig. 5.6-19. This equipment has a double measuring range of 0 - 20 ppb and 0 - 200 ppb. The principle is such that sample water is put in a scrubbing tower, oxygen is extracted by hydrogen gas generated by electrolysis and then detected with a platinum-gold electrode. For span check, oxygen is generated by electrolysis of water and a liquid of constant concentration can be produced. An example of the operating analysis meter is shown in Fig. 5.6-20 on recording paper. The figure illustrates a change in concentration obtained in a characteristic test on the deaerator and also shows Winkler analytical values. As will be apparent, the analysis meter operates well, which matches Winkler values. To keep the meter working in good condition, calibration with a standard solution will have to be carried out correctly. For analysis on the equipment, it is necessary to contrive how to avoid leakage and obtain better substitution of the sampling line.

(c) Analysis of scale and brine

Compared to the analysis methods given as above on total carbonic acid and dissolved oxygen, scale sticking on the heating tube and sludge are to be analyzed after operation of the plant. It is also desirable to trace the changes of scale component in the circulating brine during the operation. These items for analysis being correlated with each other, items and methods are shown collectively in Table 5.6-1.

After observing scale and sludge for sticking condition, the sample is subjected to emission spectrochemical anslysis, for qualitative examination of the component, and identified, at the same time, on an X-ray diffraction device for identification of the structure. Organic matter in the sample, or decomposition by heating are analyzed according to the gravimetric process at around 1,000°C or the differential thermal gravimetric process.

Next, scale and sludge are dissolved analyse the components. As general components, calcium (Ca), magnesium (Mg), silica (SiO_2), sulfuric acid radical ($SO_4{}^{2-}$), iron (Fe), copper (Cu), aluminum (Aℓ) and titanium (Ti) are present. Each analysis method is shown in Figs. 5.6-21 - 5.6-25 along with the dissolution method of the scale and analysis method of the brine. The colorimetric method, turbidimetric method and gravimetric method apply to silica and sulfate ions. The atomic absorption spectrophotometry method is popular for all these metal components-calcium, magnesium, iron, copper, aluminum and titanium. However, it is desirable to use a chemical method together with it because of interfering ions and sensitivity.

Meanwhile, chlorine and COD necessary for obtaining density and degree of pollution of the seawater and brine are illustrated collectively in Fig. 5.6-26. The chlorine ion method of analysis of density in the scale is shown therein. COD is oxygen quantity which is required chemically, and the potassium permanganate alkaline oxidation method is used more than all the methods.

5.6.3. Ball Cleaning

(1) Principle of ball cleaning

Acid cleaning, water jet cleaning and brush cleaning are available for elimination of stains on the heating tubes. However, those require to have the plant shut down and hence are effective only at regular inspection.

Ball cleaning is a method for cleaning the heating tubes inside by throwing sponge balls into the brine and thrusting them in the tubes to pass through together with the brine because of the differential pressure working at the inlet and outlet of the heating tubes. This method is available for eliminating stains on the heat transfer tubes while the plant is kept in operation.

(2) System of ball cleaning equipment

A general system of ball cleaning equipment is shown in Fig. 5.6-27. The equipment consists of a ball catching unit, a ball recovering unit and a ball circulating pump. Stains on each heating tube can be eliminated by throwing the sponge ball into the ball recovering unit, allowing it to flow into the inlet of the condenser for the heat recovery section of the brine circulating system on the stream from the ball circulating pump, allowing it to pass through the heating tube for the heat recovery section condenser and further allowing it to pass through the heating tube for the brine heater. The ball is separated from the brine by the ball catching unit and thus pulled out of the brine heater outlet, boosted by the ball circulating pump to pass the ball recovering unit, and then thrown again into the brine at the condenser inlet for the heat recovery section. The ball recovering unit can be controlled here by valve operation so as to allow the ball to pass through or recover it.

(3) Construction and operation of ball cleaning equipment

(a) Ball catching unit

The construction of the ball catching unit is shown in Fig. 5.6-28. This unit is arranged in the outlet brine piping of the brine heater, operating to separate the ball and extract it from the brine. There are inclined grids arranged in two stages internally in the direction of flow, and along these grids the ball is caught in the catching tube. The grids are of such construction as they can be operated by handle control from outside, and foreign matter sticking to the grids during operation can be washed off in closed position of the grids.

(b) Ball recovering unit

The recovering unit is a double cylindrical container, incorporating a wire gauze basket and operates to recover the ball in the basket or allow it to pass through by exchange valve. Then, the ball can be thrown into the system and extracted by the ball recovering unit.

(c) Ball circulating pump

The pump boosts the brine containing the ball, using a bladeless pump so as not to damage the ball.

(d) Sponge ball

Fig. 5.6-29 shows the sponge ball. The ball is of continuous cell sponge and so arranged as to have nearly the same specific gravity as water when water is absorbed. The material is mainly natural rubber, withstands brine at 130°C, flows smoothly through the tube and is elastic enough to remove sludge which sticks inside the tube effectively when it flows through.

(4) Cleaning effect

For prevention of scale on multistage flash evaporating aeawater desalination plants, the pH control method is generally employed. In this case, corrosive products on the vessel are turned into sludge, because the operation covers a long period of time, and thus stick on the heating tubes. Therefore ball cleaning becomes effective. The following gives cases wherein ball cleaning applies:

(a) Desalinated water quantity 100 t/d, 10-stage flash evaporation plant

Operating conditions	Brine max. temperature:	120°C
	Brine concentration ratio:	2.0
pH controlling conditions	Circulating brine pH:	7.2 ± 0.2
Ball cleaning conditions	Heating tube ID:	14 mmø
	Ball diameter:	14 mmø ± 0.5 mm
	Cleaning hours:	2 hr/day
	Ball circulating frequency:	0.5 pc/min./ heating tube 1 pc.

From carrying out 2,000-hour continuous operations under the conditions above, no deterioration was observed in the performance ratio of the plant, and the stain sticking rate could be decreased as much as shown in Fig. 5.6-30 in comparison with a case where ball cleaning was not effected.

(b) Desalinated water quantity 3,000 t/d, 39-stage flash evaporation plant

Operating conditions	Brine max. temperature:	120°C
	Brine concentration ratio:	2.0
pH controlling conditions	Circulating brine pH:	7.2 ± 0.2
Ball cleaning conditions	Heating tube ID:	17.0 mmø
	Ball diameter:	17 mmø ± 0.5 mm
	Ball circulating frequency:	0.6 pc/min./ heating tube 1 pc.
	Cleaning hours:	60 min.

Fig. 5.6-31 shows the effect of ball cleaning in performance ratio according to the time passed. By effecting ball cleaning where the performance ratio decreased to 9.8 from 10.7 for 300 hours, the said ratio recovered to the initial value of 10.7.

(5) Operation and maintenance

No operation is particularly required for the ball cleaning equipment other than the throwing and recovery of the ball. No rotodynamic machines are arranged other than the ball circulating pump, which facilitates maintenance work in consequence. Automated equipment has been completed for this purpose.

5.6.4. Regular Inspection

(1) Regular inspection of all equipment

It is important to detect abnormities arising in all equipment as early as possible by continuing inspection before start-up and during operation of the plant. To effect maintenance thoroughly by general regular inspection which will be carried out once a year is to ensure trouble-free operation with the plant.

Items for regular inspection are shown schematically below:

(a) Desalination plant body

(i) Evaporator

1. Leakage
2. Elimination of stains on heating tubes
3. Check for corrosion
4. Check and replacement of packing
5. Interior cleaning; deposit elimination
6. Check on rubber lining and resin coating
7. Check for abnormities on expanded parts of tubes
8. Cleaning and check of strainer interior

(ii) Brine heater

(iii) Drain cooler

(iv) Ejector condenser

(v) Vent condenser

As in the case of evaporator

(vi) Ejector

1. Leakage
2. Overhauling
3. Cleaning and check of strainer interior

(b) Tank, pump and piping valve

(i) Tank

1. Leakage
2. Interior check, cleaning and elimination of deposits
3. Check for thickness of rubber lining

(ii) Pump and driving turbine

1. Overhauling
2. Check for corrosion on runner
3. Replacement of gland packing
4. Greasing
5. Lubricating oil change
6. Cleaning and check of strainer interior
7. Opening and check of check valves

(iii) Piping

1. Leakage
2. Check for thickness of rubber lining
3. Check for thickness of stainless steel tube
4. Check for clogging of sulfuric acid guide tube
5. Cleaning and check of drain tray

(iv) Valves

1. Leakage
2. Overhauling (seat face, handle)
3. Handle lubrication

(v) Safety valve

    1. Overhauling
    2. Seat face servicing
    3. Blow-off pressure ensuring

(c) Other equipment

For electric equipment, insulation resistance and operating condition will have to be checked generally. Controlling units are to be checked for abnormities in operation, to be carried out an overhaul, and instruments are to be checked for abnormalities in indication.

(i) Power supply equipment

Metal-clad switchgear, control center, site control panel, relay, meter, timing reactor, current transformer, instrument transformer, transformer

(ii) Electric load equipment

Motor, solenoid valve

(iii) Controlling unit

Control panel, recorder

(iv) Controlling units

Indicating process controller, manual actuator, computer, indicator, E/P converter (pneumatic), controllers (temperature, pressure, water level, pH), power pack

(v) Transmitters

Transmitter, pH meter, orifice

(vi) Control valves

Diaphragm control valve and piston valve, motor valve

(vii) Switches

Pressure switch, temperature switch, limit switches, level switch

(viii) Others

Thermocouple, pressure gauge, thermometer

(2) Cleaning of heating tubes

Acid cleaning, water jet cleaning and brush cleaning are used to eliminate scale or sludge sticking on the heating tubes. Acid cleaning is effective for elimination of alkali scale, and water jet cleaning and brush cleaning are effective for elimination of sludge.

(a) Acid cleaning

With $CaCO_3$ and $Mg(OH)_2$ as principal components, alkali scale can easily be dissolved in acid. The acid cleaning piping is as shown in Fig. 5.6-32.

Cleaning process

- Cleaning agent: Hydrochloric acid 3 - 5%
  Inhibitor 0.3 - 0.5%

- Treating conditions

  Carry out cyclic treatment for 6 hours along the guidelines until the dissolved Ca^{2+} and Mg^{2+} become nearly constant at 60°C ± 5°C.

- Measuring items

  Temperature, HCℓ density, Ca^{2+}, Mg^{2+} ..... 1 time/1 hr.

Cleaning process

- Treating conditions

  Wash in water until pH becomes more than 5. Neutralize the initial washings with caustic soda and then discharge.

- Measuring items   Temperature, pH ..... When possible

(b) Water jet cleaning

Inject pressurized water of several tens or several hundred atm. into the heating tube and thus eliminate sludge in the stream. Effective where there is no scale separated.

(c) Brush cleaning

This is a method to mechanically eliminate matter sticking inside the heating tubes by brushing in an air current or water stream. Effective cleaning will be obtainable through changing the brush specification according to the conditions of sticking matters. However, the working efficiency of this method is somewhat inferior and care should be taken not to mar the heating tube surface.

5.7. Heat Utilizing Cycle

5.7.1. Utilization for Dual-Purposes with Power Plant

(1) General Trend

Steam works overwhelmingly as the heat source for the seawater desalination plant operating according to the multistage flash evaporation process.

The quantity of the steam heat is the sum of so-called sensible heat and latent heat; in the power plant, the sensible heat only is utilized for the purpose and the latent heat is wasted outside the system through a condenser or by some other alternative means. Then, the steam latent heat can be utilized effectively for the seawater desalination plant.

It is well known that the latent heat covers a considerable part in quantity of the whole that steam provides. Such being the circumstances, it can be presumed that advantage can be obtained from utilizing the steam for dual-purposes, with its sensible heat used for power generation and latent heat working as the heating source for desalination, instead of planning both the power plant and desalination plants separately. This may be further possible for lightening loads on the condenser, particularly in equipment cost, and also for ensuring a scale merit on the steam generating equipment.

In view of the situation mentioned above, the matter of dual-purpose plants combining the power plant and desalination plant has been discussed for a long time bringing forth a variety of literature. What is notable above all is the so-called Bolsa Island Project, with which Bechtel was entrusted for investigation by the Metropolitan Water District - MWD, Southern California. Seeing the introduction of a seawater desalination plan as significant as well as economical so as to cope with the future trend of sypply and demand for water and also with possible unexpected accidents on Colorado and Los Angeles headraces the MWD depends on (these headraces pass on a fault of the stratum in an earthquake zone and consequently an earthquake may cause an accident at any time), MWD, responsible for securing water for supply and demand in Southern California (including Los Angeles), made a contract with the authorities concerned, of the Department of the Interior and Atomic Energy Commission in August 1964, for investigation on economic and engineering feasibility and preliminary design of a dual-purpose plant 190,000 - 570,000 m^3/day for fresh water and 150,000 - 750,000 kW in power generation. Bechtel was actually entrusted with investigation in December 1964.

The main point of the result was that cheap desalination water would be obtainable through constructing a dual-purpose plant provided with 2 light-water type reactors using nuclear fuel and generating 570,000 m^3/day of water and 1,800,000 kW of power on an artificial island 174,000 m^2 or so in area and about 800 meters off Hunchington Beach in the suburbs of Los Angeles. The water cost was estimated at ¥20.4 per cubic meter at that time. The cost would be calculated according to the Power Credit Method described later.

The necessary funds were estimated at 444,000,000 dollars at first, which was divided among MWD, the electric power company and the Government, and that part for the Government share was signed by the President at that time in 1967. Then, a review was made of the construction expenses resulting in increasing the necessary funds up to 765,000,000 dollars due to factors of extension of the term of work, rise in equipment costs and interest, strengthening of safety standards, increase in generating capacity, etc., thus interrupting execution of the project.

The next important literature is the Technical Report Series No.69 "Costing Methods for Nuclear Desalination" (1966) by the International Atomic Energy Agency (IAEA) in which the economy of a dual-purpose plant was taken into consideration.

As for Japan, the cost of a dual-purpose plant working for thermal and atomic power generation was estimated in the "Report on Technical Development of Seawater desalination" by the National Institute of Resources, Science & Technology Agency in July 1967. Further, the Japan Atomic Industrial Forum held a meeting for research on atomic power desalination in April 1968 to examine the feasibility of combining the atomic reactor and the desalinating plant and estimate desalinating costs. The result was submitted as the National Report to the International Atomic Energy Symposium on Nuclear Desalination held under the sponsorship of IAEA at Madrid, Spain in November 1968.

In parallel with the trends above much literature has been presented in various magazines personally, or in other names since then - to say nothing of the public reports, and thus it can be said that literature through which common knowledge on dual-purpose plants will be obtainable relatively easily.

However, figures given in this literature represent the situation before the Oil Shock, and hence modification is required at present. In a dual-purpose plant the brine heating steam works chiefly where the power plant and the desalinating plant come in contact, therefore merit from designing dual-purpose plant can be figured out through determining the steam cost. From the above point of view, the heat utilizing cycle for the dual-purpose plant and IAEA's costing method are described briefly as follows. An actual case wherein the steam cost was calculated by applying the said method is also introduced.

2) Heat Utilizing Cycle for Dual-Purpose Plants

(a) Combination of condensing type with extraction turbine

This is a method to utilize the steam bled from the intermediate stage of the condensate turbine as a heating source for the desalination equipment. The method is available where the ratio of water to electric power is usually small, depending upon the performance ratio at the desalinating plant. A rated operation can be maintained for desalination regardless of fluctuations in demand and supply of the power. Fig. 5.7-1 illustrates an account obtainable through combining 100,000 m^3/day desalinating multistage flash evaporating equipment with a 1,000,000 kW power station. This is estimated to work as a system whereby a new and powerful thermal power station is prepared owing to lack of water in a city with population of 1,000,000 inhabitants. Table 5.7-1 is given by estimating the generating cost and the steam cost to supply to desalinating plants under a system like that. From fixing the power cost according to the Power Credit method, power can be generated more from a lower rate of desalinating plant operations, therefore the steam cost is kept cheap. However, since the power cost under the Power Credit method is calculated on the basis of the rate of generating plant operation being at 70%, the profit from operating the generating plant for 310 days is all reduced to the steam, and thus the cost becomes cheap.

(b) Combination of condensing type with back-pressure turbine

This is a method to utilize the steam discharged from the back-pressure turbine as a heating source for the desalinating equipment. The method is usable where the ratio of water to electric power is usually large. Thermal efficiency is best obtainable for the whole plant, and the method is suitable for a plant working mainly for desalination. The Matsushima Coal Mine, Nagasaki Pref. produces 8,000 kW of electric power and 2,650 m^3/day of water by this method. The system is suitable generally for combination with a desalination plant of large capacity. Fig. 5.7-2 shows a low sheet available by the combination of a 500,000 m^3/day desalting plant with an 800,000 kW power station. Table 5.7-2 indicates power costs and steam costs supplied to the desalinating plant which are estimated according to the system, showing a similar tendency to the case of a bleeder system. In any case, cheap costs are obtainable from a lower rate of plant operation, so far as the steam cost is concerned.

(3) IAEA's System for Cost Distribution

In a dual-purpose plant the steam generated by boiler is utilized for both the generating and desalinating plants, and hence the method on how to distribute the costs must be prearranged. LAEA has submitted eight ways. The methods are not necessarily all that are possible, however, the following refers to these systems as best for reference. It is necessary to select the distribution system according to the character and scale of the dual-purpose plant.

(a) Proportional distribution based on annual expenditure of an alternative single-purpose plant

This is a method to distribute gross expenditure of the dual-purpose plant proportionally according to costs at which alternative single-purpose plants for production of water and power in the same quantity as the dual-purpose plant would produce water and power.

(b) Proportional distribution based on generated power

This is a method of distribution based on the ratio of generated power on the assumption that the same steam generating equipment is used at a single-purpose generating plant when distributing expenses for the steam generating equipment to the power generating side and water desalination side out of the gross expenditure for the double-purpose plant. The desalination cost is the sum of the expenses distributed as above and that being caused directly by water desalination.

(c) Proportional distribution based on effective energy

This method is to distribute the annual expenditure common to the dual-purpose plant in proportion to the gross quantity of effective energy used for water desalination and power generation; expenses for common parts like offices and land which have nothing to do with energy consumption must be distributed by an arbitrary method.

The above methods according to proportional distribution are those of distributing gains by arranging a dual-purpose for the generating side and desalination side (though there may be a methodological difference), which may be considered acceptable for both. Then, the credit method given below is that of allowing the gains obtained through the dual-purpose to either of the products preferentially.

(d) Power credit of generating cost for generation-specialized plants of a net output

This is a method to deduct annual expenditure for an alternative single-purpose power station having the same output as the net output of the dual-purpose plant from gross expenditure for the dual-purpose plant and to take the remainder for the desalination cost.

(e) Power credit of generating cost for generation-specialized plants with larger net output

The method is a variant of (d), taking such a case as will set up a generation-specialized plant with the same steam generating equipment instead of a dual-purpose plant.

(f) Power credit of kWh purchasing price by an electric power company

The method is to sell electric output from a dual-purpose plant to an electric power company and to take the remainder available by deducting the earnings from gross expenditure of the dual-purpose plant for the desalination cost.

The desalination cost can be ascertained cheaply by the above methods on credit.

(g) Water credit of desalination costs for an alternative single-purpose desalination plant having equivalent production to the dual-purpose plant.

(h) The Dam, headrace, etc. will be used for application of the 'method of substitution with water credit for the desalination cost which corresponds to an alternative feed water method available for minimizing the cost.

LAEA's report indicates a result obtained through estimating the water and power costs according to these methods which is given in Fig. 5.7-3 and Table 5.7-3. It goes without saying that the desalination cost lowers accordingly in the order of power credit, proportional distribution method and water credit. What is to be noted here is that the water cost fluctuates sharply according to the distribution system employed.

(4) Steam Cost Accounting

(a) Preconditions

i. The plant shall be rated at 1,000,000 kW in generation and 500,000 m^3/day in desalination.

ii. The applicable cost distribution system shall be the proportional distribution method (distribution method a) and power credit method (distribution method d).

iii. Performance ratio of the desalting plant shall be 14 for the proportional distribution method and 12 for the credit method.

iv. Construction costs of the desalting plant shall be 12 billion yen for each of the plants of 12 and 14 in performance ratio, including 420 million yen for the site expense.

v. Fuel cost shall be 30 yen/ℓ and 35 yen/ℓ.

vi. Others are subject to the costing standard for large projects.

(b) Conception and expression for computation

i. Proportional distribution method

In conceiving such an alternative plant for a single purpose as will produce water and power of the same quantity as the dual-purpose plant, gross expenditure of the dual-purpose plant is distributed in accordance with expenses for the alternative plant. Those for which fixed expenses consisting of interest, depreciation, tax and insurance on the desalination side and working expenses necessary for operation, servicing and replacement and cost of chemical supplies are deducted from expenses to be born by the desalination side are expenses of power and steam necessary for desalination.

To determine the steam cost, a distribution of the power and the steam must be antecedent, which is presumed as follows. The power cost is determined by dividing expenses to be born by the generating side by net power output, and those for which the necessary quantity of electricity is multiplied are to be expenses necessary for power. Thus expenses necessary for the steam are obtainable and the steam cost can be determined.

Steam cost $v = \frac{y_w - (a_2 + m_2 + d_2) - P_w \cdot C_e' \cdot H}{E_s}$

Expenses on desalination side $y_w = x_t \cdot \frac{Z_w}{Z_e + Z_w}$

Expenses on generating side $y_e = x_t - y_w$

Power cost $C_e' = y_e / E_e$

Operating hours H = 365 x 24 x rate of utilization

here, x_t: annual expenditure of dual-purpose plant

Z_w: annual expenditure of alternative single-purpose desalination plant

Z_e: annual expenditure of alternative single-purpose generating plant

For other details see figure and table.

ii. Power credit method

The generating cost of a generation-specialized plant with the same output as a dual-purpose plant is determined, and the expense corresponding thereto is born by the generating side with the remainder by the desalination side. The steam cost is obtained through dividing the expense born by the desalination side by the quantity of steam.

Steam cost $v_1 = \frac{x_t - C_e \cdot P_b \cdot H - (a_2 + m_2 + d_2)}{E_s}$

(c) Result of computation

The result is as shown in Table 5.7-4.

5.7.2. Waste Heat Utilization

What is called the production process involves energy consumption. Seawater desalination is not exceptional. In the evaporation process, efficient utilization of energy is expected from arranging the plant to operate for double purposes. However, utilization of waste heat is possible for further energy saving.

A heat source relatively low in temperature, 100°C or so, can be utilized for seawater desalination under the evaporation process, which is convenient for waste heat utilization. A result obtained through investigating the quantity of waste heat in 1973 is cited as shown in Table 5.7-5. The steel industry and electric power plants are most noticeable. After these are ceramics and metal smelting. Burning-up of industrial wastes in note worthy.

Meanwhile, Fig. 5.7-4 shows the temperature range of such waste heat. At a high temperature level is the sensible heat of solid material in the process of iron making manufacture, which is subjected to recovery by way of fluid. At a low temperature level is the cooling water, which is most feasible for use in seawater desalination. Where the sensible heat of exhaust gas is recovered in the form of hot seawater, fresh water is then obtainable through evaporating the hot seawater.

A general waste heat recovery and utilizing process is shown in Fig. 5.7-5. The heat discharged as sensible heat can all be recovered in the form of hot seawater, which is available for use in seawater desalination as a heat source.

The Society of Seawater Science of Japan reports in their research that the said heat is abundant at steel works where the scale of annual production is 6 to 7 million tons, and is capable of working as a heat source for seawater desalination multistage flash evaporating equipment, and thus the following three cases are pointed out:

(Case 1) Utilizing hot drainage at rolling mills

This is to utilize the heat of the skid cooling water of the heating furnace in the hot strip mill, thereby obtaining 2,000 t/h of hot seawater at 43°C.

(Case 2) Utilizing hot drainage of coke oven gas cooler

Hot seawater of 67°C can be obtained at 2,860 t/h from a coke oven gas cooler.

(Case 3) Utilizing sensible heat of low temperature exhaust gas

Hot seawater of 67°C can be obtained at 1,386 t/h through recovering the heat from a coke oven gas, exhaust gas of a hot store and sintering plant by means of a heat exchanger.

A trial design has been made of a layout which will recover waste heat as hot seawater and lead it to a waste heat utilizing seawater desalination plant on the shore. The equipment works on a concrete vessel high flow rate long tube type flash evaporation process, and the process contrived by the Tokyo Industrial Laboratory, which turns a part of the cooling water flowing in the condensation heating tubes into flash brine through a bypass on the way, is applied for utilization of waste heat. Later construction expenses, desalination cost and energy source units are figured out in each case. These are shown in Table 5.7-6.

The desalination cost is 157.1 yen/m^3 obtained through utilizing the hot drainage (Case 2) of a coke oven gas cooler having a comparatively large capacity of 14,000 m^3/day at 67°C hot seawater temperature. This cost is nearly the same as that of 153 yen/m^3 obtained by standard type flash evaporating equipment of 100,000 m^3/day in capacity with a maximum brine temperature of 120°C which was designed on trial in accordance with the positive results of the large project. In taking up energy consumption, the unit for which the energy required for producing 1 m^3 of fresh water is converted to fuel oil (ℓ) is given at 1.4 ℓ/m^3 by using an ejector for the vacuum system and 0.45 ℓ/m^3 by using a vacuum pump driven by motor as in Case 2 and Case 3. Close examination will have to be made for an excessive difference as above. However, the energy unit converted to fuel oil for double purposes is taken at 2.44 ℓ/m^3 through operation of standard multistage flash evaporating equipment. Therefore the unit might save a little under 20% by using waste heat utilization of 0.45 ℓ/m^3 thus saving energy exceptionally. It is said that even the back penetration process, which is a possible energy saving seawater desalination process is available only at 1.7 ℓ/m^3 or so in energy unit, and hence seawater desalination utilizing waste heat must be an attractive subject for examination.

Besides, a single-stage flash evaporation process for desalination of seawater which uses the heat of cooling water at power stations or various industrial plants, including oil refineries, or of natural hot water, was suggested fifty years ago. Tests were carried out carefully by G. Claude, a Frenchman, and his cooperator in the 1920's through 1930's on how to utilize the temperature difference of seawater between the bottom and surface. Further, design and tests were performed by Energie des Mers, a French corporation, for the construction of a generating/desalination plant utilizing natural temperature differences at Abidjan, West Africa under the support of the government and an electric power company after World War II.

Meanwhile, in the United States, E.D. Howe has proven the possibility of operating single-stage flash evaporating equipment by utilizing a temperature difference of 8.3°C at California State College. And Friscom-Russel also carried out a test on low temperature single-stage flash evaporating equipment and concluded that it can be put to practical use, presenting a report in 1958.

The above idea of using waste heat directly on a single-stage flash evaporation is attractive. However, a method to get a further temperature difference of 2.8 - 5.5°C by combining it with solar heat has also been suggested. Utilization of the waste heat of an atomic power station and a large diesel engine is then taken into consideration.

Recently new schemes have been submitted on seawater desalination by waste heat utilizing low temperature vacuum evaporation to cope with soaring energy costs.

5.7.3. Conception of Kombinat

The actual idea a of kombinat is for correlative enterprises turning out many intermediate products, which were first concentrated around the petrochemical industry, for rationalization of the process. In Japan a geographical position for industry would have to be secured, as planned in the past, when the economy had higher growth, and assurance made doubly sure by securing an apartment-house complex, fishery compensation, and traffic facilities in advance of construction of the kombinat. Furthermore, a kombinat to ensure both agriculture and industry and pollution-free atmosphere has been advocated so far.

Since production and supply of water and electric power necessary for a kombinat are considered as being public utilities, the government authorities have repeately taken it into serious examination by conducting a conceptional design on a plant operating for the double purpose of desalination and generation. Main investigations and contents conducted so far are shown schematically in Table 5.7-7.

The "Report on Acceleration of Thermal Power Multi-purpose Utilization" submitted by the National Institute of Resources, Science and Technology Agency in February 1967 looked at a kombinat including a desalination plant and heat supply power station, pointing out for promotion:
1) reflection on city planning, housing planning, geographical position for industry planning, industrial service water planning, water-supply utility planning and power development planning, 2) government subsidy in finance and taxation, 3) arrangement of legislation on heat supply, 4) promotion of utilizing technical development and the arrangement of a technical development structure. The Energy Policy Division, Secretariat for Ministry of International Trade and Industry, has under trial design four types of kombinats of a large light-water reactor

and hot gas furnace including a seawater desalination plant 350,000 - 1,000,000 m^3/day in capacity. Fig. 5.7-6 is a bird's-eye view to show an example. Iron manufacturing, aluminum smelting, soda industry and petrochemical industry are arranged around the atomic reactor, and there is a seawater desalination plant on the sea side having common equipment with waste disposal facilities. This research is for examination of measures for 1) intensification of know-how and scaling-up of equipment, 2) systematization of production management information on markets, circulation, technical skills, etc., 3) diffusion of area development, 4) concentrated management of environmental problems, 5) stable supply of quality and cheap energy.

The Public Utilities Bureau, Ministry of International Trade and Industry, made an examination of synthetic utilization of heat by combining a 1,180 MW light-water reactor, a 1,500,000 m^3/day evaporation process seawater desalination plant and a petrochemical plant coping with 4,500 MW thermal output. Technical problems and economical properties on the conception of system engineering and control were taken into examination through quantitative trials on feasibility. Observation was further conducted on application of facilities for multi-purpose utilization and safety of a model study on how the kombinat would be feasible and, at the same time, economicy was evaluated by applying various methods for cost allocation to distributed thermal energy according to the said model.

5.8. Desalination Cost and Overall Evaluation

5.8.1. Costing Standard

The matter of desalination cost may be based on the measure of technical standards or how much fresh water can be obtained at a particular site.

When constituting the measure of technical standards, the costing standard is so set up as to compare differences in cost by technical skill as distinctively as possible, with geographical position and other conditions like raising of the fund averaged.

The costing standard for desalination was set early by the Office of Saline Water, U.S.A. The said Office giving priority to underground brine instead of seawater for desalination, the evaporation process is not dominant in the standard but various processes including the diaphragm process can be taken for comparison. Then, costing has been done under the electric dialysis process, the reverse osmosis process, the multistage flash distillation process, the vapor compression-vertical tube evaporator-multistage flash distillation process and the vacuum-freezing vapor-compression process in accordance with the desalination costing standard set up in 1970. The results are shown in Table 5.8-1. The capital costs are itemized in 5, and the operation and maintenance costs are itemized for labor costs, general and administrative costs, maintenance costs, chemical cost, electric power costs, etc. The value of desalination cost per m^3 is that reported in 1970 and hence appreciably different from the present cost. As compared in relative value, the process using diaphragms, like electric dialysis and reverse osmosis is cheap, but it must be taken into consideration that the salt content of the brackish water is 1/10 or so of seawater. In the evaporation method, a complicated process is lower in price. The cost is considered to be one of the available data with which to look back upon the past course when a priority was given to the reverse osmosis process and the combination process of falling film vertical tube multi-effect and multistage flash distillation for technical development in America since then.

In Japan seawater desalination costing standards have been set up according to the multistage flash evaporation method for which Agency of Industrial Science and Technology, Ministry of International Trade and Industry set about developing under the system for National Research and Development Program (large-scale project). Given in Table 5.8-2 are standards which were established in April 27, 1977 for evaluation of the results obtained through 8 years' researches.

In the standard for capital cost calculating, the seawater desalination plant and the construction cost covers the main equipment concerned with fundamental design and the site work concerned with geographical position, however, both the factors are arranged into one in this report because it is difficult to generalize for identification. For better clarification of the evaluation on developing technique, it is preferable nevertheless to classify the expenses into main equipment, consisting of the expenses for the evaporator body, auxiliary machinery, shop assembly and material and site work for main equipment foundation, pipin thermal insulation, instrumentation, coating, etc.

Intake/drainage equipment, maintenance equipment and building are then taken into consideration as ancillary equipment. The intake equipment includes an electrolytic chlorine generating device, the intake pump is included in the main equipment, and the building comprises office premises, warehouse and control room.

In the operating cost, a dual-purpose plant available by combining a 1,000,000 kW generating plant and a 500,000 m^3/day (100,000 m^3/day in 5 units) desalination plant is assumed for costing of the heating steam. Cost allocation stands on the basis of the power credit method with the propertional distribution method partly referred to.

For operating personnel concerned with operating labor expense, personne expense per capita is fixed, with data of the Office of Saline Water, U.S.A. working for reference, in consideration of the pay level for local self-governing bodies.

For chemicals, concentrated sulphuric acid (98%) and a defoaming agent are considered. The rate for interest and amortization are fixed according to government bond and publicly subscribed bond and also to durable years and residual value.

For insurance, that against damage for the chemical plant is applied.

Operating days are taken at 310 days per year to cope with the regular inspection term for new and powerful thermal power stations.

A costing for seawater desalination by 100,000 m^3/day multistage flash evaporator according to the costing standard given above is as shown in Table 5.8-3. In comparison with Table 5.8-1, what is largely different as a relative value is energy cost. At first when study for development of seawater desalination was commenced in 1969, the construction expense was taken at 50,000 yen per unit capacity of m^3/day and the desalination cost at the 30 yen level as guidelines. For rough estimation of the cost breakdown, the capital was at 15 yen and the energy cost at 15 yen or so to equal each other. In Table 5.8-3, however, the ratio of capital to energy cost is 1 to 2.5 for the recent desalination cost, because of energy soaring sharply in cost. That the capital is at 140,000 yen per m^3/day, lower than three times 1969's, can be regarded as attaining the target satisfactorily in a sense of curtailing the cost for construction expense.

5.8.2. Overall Evaluation

Since the seawater desalination under the evaporation process involves a large type of technique, influence not only on cost but also on environment and community will have to be taken into consideration. It is necessary to push examinations of the problem as a total system for which counsel will be invited from many people on the policy of how the seawater desalination is to be adapted into cities and industries as a water supply source.

There may be conceived several viewpoints for how to tackle areas of the total system, however, Fig. 5.8-1 gives the situation plan in

7 sectors. Namely, an inevitability lies in economic activity that endorses human life materially, economic activity stimulates correlative industries and thus promotes local activity. It further brings about an increase in water demand to promote technological development of seawater desalination and water reproduction. Technique must be coordinated with consideration for environmental problems, which may work in a direction to suppress economic activities, too.

In such a relation of cause and effect with the system, a plan of the system with an intensive correlativity can be plotted out taking up particular items. For example, Fig. 5.8-2 illustrates a plan of a system as is related with seawater desalination for demand and supply of water, including rivers. The river water is obtainable originally by rain. Water in a necessary quantity is then fed from rivers, for which a method to take in running water from an intake weir through branching directly and a method to use water after storing in a dam, are available. In the case of a dam, stored water can be taken out for use regardless of rainfall, however, everything is left to the rate of flow on the day in the case of an intake weir. Where the seawater desalination plant is to work for such a system, the processed water rises in cost as compared with the river water and the energy consumption inevitably runs high, thus operating the plant only in the dry season. Since the index of drought can be expressed in water level of the dam, the plant operation is to be commanded by adjusting the index from a long-range weather forecast. Now, if the stream condition of a river for water supply against precipitation can be predicted, a simulation will be workable on the seawater desalination plant for the rate of operation. Thus the scale of seawater desalinitio to plan can be obtained through studying the case wherein the plant capacity is changed.

Fig. 5.8-3 is that in which the system is further magnified for better energy saving. As a seawater desalination plant, the system includes standard type multistage flash evaporating equipment, waste heat utilizing evaporating equipment and reverse osmosis equipment, having heat and power supplied from the power station. For the waste heat source, iron works and incinerator are included as representative. Where seawater desalination is intended to cover a lack of river water for such a system, the case may further involve complication. It is therefore necessary to design beforehand such a plant as will work for dual-purposes in combination with a power station so as to keep the desalination plant for temporary operation.

The case where the system is combined with nuclear power generation may invite an argument on technology assessment. Fig. 5.8-4 shows a loop illustrating the cause-and-effect relation thereof. Items of influence upon environment and its countermeasures, contribution to resource energy problem, etc. are correlated.

5.9. Peripheral Technology

5.9.1. Intake/Outfall Equipment

(1) Intake equipment

When intake equipment for raw seawater is designed for desalination, it is necessary to examine the following requirements in view of engineering, execution of work and maintenance:

(a) Water should be ready at all times in the necessary quantity. Care should be taken so as not to cause unavailability of water when required due to daily changes in the tide, drifted sand and sedimentation of suspended solid matter.

(b) Water temperature should be moderately low.

(c) Water properties (physical and chemical) should be superior and stable.

(d) Such factors as will cause corrosion on the intake equipment and cooler should not be allowed.

(e) Avoid a zone which will allow the river water to flow in or the industrial drainage to diffuse.

(f) Take in water with minimum larval content of sticky organisms.

(g) The sea bed shape should be advantageous for stability of the intake equipment and free from damage by typhoons and surge.

(h) The site should be kept from troubles with steamer lanes, harbors, fishery rights, etc.

1) Selection of proper water to take in

The water varies according to the water area but can be classified for examination as follows:

(I)	Surface layer	Layer 2.0 - 3.0 m below the surface of the sea
(II)	Bottom layer	Layer 1.5 - 2.0 m above the sea bed
(III)	Layer covering the area between (I) and (II)	

The surface is subjected largely to influence from the atmosphere and in such a layer the life of marine animals and plants is activated by daily or yearly fluctuations in water temperature, and supply of oxygen or ultra-violet rays and the density of plankton is appreciably high. The surface has much drifting material and suspended matter and is also a zone where river water and industrial drains run out and diffuse in harbors, estuaries or gulfs. The bottom layer is most subject to an influence by the sea bottom nature and has some portions where the water stagnates. In shallow sea the water is agitated by waves in the windy season and thus deteriorates in property. A intermediate in quality constitutes stable layer water and doesn't change extremely in characteristics.

2) Type of sea water intake equipment

The type of seawater intake equipment is classified into a surface intake system or a deep water intake system according to the water layer intended. For the former a usual inlet conduit (open conduit, covered conduit) type is popular, which is generally free from conditions for water temperature, water property, etc. For the latter, a workable technique is being developed as the result of basic hydrological experiments, theoretical analysis and field tests carried out through researches by each laboratory or private enterprise recently to satisfy the high standard of intake conditions.

For the deep water intake system, the following fundamental types are possible:

(i) Curtain wall system

(ii) Submarine intake pipe system
   (a) Horizontal intake type
   (b) Vertical intake type

(i) Curtain wall system ..... This is a system to take in water through an opening between the sea bed and the bottom of a curtain wall which is fixed or suspended in front of the headrace port of the pumping basin. This system is employed extensively for any intake quantity subject to securing a necessary depth (over 7.0 - 8.0 m at water temperature rise; bed being 4.0 m) in anchorage and gulf. But it is not suitable for a zone which is severe in wave and climatic conditions or a shallow coast. For anchorage and gulf, suitable measures will be necessary to cope with the inflow of bottom colloids sediment and the cumulation or inflow of drifting material and jellyfish; construction expense falls in relation as the capacity becomes large. Where the wall bottom end depth is not satisfactorily obtainable from perpendicular mixing being accelerated in the thermocline by wind and waves, the rate of entrainment will have to be expected in the planning.

(ii) Submarine intake pipe system ..... This is a system to take in water through a headrace pipe like a steel pipe, iron pipe or concrete pipe which is laid on the sea bed to a spot where intake is to be done or an arbitrary spot where the deep water intake is available; there are two types classified, horizontal and vertical intake systems, according to the shape of the intake port. Both can be installed relatively free from being restricted by the shore and sea bed configuration at the water intake area.

The intake port is miniaturized three-dimensionally and laid on the sea bed with its top left free, therefore drifting material and jellyfish do not stay and the rate of entrainment can be kept relatinery free from descent of the thermocline. Installed on the sea bed, the intake port structure is secure against wave conditions as compared with the curtain wall system. The bed soil can be prevented from flowing into the intake port by arranging its rise from the sea bed in a hydraulic design according to the rate of flow in the port. Then oscillations of the open sea wave are attenuated in the headrace pipe up to the pumping station, giving trouble for the pumping operation.

3) Flow sheet

A flow sheet for the intake equipment is given in Fig. 5.9-3.

The seawater is led naturally to the pumping station on the land from the vertical intake port through the embedded pipe. The pumping station is a reinforced concrete underground structure and designed hydraulically to have a rectifying effect. The gate and screen equipment are prepared as ancillary equipment. Treatment by chlorine injection is to be conducted through the intake port for suppression of trouble by sticky organisms, for which electrolytic chlorinating equipment will be arranged adjacently to the pumping station. A vertical diagonal flow pump is intended for the seawater pump.

4) Intake pipe and ancillary equipment

(i) Intake pipe ..... A steel pipe, cast iron pipe, concrete pipe and plastic pipe with a circular section are generally used, which are selected in consideration of intake quantity, geological features around the conduit laid, decompression workability and economical property. The steel pipe is more popular in the case of a power station for which seawater is largely used. The flow velocity in the pipe is taken generally at 1.5 - 2.5 m/sec. in consideration of head loss and measures against troubles by sticky organisms.

(ii) Screen equipment ...... A bar screen with a rotary rake and rotary screen which work efficiently for recovery are used according to the kind of drifting materials (vinyl, waste matter, seaweed, jellyfish, etc.) coming in through intake port.

(iii) Stop log ..... Provided for cleaning of the pumping station and maintenance of the pump, screen, etc.

(iv) Seawater intake pump ..... For selection of the type, the following will be taken into consideration:

- Must be a type to cope well with necessary intake quantity and total head and also efficient to operate.
- Limited in area for installation; simple in engineering construction.
- Simple in operation.

Such a type of pump is preferable as is profitable in NPCH against fluctuation of the suction level due to tide fluctuation and other causes.

The vertical diagonal flow pump is suitable as it satisfys the above requirements and has worked at various power stations. It is necessary to give priority to seawater resistance for the material.

(v) Measures against troubles by sticky organisms ..... Chlorinating treatment is most popular for measures against choking trouble, deterioration in heating efficiency, etc. which are resultant from seaweed, shellfish or other organisms sticking to and growing on the piping and equipment.

For this device a process of injecting hypochlorite produced by direct electrolysis of the seawater is perfect and easy in operation. The injection quantity must be selected according to the water property and objective organism; chlorine residual has to do with corrosion on the equipment material and is also capable of exerting an influence on the organisms living in the water area in which it is discharged. Care should be taken accordingly.

(vi) Corrosion preventive measures ..... For corrosion preventive measures on the steel material used for intake equipment, there is employed a method to make corrosion allowances on the raw material and carry out coating with a paint of the tar epoxy resin group and cathodic protection in combination.

(2) Discharge equipment

A large quantity of brine and hot drainage are discharged from the seawater desalination equipment working under the evaporation process. As the rises of water temperature and salt density are estimated at 7°C and 5 - 10 ‰ respectively against the raw seawater taken in, the equipment having a hydraulically designed configuration with which to lighten the water area environment will have to be examined for the following conditions:

(i) To mix the discharged drain with ambient seawater quickly for dilution.

(ii) The flow velocity in the sea is not to hinder sailing vessels due to the discharge.

(iii) To minimize the influence to be exerted on the ecological groups in the diffused area.

(iv) The discharge is not to cause a change in the shore shape or a scour in the sea bed.

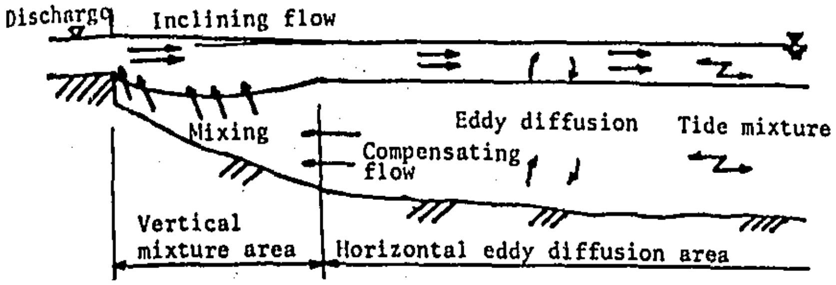
(v) The shore mode and exclusive use of public sea surface are kept at a minimum.

(vi) The discharged drainage and the intake are not to be recirculated.

a) Discharge system and characteristics

The method of drainage into the water area is classified roughly into surface discharge and submerged discharge.

(i) Surface discharge ..... The layer at 0 - 3 m is generally taken for the depth through which drain is released directly in the water area from the sea wall, etc. Drainage is generally lighter in density than the seawater and develops to a density flow rising to the surface through its buoyancy. In the neighborhood of the discharge outlet, the lower layer water is entrained therein by a large shearing stress to dilute it. After passing the said area, the layer of drainage is stabilized to a density flow, therefore the upper layer of drainage is diffused and diluted by eddy diffusion and mixing of the tide, but entrainment from the lower layer cannot be expected. Features of the surface discharge are as follows:



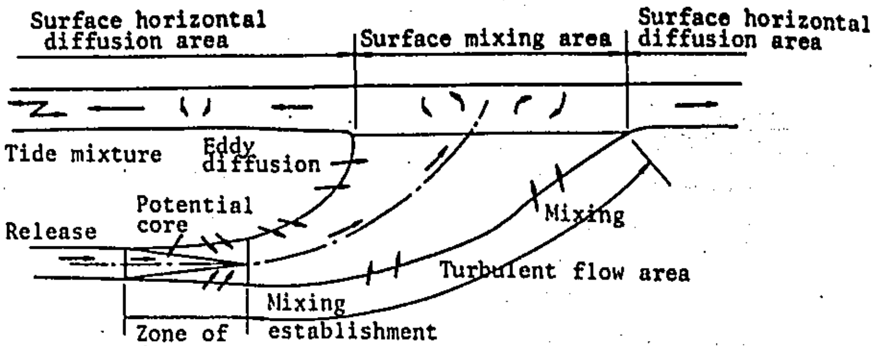
- Direct discharge through an open channel from the sea wall has led to many satisfactory achievements.
- Since the velocity of surface flow is subjected to an influence of that of discharged flow, the discharge velocity must be moderated accordingly. Therefore, a large quantity of the discharge flow must be coordinated with widening the discharge port.
- The discharge velocity cannot be so large, and hence initial mixing cannot be expected to be high.
- Dilution occurs mostly in the horizontal eddy diffusion area. Almost not diluted by entrainment from the lower layer this area is easy to widen in the diffusion range.

(ii) Submerged discharge ..... This is a system to dilute the drainage discharged through a port arranged near the sea bed by allowing the ambient lower layer to entrain therein while it reaches the surface by buoyancy. This system is further subdivided into the following:

- Horizontal discharge pipe system
- Vertical discharge pipe system
- Multi-nozzle system

The drainage coming out of the release port form a plume, reaches the surface and then becomes a density flow to diffuse on the surface. The process can be divided into the following four areas:

- Zone of flow establishment ..... The area is provided with a core holding the rate of flow and the density at the release outlet port; mixing is conducted by shearing stress at its periphery.



- Turbulent flow area ..... The turbulent flow develops completely and the radial distribution of rate of flow and density presents nearly Gaussia distribution. The central axis of the plume change the angle slowly upward by buoyancy and is subject to extensive mixing.
- Surface mixing area ..... This is an area covering the surface where the plume has arrived with mixing activated internally but free from outside. Diluti cannot be expected in consequence.
- Surface horizontal diffusion area ..... Upon arriva the plume diffuses of the surface vortically and i subjected to tide mixing, however, incoming of low bed water is limited. It works as in the case of the horizontal eddy diffusion area of the surface discharge.

The following can be given as features of the submarine release:

- The flow velocity slows appreciably the process of the plume rising. Therefore, the discharge veloci can be made comparatively high under restriction o the surface flow arising from release.
- That the discharge velocity can be quickened is possible to minimize the construction of the discharge port.
- That the discharge velocity can be quickened is available to allowing the peripheral water to entr more.
- The turbulent flow area continues as far as the surface, allowing the peripheral water to mix in securely during the course. Each type of the submarine release is shown in Fig. 5.9-4.

(iii) Flow of discharge equipment ..... Brine and cooled drainage discharged from the seawater desalination equipment are introduced into a mixing tank on the shore through a conduit. After being mixed thoroughly in the mixing tank, the drainage is jetted through a nozzle arranged at the discharge spot by way of a drainage conduit. The jetting speed is determined according to the fluidity, depth, bed nature and biological environment of the water basin, however, it is normally 1.5 - 8.0 m/sec. horizontally. After reaching the surface of the sea, the plume diffuses horizontally. However, the design is such that the velocity of flow on the surface will be limited to 25 - 50 cm/sec.

5.9.2. Seawater Pollution and Measures

Pollution of castal seawater has grown on a worldwide scale or late, and hence it is necessary to examine polluted seawater as used for desalination. When using polluted seawater for desalination by the evaporation process, the following is to describe (1) behavior of the pollutant or, particularly, mixing of the pollutant in produced fresh water, (2) microbial corrosion, and (3) removal of the pollutant.

(1) Behavior of pollutant

Where raw seawater polluted with city sewage or industrial waste is supplied to the desalination equipment, the pollutant in the seawater is capable of being mixed into the produced fresh water, and hence its behavior must be grasped thoroughly. A description will be given here on the three items: (a) vapor-liquid equilibrium relation of the pollutant, (b) additive test on multistage flash evaporating equipment, and (c) conjecture on behavior of the pollutant.

(a) Vapor-liquid equilibrium relation of the pollutant

(i) Ammonia

Ammonia comes from a chemical change in animals, plants, city sewage, industrial waste. There is a zone in which ammonia comes up to 1 - 2 ppm in the innermost part of Tokyo Bay. The vapor-liquid equilibrium relation between ammonia and water is largely influenced by the pH of the solution: as shown in Fig. 5.9-5, regardless of liquid composition x in a system with seawater, the connectration of ammonia y comes to zero at around pH 6 to keep ammonia from being concentrated in the produced fresh water. However, the vapor-liquid equilibrium ratio (K = y/x) is 1 at around pH 6.8 and more than 6 at around pH 8. The influence of salt content in the seawater and the system pressure is small as compared with pH. In the case of a weak electrolyte like ammonia, the pH of the solution generally exerts a big influence on the vapor-liquid equilibrium relation, which is due to pH having to do directly with ionic equilibrium. In the case of ammonia, the equilibrium relation is given as:

$$NH_3 + H_2O \rightleftharpoons NH_4^+ + OH^-$$

and thus OH^- ions increase as the pH of the solution rises, the ammonia molecules increasing through reaction and shifting to the left side. Ammonia in the vapor phase also increases as a consequence.

(ii) Nitrous acid

Nitrous acid is capable of existing in a reducing environment and increases from pollution rising, as in the case of ammonia. However, nitrous acid is under the following equilibrium in the solution:

$$NO_2^- + H^+ \rightleftharpoons HNO_2$$

reaction comes to the left side in the area where pH is high, presenting an ionic state influence, and nitrous acid is not concentrated in the vapor phase. Fig. 5.9-6 shows at the same time an influence of pH on nitrous acid against the vapor-liquid equilibrium ratio and a dissociation equilibrium curve at 25°C. SW given in the figure is a dissociation curve corrected by means of the Killand equation due to the seawater salt exerting an influence on ion activity.

(iii) Cyanogen

From its poisonous character, cyanogen must not be detected at all according to the standards of the Water-service Law. However, the vapor-liquid equilibrium ratio of cyanogen is extremely large and is about 20 at pH 6 - 8, and hence there may arise an unexpected case where it is concentrated in the produced fresh water, should much cyanogen be contained in the raw seawater. Where, however, concentration of cyanogen is below 0.2 ppm, the vapor-liquid equilibrium ratio becomes almost zero, perhaps from its forming a complex with metal ions in the seawater. Transfer to the produced fresh water will thus hardly occur.

(iv) Hydrogen sulfide

It is often the case that mud and water are black in a water basin where pollution is exceedingly advanced, and this diffuses hydrogen sulfide in the air in summertime to cause offensive smells and corrosion on metal. Hydrogen sulfide is present in solution in S^{2-}

at a pH more than 8. However, it turns to H_2S gas in the area below pH 8, therefore H_2S hardly transfers to the produced fresh water in the flash evaporating equipment and is taken off as non-condensable gas through the vent line. Meanwhile, raw seawater is mixed with sulfuric acid up to pH 4 or so in the decarbonation pretreatment, therefore H_2S will largely be removed simultaneously with decarbonation.

(v) Phenol

It is known that substances of the phenol group are present in traces in natural seawater; it is one of the typical pollutants in the chemical industry area and thus prohibited from exceeding 0.05 ppm by the standards of the Water-service Law, and further is a cause of offensive smells. Phenol forms a minimum azeotropic mixture together with water to highly increase its vapor pressure, therefore it is capable of being concentrated in the produced fresh water if it is mixed in the raw seawater. Since phenol, O-cresol, m-cresol, thymol and α-naphtol are small in dissociation in aqueous solutions, influence of pH on the vapor-liquid equilibrium ratio is minimized. However, an influence of salinity has been conspicuous through a salting-out effect. Fig. 5.9-7 represents the influence of salinity on the vapor-liquid equilibrium ratio of each substance, whereby it is found that those other than α-naphthol below 1 in K value will be concentrated in the produced fresh water.

(vi) Alcohol

The vapor-liquid equilibrium ratios of alcohols such as methanol, ethanol, n-propanol, i-propanol, n-butanol, i-butanol and i-amyl alcohol are appreciably large and also subjected to the salting-out effect. Therefore these are capable of being concentrated in the produced fresh water.

(vii) Organic acid and organic alkali

Organic acids like formic acid or acetic acid are almost in ionic form at around pH 7 in desalination equipment under the evaporation process and hence do not transfer to the produced fresh water. On the other hand, organic alkalis like pyridine or aniline are appreciably large in the vapor-liquid equilibrium ratio, and therefore may be concentrated and transferred accordingly.

(viii) Ester

Ester phthalate and n-butyl acetate are used extensively as intermediates for the chemical industry and are also available for use on plastic as a phasticizer. These are subjected to an influence of change in pH of the solution by hydrolysis. However, they are capable of being concentrated because of being large in the vapor-liquid equilibrium ratio.

(ix) Others

Alkylbenzenesulfonic acid (ABS), which is a principal component of neutral detergent, is not totally volatile and hence does not transfer to the produced fresh water. Fuel oil is generally not soluble in water and floats on the surface of the sea. Therefore it does mix in raw seawater. However, where it has mixed for some reason or other, almost all of the light oil component transfers into the produced fresh water through steam distillation.

(b) Additive test on multistage flash evaporating equipment

Summary and result of an additive test of ammonia and phenol, which are typical pollutants, carried out on the 3,000 m^3/day multistage flash evaporating equipment in the Chigasaki Test Facility are given as follows:

Operating system	Brine recirculation
Desalinating capacity	3,000 m^3/day
Performance ratio	11
Number of stages on heat recovery section	36
Number of stages on heat rejection section	3
Max. temperature of circulating brine	121°C
Concentration ratio of circulating brine	1.54
Concentration ratio of discharged brine	1.78

Result of the ammonia additive test is shown in Table 5.9-1 and Fig. 5.9-8. In the circulating brine of pH 7.0 - 7.3 the ammonia concentration ratio between raw seawater and produced fresh water is 1.5, which corresponds nearly to the vapor-liquid equilibrium ratio at 1.4 in pH 7 seawater.

In the phenol additive test, the rate of distillation the produced fresh water is thymol > O-cresol > m-cresol > phenol in that order, which also corresponds to the vapor-liquid equilibrium ratio. The result of the phenol additive test is given in Fig. 5.9-9.

(c) Conjecture on behavior of pollutants

It is impossible to carry out an additive test on each pollutant in seawater for behavior in desalination equipment. However, their behavior and distribution must be known before considering a counterplan for elimination and treatment of the pollutants. A description will be given with reference to computer simulation methods according to a numerical value model which is based on the vapor-liquid equilibrium data.

In referring to Fig. 5.9-10, assuming X is the concentration of a pollutant in raw seawater, x the concentration in the circulating brine, y the concentration in the produced fresh water, k the vapor-liquid equilibrium ratio, B the discharged brine quantity, D the produced fresh water quantity, H the supplied seawater quantity, L the circulating brine quantity, subscript i the number of stages, the concentration of pollutants in the brine flowing into the vessel at the first cycle is

$$x_0^{(1)} = \frac{HX}{L_0} \tag{1}$$

the superscript (1) representing number of cycles.

From the vapor-liquid equilibrium relation at the first stage and mass balance:

$$y_1^{(1)} = K^{(1)} \cdot x_1^{(1)} \tag{2}$$

$$L_0 x_0^{(1)} = L_1 x_1^{(1)} + D_1 y_1^{(1)} = L_1 x_1^{(1)} + D_1 K_1^{(1)} x_1^{(1)} \tag{3}$$

In transforming the eq. (3)

$$x_1^{(1)} = \frac{L_0 x_0^{(1)}}{L_1 + D_1 K_1^{(1)}} \tag{4}$$

$y_1^{(1)}$ can be obtained through calculating the $x_1^{(1)}$ and substituting it for the eq. (2).

At i stage,

$$x_i^{(1)} = \frac{L_{i-1}\ x_{i-1}^{(1)}}{L_i + D_i K_i^{(1)}} \qquad (5)$$

and at i stage at P time,

$$x_i^{(P)} = \frac{L_{i-1} \cdot x_{i-1}^{(P)}}{L_i + D_i K_i^{(P)}} \qquad (6)$$

therefore the concentration of pollutants in all the produced fresh water at the cycle of P time is

$$y_T^{(P)} = \frac{\sum_{i=1}^{n} D_i y_i^{(P)}}{D} \qquad (7)$$

For further details on the above calculation, refer to the literature (11).

If $x_i^{(P)}$ and $y_T^{(P)}$ converge in the above calculation, then behavior and distribution of the pollutants in the equipment in a steady state can be obtained. Transient phenomenon can also be conjectured by changing the input data properly. However, the vapor-liquid equilibrium ratio K must be formulated with pH, temperature (pressure) and salinity working as parameters.

A comparison of the results of an additive test on phenol with that of calculation is as shown in Table 5.9-2.

(2) Microbial corrosion

Where the corrosion of construction material by polluted seawater is problematical, the behavior of sulfate reducing bacteria must be taken into consideration. There is a volume of sulfate ions contained in seawater, and if an organic pollutant flows in it, sulfate reducing bacteria multiply to reduce sulfate ions to hydrogen sulfide, and then the hydrogen sulfide is discharged. The fact is that this hydrogen sulfide accelerates the corrosion of metal. Now, growing conditions of the bacteria will be taken into consideration firstly. An influence to be exerted on corrosion of iron and copper will be described next and then preventive measures will be conceived.

Sulfate reducing bacteria are anaerobic and hence grow where there is no oxygen present. Optimal pH is 7 or so and temperature is 25 - 35°C. However, high temperature bacteria growing at 50 - 60°C are also found. As nutritive sources for micro-organisms to grow, (a) carbon sources, (b) nitrogen sources, (c) inorganic salts and (d) growth factors (like vitamins) constitute energy sources and a living body will be necessary. The energy metabolism of sulfate reducing bacteria is peculiar as it performs so-called sulfate respiration by means of sulfate ions instead of oxygen. That is, hydrogen sulfide is generated by energy which is obtained through oxidation of organic acids limited to lactic acid, pyruvic acid, formic acid, etc. or pydrogen under the following formula:

$$2CH_3CHOHCOOH + SO_4^{2-} \longrightarrow 2CH_3COOH + H_2S + 2CO_2 + 2OH^-$$

or

$$5H_2 + SO_4^{2-} \longrightarrow H_2S + 4H_2O$$

Organic acids or hydrogen which develop to an energy source can easily be produced by coexistent micro-organisms through an organic matter decomposing process. And a carbon source or nitrogen source which constitute the living body multiply the bacteria faster when they are organic. Therefore, sulfate reducing bacteria will find a habitable place where organic matter and sulfate ions are abundant-both with no oxygen present. For example, in the bottom sediment of polluted coastal water basins or in the sludge deposited in desalination equipment using polluted seawater. Hydrogen sulfide is generated actively in bottom sediment in the summer where water temperature becomes particularly optimal.

The matter of how behavior of sulfate reducing bacteria is correlated with corrosion of metal is taken up for examination first with reference to iron. Corrosion of iron in a neutral aqueous solution under oxygen-free conditions takes place electrochemically in most cases and is conceived as:

$$Fe \longrightarrow Fe^{2+} + 2e^-$$

in anodic reaction and

$$2H^+ + 2e^- \longrightarrow H_2$$

in cathdic reaction. For sulfate reducing bacteria, hydrogen is an energy source, and iron ions are an essential inorganic salt and further function to fix hydrogen sulfide excreted as iron sulfide. Therefore corrosion on iron gives the bacteria preferable growing conditions. Also, an assertion was made in the past with reference to corrosion of iron, that corrosion is accelerated because bacteria take off hydrogen. However, the assertion that the generated hydrogen sulfide functions for acceleration might be correct according to a test result obtained by the writers. Then, corrosion of iron by hydrogen sulfide is accompanied with formation of a sulfide film, which makes the phenomenon complicated. Fig. 5.9-11 shows a relation between the concentration of hydrogen sulfide kept constant in artificial seawater and the corrosion rate of mild steel. The initial corrosion rate where the sulfide film has not yet been formed simply depends on the concentration of hydrogen sulfide. However, dependability on concentration is complicated for a steady corrosion rate. The film is not formed in a low-concentration zone and the same dependability as the initial stage is shown. A corrosion resistant film is formed in a medium-concentration zone and the corrosion rate decreases accordingly. In a high-concentration zone the film is formed but inferior in corrosion resistance and shows the same dependability on concentration as the initial stage. However, in the case of corrosion arising practically, the concentration of hydrogen sulfide fluctuates to make the phenomenon further complicated. In addition, there may arise galvanic corrosion of the film and the base metal even after hydrogen sulfide has gone.

Meanwhile, hydrogen sulfide is also influential on the corrosion of copper. In this case corrosion accelerating action of the complex salt forming agent like ammonia which is a metabolic product is conceivable at first. Then conversely a bactericidal action of copper ion is conceivable also. However, dissolved copper ions are fixed as copper sulfide in such areas where hydrogen sulfide has been generated, and hence such an action is not powerful. In the case of corrosion of copper by hydrogen sulfide, the sulfide film is inferior in corrosion resistance and has no concentration zone where corrosion resistant film is produced, like iron. According to the writers' test, however, in case corrosion resistant oxide film was formed before coming in contact with hydrogen sulfide, corrosion did not advance any further even after hydrogen sulfide was incoming. In addition, the fact is that a method to form corrosion resistant film on heat transfer tubes of the copper group for condensers practically through injection of iron ions into seawater has been effective, and hence surface conditions in which hydrogen sulfide comes in contact may exert an influence upon the corrosion rate in the case of corrosion on copper.

From what has been described above, measures against polluted seawater will be taken up on heat transfer tube material. It is preferable to use such material as is free from corrosion by hydrogen sulfide, like titanium, in the heat rejection section allowing the raw seawater to pass. However, in the heat recovery section hydrogen sulfide in the seawater disperses in the pretreatment stage, and hence if the prevention of hydrogen sulfide generation in the heat transfer tube is possible through applying measures against sludge, copper alloy with corrosion resistant film formed on it through an iron ion injection process may be used. Organic matter and bottom sediment are not to be allowed, where possible, into the equipment, for which it is conceivable to set up a settling tank or filter at the supplied seawater inlet port. It is then necessary to keep the intake water under supervision at all times for quality.

(3) Elimination of pollutants

(a) Fundamental test for ammonia adsorption by zeolite

Ammonia is one of the pollutants that come in fresh water obtainable through treating the polluted seawater under the evaporation proces. There are 1 zeolite adsorption method, 2 ion-exchange resin method, 3 chlorination —active carbon adsorption method, 4 stripping method, 5 biochemical decomposition method, etc. available for elimination of ammonia. However, the stripping method and biochemical decomposition method are excepted, particularly when ammonia mixed in fresh water produced under the multistage flash process is estimated at a minute amount. In addition, data necessary for design are ready, except the zeolite adsorption method. Therefore the zeolite adsorption method is specially taken up for study, and then the rate equation has been obtained through fundamental study on the ammonia adsorption rate by natural zeolite of the clinoputylolite group and its mechanism.

Next, an ammonia eliminating device to treat 100,000 m^3/day fresh water was designed for various methods according to the result above and literature on other elimination methods, and a comparison was made between the features and economy.

a. Fundamental study on ammonia elimination by natural zeolite

An ion exchange reaction was ensured from examining the reaction form of zeolite with an ammonia aqueous solution at the fundamental test under the batch method. Now, therefore, the formula of ion exchange reaction was assumed as follows:

$$\frac{dX_A}{dt} = k_1 C_{AO} (1 - X_A)(L - X_A) - k_2 C_{AO} X_A^2 \qquad (1)$$

where, X_A: amount of change in concentration of ammonia in the solution, C_A: concentration of ammonia in the solution, C_{AO}: initial concentration in the solution, C_{AR}: equilibrium concentration in the solution, k_1, k_2: reaction rate constants, L: initial concentration ratio of zeolite to ammonia.

The value obtained through calculation of the formula in integration was found to be coincident with the experiment value. Further, the ion exchange capacity of zeolite was 1.47 meq./g and selection coefficient K NH_4/Na was 6.1. It was then observed that an adsorption of copper took place under the presence of ammonia, and thus an exchange of other heavy metals is conceivable.

Under the column experiment, meanwhile, the height (Za) of an adsorption zone in which zeolite was fixed was obtained through the following formula by means of crushed particles of 1 mm or so and the break through curve of ammonia.

$$Za = \frac{2(\theta_m - \theta_D)}{\theta_m} Z \qquad (2)$$

where, θ_D: time to reach $C/C_D = 0.05$ on the transmission curve, θ_m: time to reach $C/C_D = 50$, Z: height of zeolite charged bed of the column. The result under each condition is as shown in Fig. 5.9-12.

As will be apparent from the figure, the height of the adsorption zone is kept low irrespective of conditions, which is indicative of the charging height not being required to have special consideration taken for design of the equipment to cope with an increased capacity of 100,000 m^3/day.

Zeolite can be regenerated then repeatedly since not only NH_4^+ but also heavy metals are eliminated by means of 1 mol of sodium chloride.

b. Treating process

Since the ammonia content in the produced fresh water is regarded as minute in amount, the zeolite adsorption process, ion exchange resin process and chlorination-active carbon adsorption process will be suitable for elimination.

Zeolite adsorption process

This method is applied, in the majority of cases, to elimination of ammonia for sewerage in America. Where coexistent salt is less contained as in the case of produced fresh water, a large capacity for ammonia adsorption (15 g as NH_4/ℓ-Z) is obtainable. A basic flow sheet is as given in Fig. 5.9-13. After saturation with ammonia, zeolite can be regenerated by means of caustic soda or sodium chloride. Hydrochloric acid in the flow sheet works for acid pickling of zeolite. Drainage coming after reverse washing of the zeolite adsorption tower and pickling is discharged through neutralization water tank and a flocculation settling tank. This process is not available for elimination of oil content but a part of heavy metals can be eliminated.

Capacity of the treating equipment was 100,000 m^3/day, for which the total amount of investment was 412,395,000 yen (site area being 976.5 m^2) and total operating cost 4.26 yen/m^3 (including equipment depreciation).

Ion exchange resin process

Selective adsorption of weak acid cation exchange resin is $Ca^{2+} > NH_4 > Na^+$, and NH_4, bivalent ion and ionic heavy metals are eliminated. Brine of the multistage flash equipment can be used as a regenerant. Assuming the worst quality of produced fresh water to be 1,000 times that of the seawater, treated water more than 4,000 times per resin unit volume can be obtained. Disposition of the regenerated drainage is carried out only in a neutralization water tank (Fig. 5.9-14). This process is not available for elimination of phenol and oil content but a part of heavy metals can be eliminated. For equipment of 100,000 m^3/day in capacity, the total amount of investment was 532,500,000 yen (site area being 1,052 m^2) and total operating cost was 4.65 yen/m^3 (including equipment depreciation).

Chlorination-active carbon adsorption process

Reaction for decomposition of ammonia by chlorination is as follows:

$$NH_4^+ + HOCl = NH_2Cl + H_2O + H^+$$

$$2NH_2Cl + HOCl = N_2 + 3HCl + H_2O$$

An examination of how to eliminate ammonia from the sewerage is also under way after the study on a pre-chlorination method for Water works. There is a report that necessary chlorine becomes less in quantity from combination with the active carbon adsorption process. Phenol, oil content and a minute amount of heavy metal will be pollutants of the produced fresh water, therefore this process is most popular. The necessary quanti of chlorine will be 5 to 10 times that of ammonia. For equipment of 100,000 m^3/day in capacity, the total amoun of investment was 1,091,050,000 yen (site area being 1,935 m^2) and total operating cost was 9.30 yen/m^3 (including equipment depreciation). A flow sheet is shown in Fig. 5.9-15.

5.9.3. Management of Product Water

(1) Quality of product water

Water obtained through the evaporation process is distilled water and hence is extremely pure. A normal plant ensures total dissolved solids (TDS) in the product water at 50 - 100 ppm, however, rather less than the above TDS can easily be obtained through normal operation.

Purity of the product water is governed chiefly by the rate of carryover in the evaporation chamber. Therefore, the main impurities in the product water may be regarded as coming from the make-up seawater. Besides this, iron, copper, zinc, etc., which result from metal materials of the plant, are detected in minute amounts.

Where polluted seawater is supplied, volatile matter contained in it is capable of being mixed in the product water. A description of this has been given in paragraph 5.9-2.

In any case the fresh water obtained through the evaporation process is of a quality to exceed by far the standards for city water (see data given at the end). However, for the purity required for water supplied to the boiler, the product water will have to be purified further.

In Table 5.9-3 is an analysis of the product water at a practically working plant.

(2) Requirement for stabilization of product water

As described in the foregoing paragraph, TDS of the product water by the evaporation process is low in comparison with river water. To minimize corrosion at the plant further, the concentration of oxygen dissolved in the evaporating brine is decreased to several or several ten ppb. Concentration of oxygen dissolved in the product water is kept accordingly low. The product water with such a low TDS and dissolved oxygen concentration is called "hungry water", and it is known that such water fed straight to the water works causes various troubles. It is normal that pure water doesn't corrode metal materials much. In the majority of cases, however, the corrosion resistance of metal depends on the oxidized film produced on its surface. The oxidized film is prevented from growing in water with dissolved oxygen extremely low in it and thus corrosion advances all the more. Then, water with too low TDS is kept in an unstable condition and stabilizes itself by dissolving the material coming in contact with it.

Such "hungry water", as mentioned, works to corrode metal materia and concrete structures, and hence treatment for stabilization wil have to be considered. So-called "red water" arises from iron having been dissolved in the product water through negligence in such treatment.

Stabilization of the hungry water will be secured by increasing hardness of the product water to suppress an inclination to dissolve materials, and then adding dissolved oxygen. It is more effective to decrease corrosiveness by keeping pH somewhat alkali For increase in hardness, a method to accelerate the concentration of calcium carbonate is generally employed. This method is to utilize the buffer action of carbonate ions.

When left as it stands in the atmosphere, pure water allows carbo dioxide gas to dissolve in it and lower pH. However, where calci carbonate has been dissolved in the water beforehand, carbonate ions turn to bicarbonate ions through the following reaction, thus avoiding a decrease in pH:

$$H^+ + CO_3^{2-} \longrightarrow HCO_3^-$$

For such stabilization of the product water through buffer action by calcium carbonate, a method to saturate the product water with calcium carbonate or to substitute magnesium for a part of the calcium is employed. Solubility of calcium carbonate changes according to temperature and pH. Fig. 5.9-16 shows a relation of solubility of calcium carbonate with pH in the temperature range 0 - 40°C. As will be apparent from the figure, the lower pH of the product water is, the higher solubility of calcium carbonate increases. This is due to dissolving the calcium, which is present in the forms of both calcium carbonate and calcium bicarbonate, actually being expressed as calcium carbonate and the ratio of calcium bicarbonate increasing which is high in solubility in accordance with pH being low.

Concentration of $CaCO_3$, a neutral calcium carbonate which is soluble in water free of carbon dioxide gas is about 12 ppm, but the solubility becomes 53 ppm for calcium bicarbonate in water equilibriated with the atmosphere.

Shown in Fig. 5.9-16 is a concentration of calcium carbonate saturated according to the pH of the product water. In case, for example, the temperature is 20°C and the pH of the product water is 8, solubility of calcium carbonate (alkalinity) is about 80 ppm. Where alkalinity is 80 ppm, on the other hand, and the pH of the product water is 7.0, the water cannot be saturated (stabilized) thoroughly. A difference between the actual pH of of the water in which calcium carbonate is dissolved and the pH at which that containing calcium carbonate is saturated is generally called the Langelier Index. The index is 7.0 - 8.0 = -1.0 in the above case; the water may be unstable proportionally as the Langelier Index is negative. To stabilize water by calcium carbonate is to keep the Langelier Index at zero.

If magnesium is substituted for a part of the calcium, the Langelier Index will be obtainable through combining calcium with magnesium. Where calcium and magnesium are present in equivalent mol, the alkalinity increases 20 - 30% more than that of Fig. 5.9-16. However, concentration of calcium is high in most cases, and nothing will diverge sharply from the figure. Where the pH of the product water is 8, the alkalinity will be kept normally at 80 ppm (as $CaCO_3$).

It is said that the concentration of dissolved oxygen necessary for an anticorrosive oxidized film to grow on the surface of metal materials is about 5 ppm. Then, the existence of dissolved oxygen is effective for suppression of the growth of anaerobic bacteria which make the water smelly. It is reported that the coexistence of dissolved oxygen and calcium carbonate is conducive to producing an anticorrosive membrane of calcium carbonate-metal oxide on the surface of iron, galvanized steel sheet and copper pipe in the pH range 7.8 - 9.0.

(3) Treatment of stabilization of product water

The stabilizing treatment described in the foregoing paragraph is harmful for water supplied to the boiler. However, it is necessary to keep the Langelier Index at zero and dissolved oxygen at least at 5 ppm for drinking water.

(a) Aeration

Water (about 8 ppm) saturated in the air will be mixed in the product water or the product water will be left for aeration directly in the spray tower or weir. There is also a method to allow the air into the piping for product water. Aeration involves no serious problem.

(b) Stabilization by calcium carbonate

(i) Mixing of natural water

This is a method to increase pH by means of suitable alkali (lime being normally used) after mixing river water of high hardness or brine, where obtainable, in the product water. It is suitable for a plant small in scale, but care should be taken for corrosiveness and health in the case of brine, which allows chloride ions to mix in together. It is generally preferable that chloride ions be kept below 250 ppm.

(ii) Absorption of carbon dioxide gas in lime solution

This is a method to obtain calcium bicarbonate through absorption of carbon dioxide gas in a lime slurry solution, which is mixed in the product water. Lime is dearer than limestone, but the equipment can be arranged compactly, and hence the method allows a cost lower as a whole than in the case of dissolving limestone directl

It is regarded as particularly economical when carbon dioxide is obtainable through pretreatment. After mixing calcium bicarbonate, the product water will be controlled for pH with a solution of calcium hydroxide.

(iii) Dissolution of limestone in acid

The method comprises mixing a solution of calcium bicarbonate in the product water, and is prepared by crushing limestone and dissolving it in acid. The method is economical where small limestone can be obtained cheaply. Of course, the product water must be controlled for pH after mixing.

(iv) Filtration through a fixed bed of limestone

This is a method to increase hardness by allowing the product water to pass through a fixed bed of limestone. While it is difficult to obtain satisfactory hardness, the method is serviceable where iron must be eliminated. Iron settles on the surface of lime and hence must be washed in acid from time to time. Dolomite burnt lightly can be used instead of limestone.

5.9.4. Diffusion of Discharged Brine

(1) Discharged brine diffusion in outline

(a) Quality and quantity of discharged brine

In the case of desalination equipment under the evaporation process, fresh water is obtained by evaporation of the heated seawater, therefore concentrated seawater somewhat higher in temperature than raw seawater is produced according to the quantity of fresh water produced. To utilize the concentrated seawater for recovery of by-products, it is fed to the ensuing process with high salinity. However, it is mixed with seawater, having cooled down the heat rejection section or condenser for the desalination equipment normally and then returned to the sea again as discharged brine. Where a thermal power plant is in operation jointly, it is discharged together with the condenser cooling water having worked at the thermal power plant.

As mentioned, the discharged brine is somewhat higher in temperature than raw seawater and has it's salinity increased accordingly. And these conditions are governed by specifications and operating condition of the desalination plant and a thermal power plant operating jointly.

It is specified by the water temperature regulations in Japan that the drainage temperature rise must be kept less than 7°C above the surrounding seawater temperature. Therefore, temperature of the discharged brine is subject naturally to the said regulations. The situation is then such that the regulations are established according to circumstances in foreign countries.

An increase in salinity depends on the ratio of the quantity of produced fresh water to that of seawater taken in and the mixing percentage of the condenser cooling water at thermal power plants. In Japan the designed maximum working rate of the condenser cooling water is specified at 3.5 - 4 m^3/sec. per 100,000 kW in output for the thermal power plant and 6 - 6.5 m^3/sec. for the atomic power plant. These values are therefore good ones to work with where a power plant is to operate jointly. The following is to describe concentration of the discharged brine, increase of salinity and quantity thereof concretely:

Fig. 5.9-17 represents a heat balance of the 6,000 m^3/day multistage flash evaporating equipment set up at Benghazi City, Libya. As illustrated, brine of 9 - 10°C in temperature rise and 12.6% in salinity increase over the intake seawater will be discharged at 1,980 t/hr (0.55 m^3/sec.) from mixing the concentrated brine with the discharged seawater.

Fig. 5.9-18 shows a flow of intake seawater, produced fresh water and discharged brine of 100,000 m^3/day multistage flash evaporating equipment developed through the big project in Japan. According to the figure, temperature rise of the effluent to the discharge tank is 9.1°C, therefore to adjust the temperature rise to 7°C through bypass dilution, 4,770 t/hr (1.33 t/sec.) of the seawater will be necessary. The rate of brine discharge from the 100,000 m^3/day desalination equipment is 21,470 t/hr, the salinity increment is 6.36‰ of the intake seawater and the temperature rise is 7°C in consequence. Then, where a thermal or atomic power plant is to operate jointly with the equipment, condenser cooling water is added to be brine discharge mentioned before, thus diluting the increased salinity. Generally in comparison with the intake seawater, the discharged brine temperature rises 5 - 10°C and the salinity increases 2 - 10‰. And the rate of brine discharge from the desalination plant itself is several times as much as the produced fresh water However, the rate of condenser cooling water from the power plant is by far more.

(b) Necessity of predi ting diffusion of discharged brine

As referred to in the foregoing paragraph, the desalination equipment takes in a large quantity of seawater, produces fresh water partly from it and then returns the major part again to the sea as discharged brine high in temperature and salinity. The discharged brine which is released to the sea assimilates with the surrounding seawater to show a uniform diffusing pattern. Namely, momentum of the discharged brine is large in the neighborhood of the effluent outlet and then it diffuses as it is mixed in the surroundin seawater. Upon losing energy, the discharged brine is kept from friction with the ambient seawater and enters in the sea area governed by diffusion. Differences in temperature and salinity of the discharged brine are reduced by mixing and dilution with the surrounding seawater, moreover, heat exchange is conducted with the air at the surface of the sea It is a matter of importance for engineering consideration on the desalination equipment to investigate such diffusion processes of the discharged brine to predict the diffusion area under various conditions.

The first reason is that the discharged brine will have to be kept from recirculation and coming in again as intake seawater. For the intake equipment, quality seawater must be obtained constantly through the year, for which the discharged brine must not be mixed in the intake seawater.

The second reason is the influence that the brine exerts on the environment. Because of impact that the discharge of a volume of the effluent into a limitted sea area might have on organisms living therein, it is necessary to grasp the diffusing conditions of the discharged brine, and particularly distributions of temperature and salinity first of all. Generally a rise of the salinity may lower the concentration of dissolved oxygen but is inclined to activate metabolism and intensify osmotic pressure working on the membrane. However, cause and effect of the natural world are correlated complicatedly, and the influence of temperature and salinity is not necessarily so simple. It is then dangerous to apply basic experimental results straight to the natural environment.

Meanwhile, most of the thermal power plants in Japan are located in the coastal regions, and hence data are comparatively ready on influence by temperature rise, i.e. problem of warm effluent. For example, the influence for warm effluent which was exerted on marine organisms by fisheries experiment stations was examined by Tsuruga Power Station of Japan Atomic Power Co. and Mihama Power Station of Kansai Electric Power Co. located in Tsuruga Penineula for 4 to 5 years after the operation from preliminary investigations made in 1964. The results obtained there are as follows:

1 There is no significant difference observed in composition and quantity of plankton in the neighborhood of the discharge outlet of the warm effluent at 20 m^3/sec. or so.

2 The biological phase of seaweeds in a zone near the discharge outlet where the water temperature rose around 4°C was effected, but recovered rapidly at temperatures lower than that.

3 An influence was not specifically observed on sticking organisms and fish.

(c) Development of diffusion predicting method

In taking the warm effluent working as an environmental pollutant through heat, the water temperature rise has been controlled below 7°C in Japan. A research on diffusion of the warm effluent has also been made actively, and a preliminary investigation is usually conducted through technolo for predicting diffusion of the warm discharge before constructing power plants.

The warm effluent being seawater with only a temperature rise, behavior of the warm discharge can be predicted by taking up heat diffusion only. However, in the case of discharged brine, there is involved a change in salinity in addition, therefore diffusion of both heat and salinity must be taken into consideration. A development of prediction for diffusion of the discharged brine was executed as a link of technology assessment under "Seawater Desalination and By-Product Utilization" of the big project by the Minist of International Trade and Industry, and present research wa made in the form of expanding the conventional predicting method for diffusion of the warm effluent. Details will be given in the ensuing paragraphs and its outline comes is as follows:

To grasp the actual behavior of discharged brine first of all, the effluent from the test module at Oita Station was measured for the diffusing condition. Then it was ensured that an influence could be retained in a limited area for ambience according to the submerged outfall system.

In parallel with the above measurement, a hydraulic test was carried out to clarify basic characteristics of the discharged brine under various conditions, and further the same condition as the diffusion of the effluent at Oita was reproduced in a model.

Next, a numerical model was developed through analyzing the hydraulic test result, by which a simulation analysis of diffusion in the three-dimensional sea area was carried out. And it was ensured that the measured value for the hot effluent and the solution by numerical model coincided with each other.

As the result of having applied the diffusion prediction thus developed to 100,000 m^3/day desalination equipment, the submerged outfall system provided to be a superior method to diffuse the discharged brine.

(2) Measurement of discharged brine diffusion

To grasp actual behavior where the discharged brine is released in the sea, measurement was carried out for diffusing conditions of the discharged brine from the Test Model at Oita. The intake/ discharge facilities arranged at the Oita Test Module were located in Otozu Anchorage.

The raw seawater is taken in by a natural intake system from a depth (-12 m) 50 m off the coast line where river water and city sewage are less influential. Then, the used discharged brine and cooling seawater are mixed in the discharge tank and released into the sea by way of jet diffusing equipment arranged on the sea bottom (-9 m) 50 m off shore for ready mixing with the ambient seawater. The intake discharge equipment is as shown in Fig. 5.9-19.

For measurement of the discharged brine from the Oita Test Module, an investigation was carried out five times, all told, during the period 1975 to 1976, as shown in Table 5.9-6, including the first beckground investigation. The measurement was performed under the situation that the distribution of salinity and the surrounding seawater temperature changed for each investigation, the following information was obtained for influence of the discharged brine through general considerations:

1) Characteristics of surrounding sea

Otozu Anchorage, where the discharged brine was released, is situated at south of Beppu Gulf and has an area 600 m wide and 2,000 m long with its mouth surrounded by breakwaters on the east and west. Otozu River and Hara River, in the inner part, flow into the anchorage. Otozu Anchorage, surrounded by breakwaters and kept from mixing with the open seawater, is influenced by an inflow of river water and has a light density of seawater on the surface layer as compared with the middle and lower layers.

2) Diffusing condition by multi-port nozzles (See Fig. 5.9-20)

The discharged brine is mixed actively in the neighborhood of the release port and is diluted through mixing with the ambien water slowly, flows upwards and extends horizontally under the surface which is influenced greatly by river water with light density. A temperature rise above 1°C effects an area 8×10^3 max.

3) Diffusing condition by vertical upward nozzles

The discharged brine coming through the undersea nozzle goes up vertically to lose its upward flow rate at the surface of the sea, then gets extends horizontally but soon settles down (2 - 5 m radial on the surface) to the middle layer with less influence by the river water, and then spreads horizontally. A temperature rise above 1°C effects an area about 6.5×10^3 m^2 max.

4) Diffusing condition by horizontal nozzles

The discharged brine goes towards the open sea from undersea and as it mixes with the ambient seawater, flows slowly upwards and then spreads horizontally in the middle bed zone which is less influenced by river water. A temperature rise above 1°C effects an area about 450 m^2 max.

As mentioned above, the diffusion of the discharged brine from the Oita Test Module varied according to the ambient sea conditions and the intake/discharge conditions. However, a temperature rise above 1°C was influential at worst giving an area 8 x 10^3 m^2 max. In case, meanwhile, the same quantity of effluent is discharged from the surface, the area is estimated at about 1.5 x 10^5 m^2 (predicted according to Nitta's formula), and thus it was ensured that the influence on ambience could be restricted within a fairly limited area by discharging with a submerged outfall system.

(3) Diffusion predicting method

(a) Plume

As one of the measures to reduce the area of influence by discharged brine from desalination plants, there is a submerged outfall system available to make positive mixing with the surrounding seawater to dilution possible. In order to understand the diffusion process of effluent through a submerged outfall, the mixing process for various discharge conditions under various sea conditions must be cleared up.

A rapid flow discharge into fluid is generally called a jet or plume. A flow without buoyant action at the jet is called a jet. A plume has both buoyancy and initial momentum and is otherwise called a gravitational jet. The flow pattern of the discharged brine in this paragraph is classified within the category of a gravitational jet to discharge a higher density of fluid as compared with the ambient seawater in the jet condition.

The diffusion of a gravitational jet varies according to the direction in which momentum acts (discharge direction, the direction in which buoyancy acts (positive or negative density difference between discharged water and ambient), the flow in the surrounding fluid and the density stratification. As regards the direction in which the warm effluent jet is discharged, a method is conceivable whereby it is discharged at certain angles against the vertical direction normally for improving dilution efficiency. However, movement vertically upwards and horizontally, which come to both extremes, can be pointed out as typical discharge directions.

In the usual theoretical development of a plume in a uniform fluid, the concept of entrainment coefficient has been introduced, and a method has been employed whereby the velocity of flow on the plume axis with the distance from jet source, concentration (or density) and the width of plume are obtained on the basis of asimilarity hypothesis. Since the boundary between the surface and the bottom cannot be considered in this theory, the analytic solution is applicable only to a process in which the plume reaches the water surface or bottom.

In predicting the diffusion of discharged brine, it is conceivable that the configuration around the discharge outlet and the influence of the surface layer may play an important part in the diffusion, and hence three-dimensional analysis will be necessary.

(b) Fundamental equations on discharged brine

The three-dimensional numerical model consists of a kinematic equation in hydrodynamics, continuous equation and mass and heat conservation equation as given below.

The fundamental equations can be expressed as follows:

1) Kinematic equation:

$$\frac{\partial U_i}{\partial t} + \frac{\partial}{\partial x_j}(U_i U_j) = -\frac{\partial \phi}{\partial x_i} - g\frac{\Delta\rho}{\rho_0}\delta_{3i} + A_h \nabla^2 U_i + A_z \frac{\partial^2 U_i}{\partial x_3^2} \qquad (1)$$

2) Mass conservation equation:

$$\frac{\partial}{\partial x_i} U_i = 0 \qquad (2)$$

where U_i: in ith velocity componet, xi: ith direction component, t: time, ϕ: ratio of pressure to density ($=P/\rho$), g: gravity acceleration, δ_{3i}: 1 for $i=3$, 0 for $i=1,2$, A_h,A_z: lateral eddy viscosity and vertical eddy viscosity respectively, ρ_0: reference density of discharged water, $\Delta\rho$: density difference between ambient seawater and plume.

For calculation of the discharged brine, an equation of state for seawater, $\rho = \rho\ (S, T)$, is used where S is salinity and T water temperature. In this case an effect of buoyancy can be simulated in combination of the mass conservation equation regarding heat and salinity with the kinematic equation of fluid.

3) Density conservation equation:

$$\frac{\partial \rho}{\partial t} + U_i\frac{\partial \rho}{\partial x_j} = \frac{\partial}{\partial x_j}\left(\frac{K_j}{\rho}\frac{\partial \rho}{\partial x_j}\right) \qquad (3)$$

where K_j: eddy thermal diffusivity.

For density calculation expressed in the above fundamental equations, the following Knudsen's formula can be used. According to Knudsen, the following relation holds between chlorinity Cℓ(‰) and density σ_0 at 0°C:

$$\sigma_0 = -0.069 + 1.4708C\ell - 0.001570C\ell^2 + 0.0000398C\ell^3 \qquad (4)$$

Then, salinity S is correlated with chlorinity Cℓ as follows:

$$S = 1.80655 \times C\ell \qquad (5)$$

Density σ_t at arbitrary temperature t is

$$\sigma_t = \Sigma_t + (\sigma_0 + 0.1324)[1 - A_t + B_t(\sigma_0 - 0.1324)] \qquad (6)$$

where Σ_t, A_t and B_t are functions of temperature only.

Since the turbulent transport coefficients for momentum, Ah and A_z are functions of the flow field, these quantities mus be related to appropriate mean flow properties. Here Prandl hypothesis is applied to plume. The component of the coefficient for eddy viscosity in the direction rectangular to the axis is taken as follows:

$$A_r = C\ell_r W_{max} \qquad (7)$$

where ℓ_r is a mixing length, which is equal to the plume hal radius. W_{max} is the plume centerline velocity and C a const equal to 0.0256.

As boundary conditions on flow and density, velocity gradien in a direction rectangular to the surface and density gradie are zero, flow rate at the solid wall is zero, density flux a direction rectangular to the wall is zero on the seawater surface, and normal velocity and released water density are given at the discharge port. Then, it is conceivable that $\partial^2\rho/\partial_n 2 = 0$ (n being a component in a direction rectangular to the boundary) as density conditions at the boundary when the flow comes out of the calculation area, and water as natural surrounding water is incoming when the flow comes in the system.

(c) Estimating diffusion of discharged brine

Discharged from a 100,000 m^3/day desalination plant is taken brine as an example.

Flow rate, temperature and salinity of the discharged brine in this case are as follows as described in 5.9.4-(1) when the natural seawater temperature and its salinity are 25°C and 32.52% ($C\ell = 18^0/_{00}$) respectively:

Through bypass dilution conducted for keeping the temperature difference ΔT_o between intake seawater and discharge at 7°C,

Discharge flow rate $Q_o = 5.96\ m^3/s$ (= 21,470 m^3/hr)

salinity concentration: $38.88^0/_{00}$

The salinity difference ΔS_c between intake and discharge is $38.88 - 32.52 = 6.36^0/_{00}$.

Using the a mono-vertical discharge pipe system, conditions for numerical simulation are set up as given in Table 5.9-7 on the basis of design values for discharged brine at a 100,000 m^3/day desalination plant. The discharged outlet is located at a position of Z = 2.7 m on the sea bottom. Fig. 5.9-22 show the distribution of flow velocities across the section of the central axis of the pump and the vertical distribution of the density respectively, which are obtained through simulation analysis on the basis of equations (1) to (3). In this case, there is a negative density difference of $\Delta\rho = 2.7 \times 10^{-3}$ between discharged brine and surroundings seawater. This is attributed to the high salinity of the discharged brine despite releasing at a high temperature difference of 7°C. Therefore, the vertical upward flow of the discharged brine is prevented from rising up to the sea surface due to the influence of negative buoyancy and the velocity of vertical upward flow is lost below the surface. Then, along with the surrounding water entraining from the upper layer, the discharged brine goes down along the outer edge of the plume and shows a tendency of dispersing along the bottom layer from the neighborhood of the horizontal distance, r/D = 10. Due to the addvective effect of discharged brine, the distributions of water temperature and salinity are almost similar to the density distribution given in Fig. 5.9-22.

As shown in Fig. 5.9-22, it is considered that the discharged brine is diluted nearly 20 - 25 times where $r/D = 10$, and hence it can be judged that the area influenced by the effluent will be restricted locally.

It can be thus recogniged that the discharged brine diffuses along the sea bottom, because the density difference between discharged brine and surrounding seawater is a negative value of 2.7×10^{-3}.

Now, a conceivable influence that the discharged water will exert on oceanophysical environment when carrying out a seawater desalination project by joint operation with a power plant is taken up for examination. When considering a power station of 500,000 - 600,000 kW capacity, it is estimated that the rate of cooling water for condensers of the power plant will be 30 m^3/s and the temperature difference between discharged water and natural environmental water will be $\Delta T = 7°C$. Assuming the natural environmental water temperature be 25°C, the density of warm effluent only water dnesity (ρ_o) and environmental water density (ρ_s) are $\rho_o = 1.019,196$ and $\rho_s = 1.021,495$ respectively, and the density difference is 2.299×10^{-3}. It is supposed in this case that the discharged water will spread on the surface with positive buoyancy.

Meanwhile, where the discharged brine is mixing with warm effluent from the power plant, the flow rate of discharged water, the temperature rise and the salinity are as follows:

Discharged flow rate = $30.0 + 5.96 = 35.96$ m^3/s
Temperature rise = 7.0°C
Increment of salinity = $\frac{30 \times 0 + 5.96 \times 6.36}{35.96} = 1.054‰$

Assuming that salinity of the environmental water is $32.52^0/_{00}$, density of the discharged water ρ_0 = 1.019,982 to bring about a density difference of $\Delta\rho = 1.513 \times 10^{-3}$ from the environmental water. Although the value is small in comparison with that of warm effluent only, there is positive buoyancy.

As the result of having predicted the diffusion by means of simulation analysis according to the conventional numerical model on warm effluent discharged on the surface layer, it is conceivable that, in joint operation with a power plant, the thickness of the layer of discharge increases by the warm effluent only as the density of salinity increases, but the diffusion of discharged brine will show similar behavior to that of warm effluent.

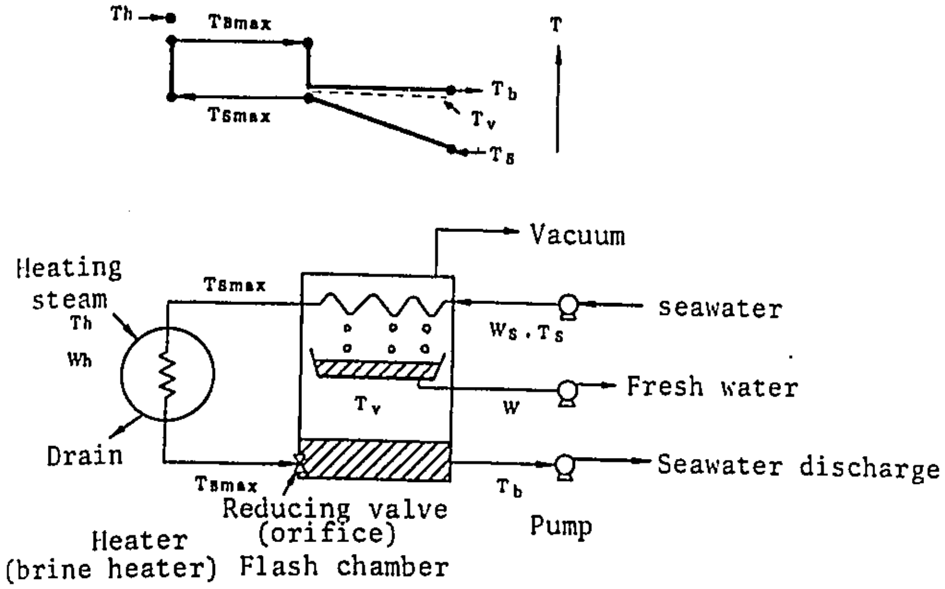


Fig. 5.1-1 1-stage flash evaporation plant (T;temperature, W;flow rate) and temperature distribution

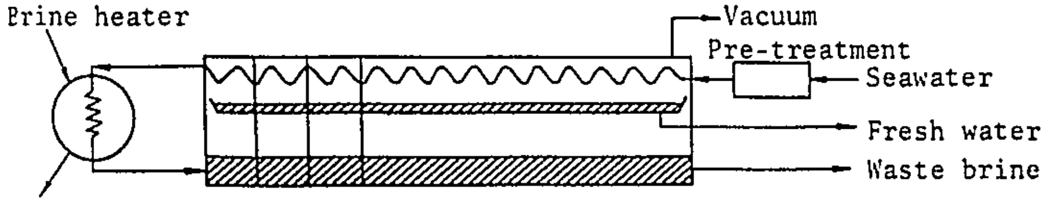


Fig. 5.1-2 Multi-stage flash evaporation plant(Reflux type)

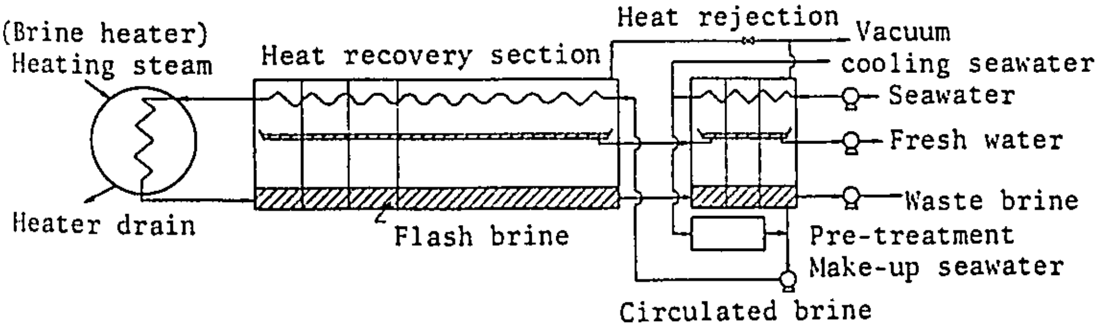


Fig. 5.1-3 Multiple-stage flash evaporation plant (Recirculation type)

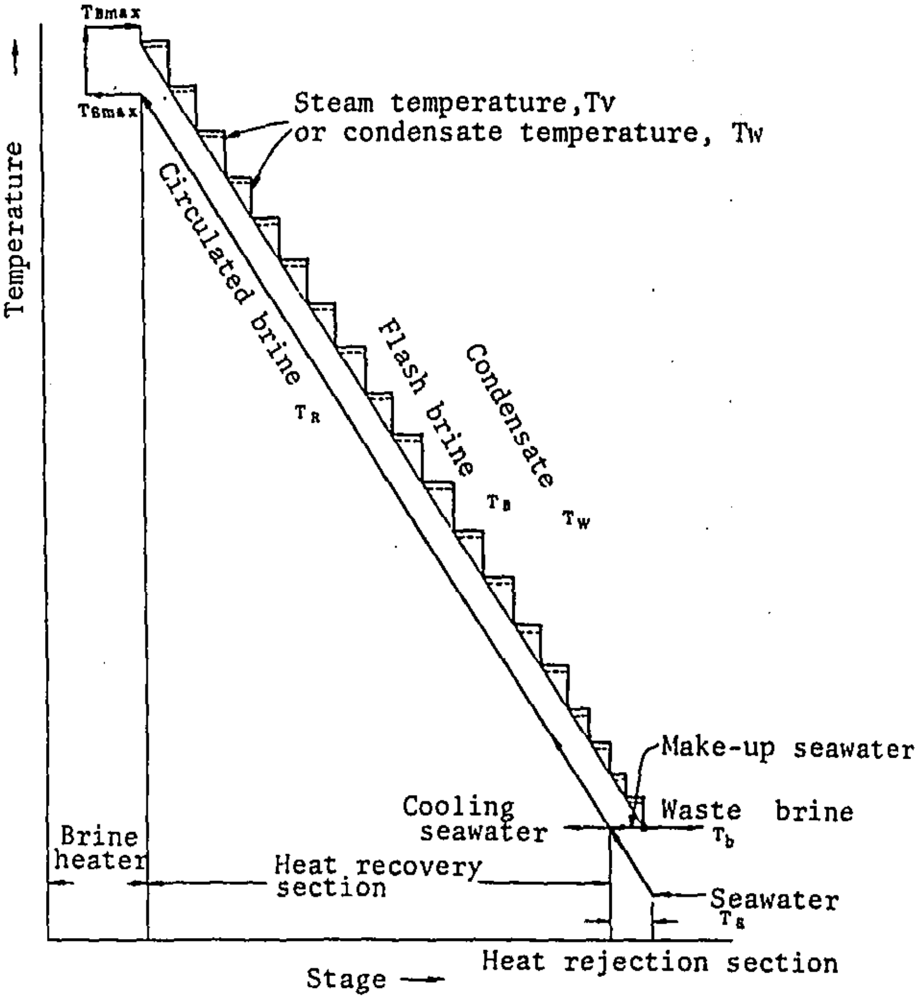


Fig. 5.1-4 Temperature distribution of multiple-stage flash evaporation plant

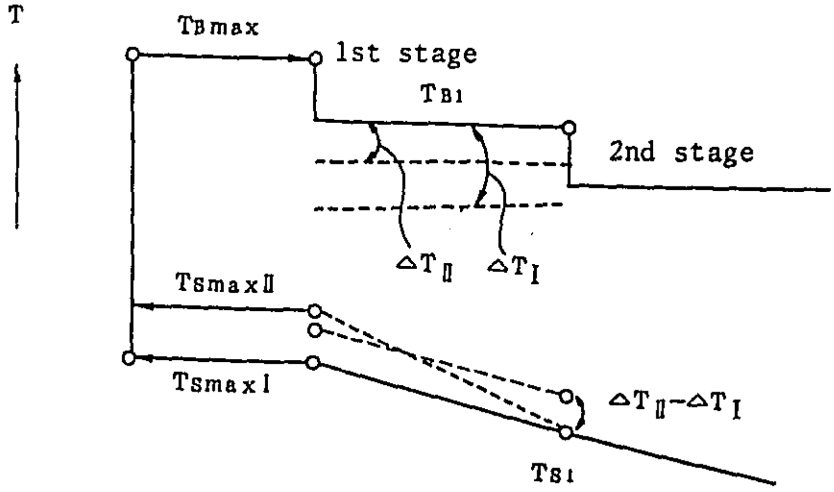


Fig. 5.1-5 Temperature distribution of 1st stage and adjoining part

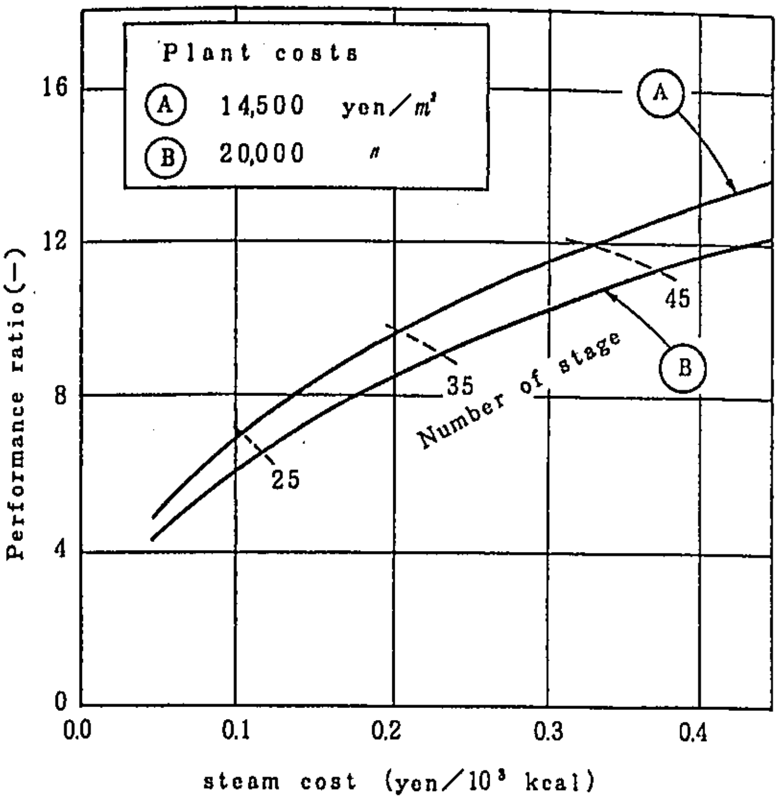


Fig. 5.1-6 An example of optimization by simplified model

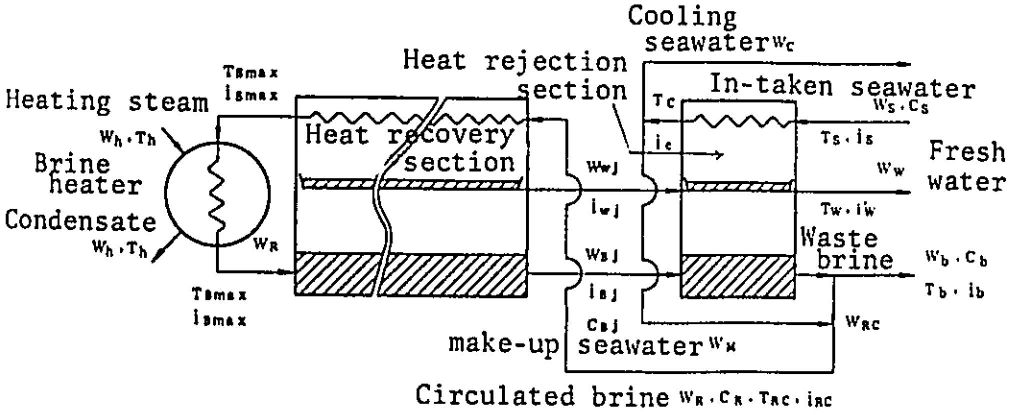


Fig. 5.1-7 Flow diagram of material balance

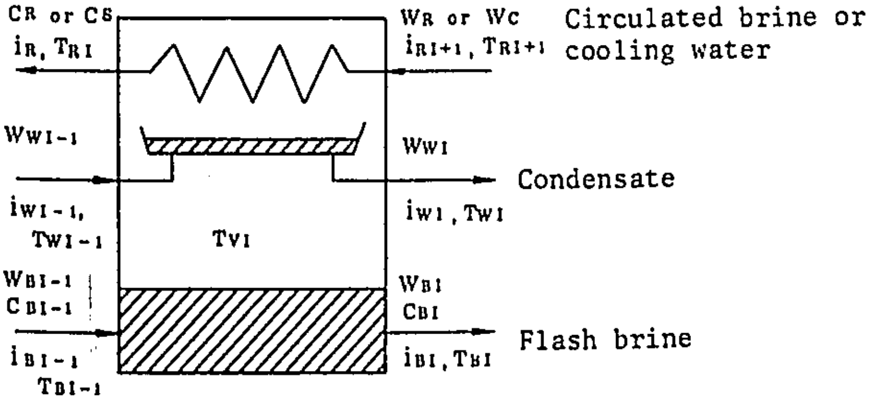


Fig. 5.1-8 Flow diagram of material balance at 1st stage

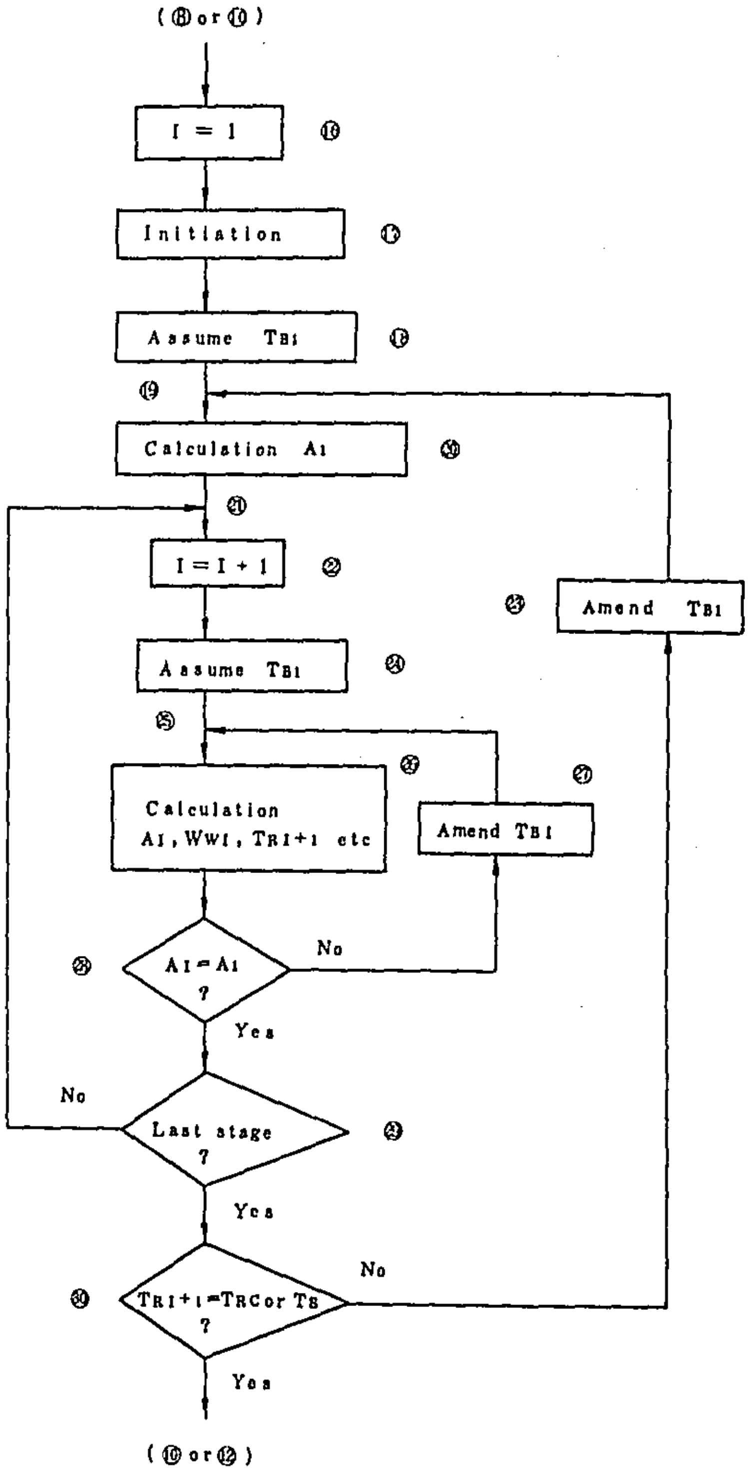


Fig. 5.1-9 Flow chart of calculating program for detailed model of MSF (1st stage)

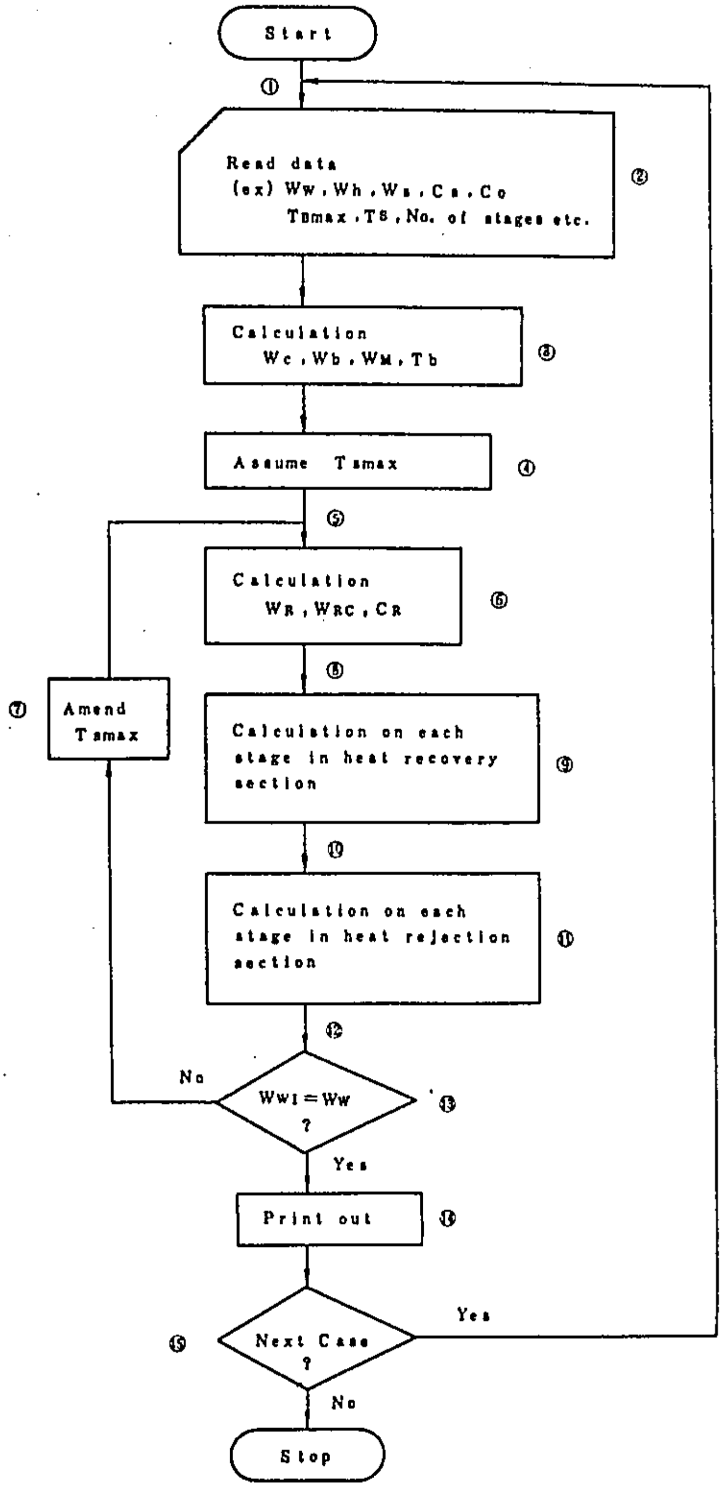


Fig. 5.1-9 Flow chart of calculating program for detailed model of MSF

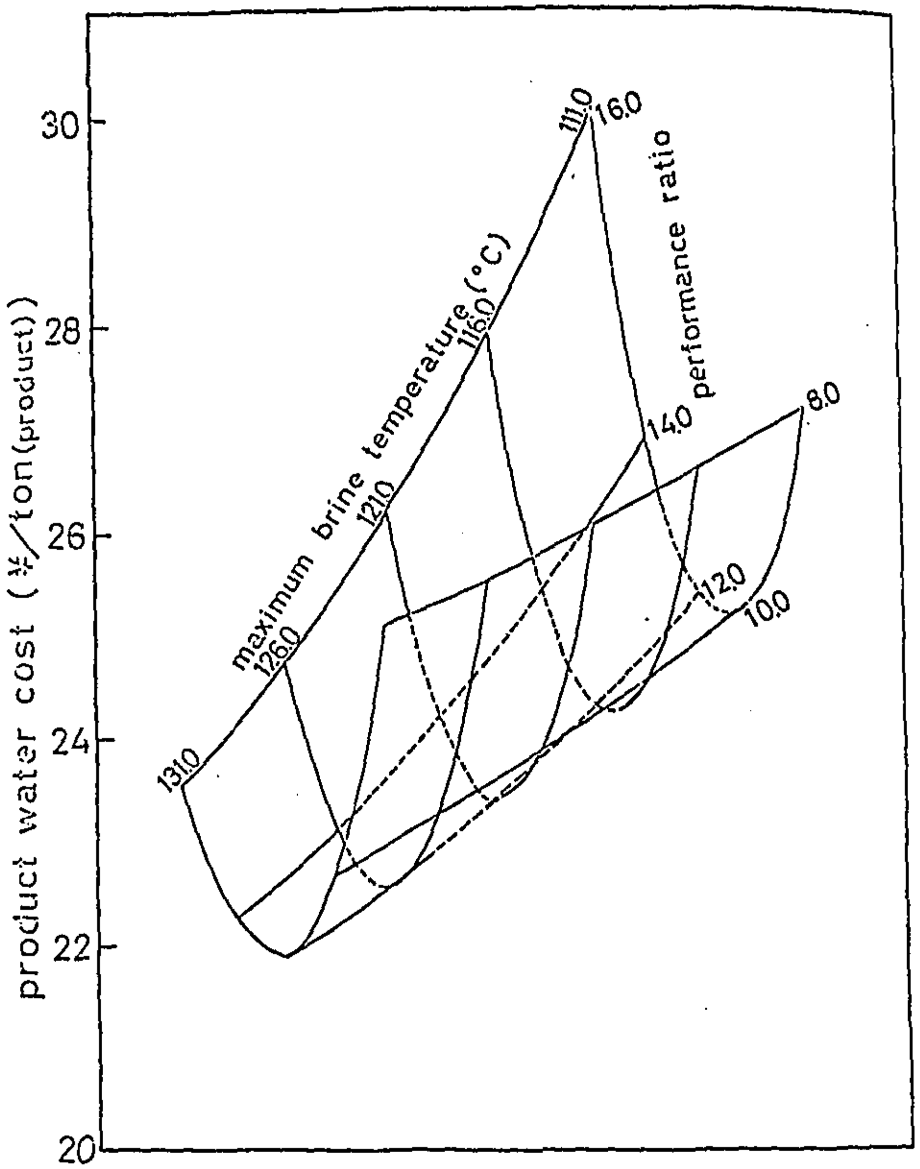


Fig. 5.1-10 Examples of optimization for detailed model

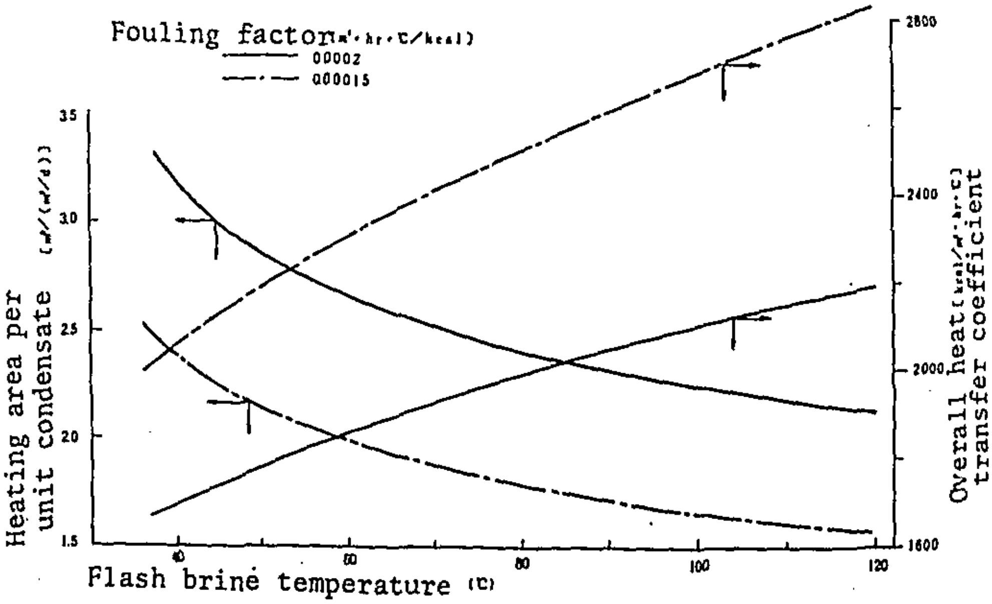


Fig. 5.1-11 Over-all heat transfer coefficient vs. heating area distribution per unit condensate

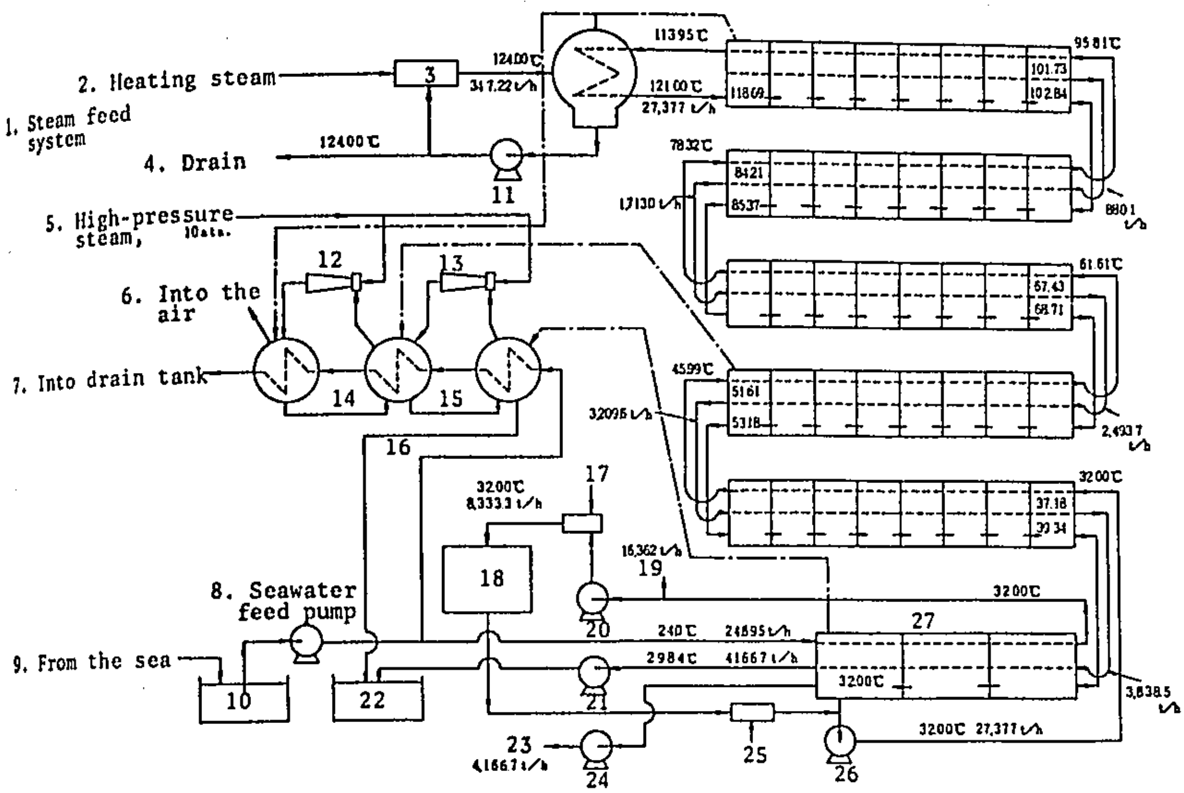


Fig. 5.1-12 Flow diagram of 100,000 t/d desalination plant (performance ratio;12)
Note: Figures shown inside the flash chambers refer to temperature °C

Notes:

1. Steam feed system
2. Heating steam
3. Superheated steam saturator
4. Drain
5. High-pressure steam, 10 ata.
6. Into the air
7. Into drain tank
8. Seawater feed pump
9. From the sea
10. Seawater pit
11. Drain pump
12. Ejector
13. Ejector
14. Drain
15. Drain
16. Drain
17. Sulphuric acid
18. Pre-treatment
19. Discharge
20. Seawater make-up pump
21. Fresh water pump
22. Fresh water pit
23. Discharge
24. Waste brine pump
25. De-bubble chemicals, Caustic soda
26. Brine circulation pump
27. Heat emission system

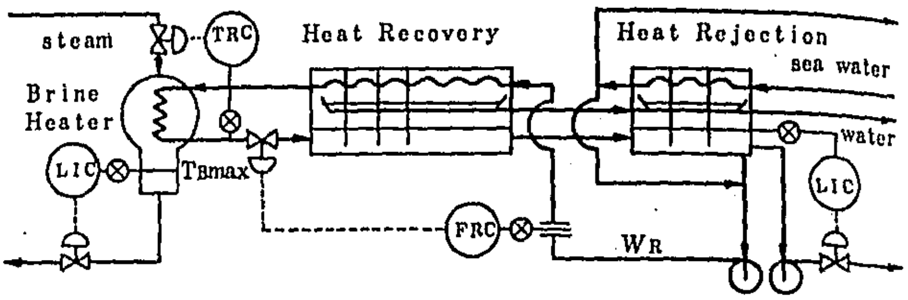


Fig. 5.1-13 General main controllers of a MSF

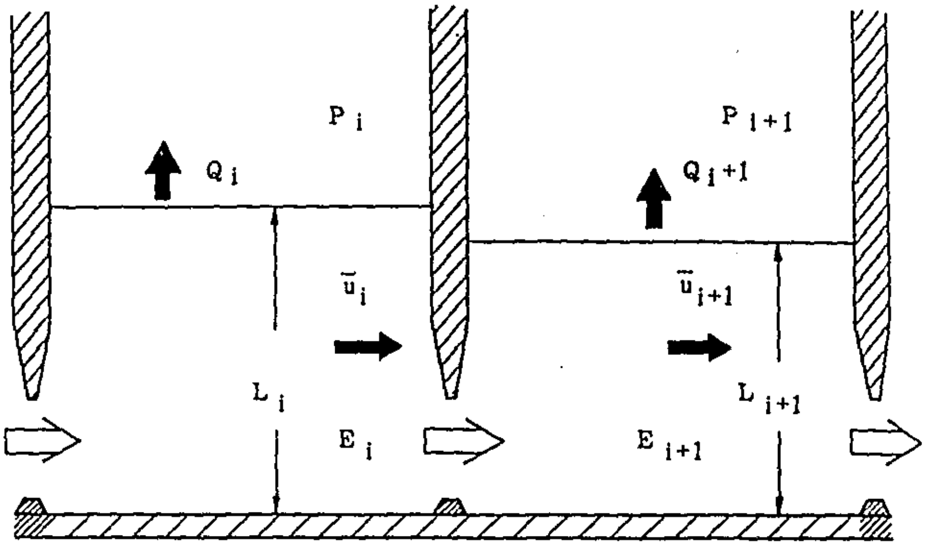


Fig. 5.1-14 Schematic diagram for interstage orifice

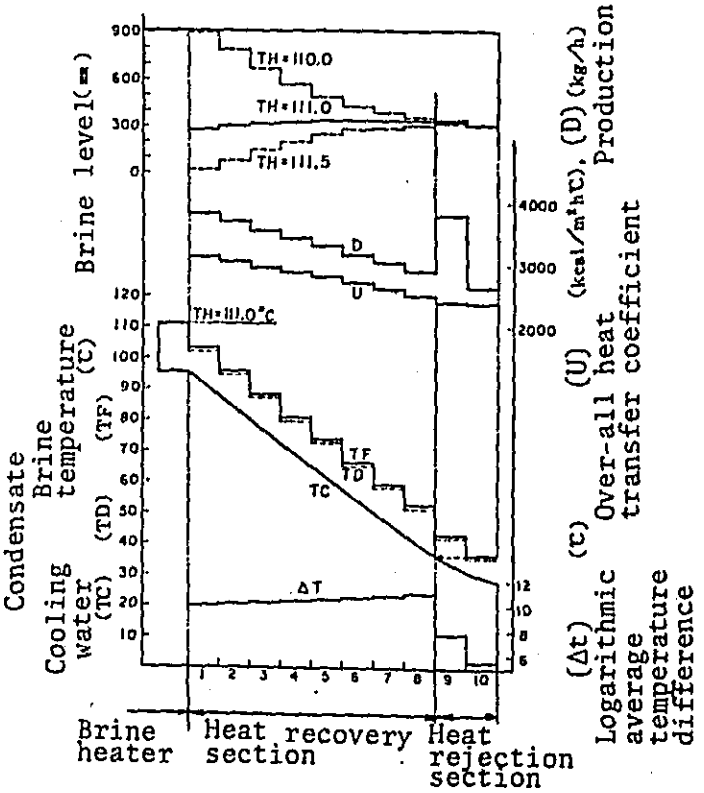


Fig. 5.1-15 State of individual stage under partial load (80%)

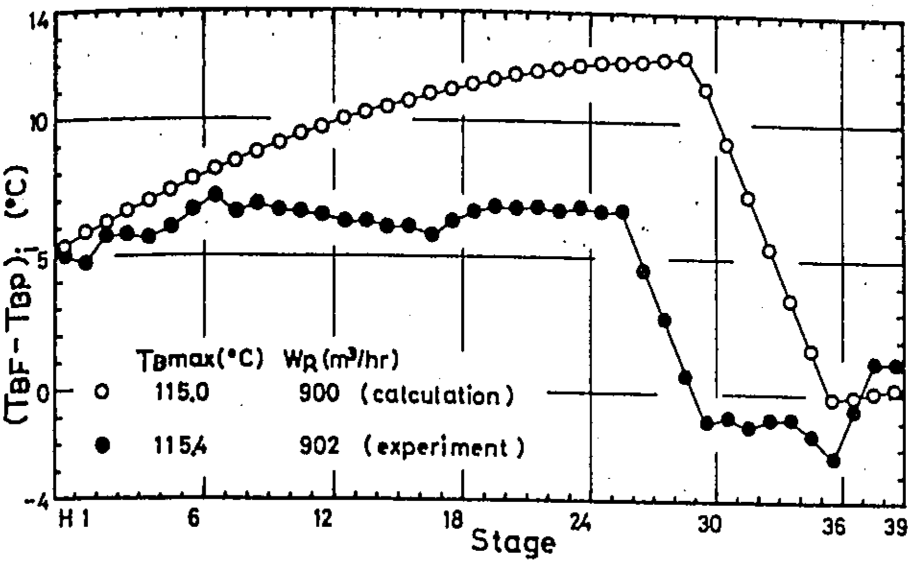


Fig. 5.1-16 Composition between simulation and experimental results for temperature distribution of flash brine under partial load

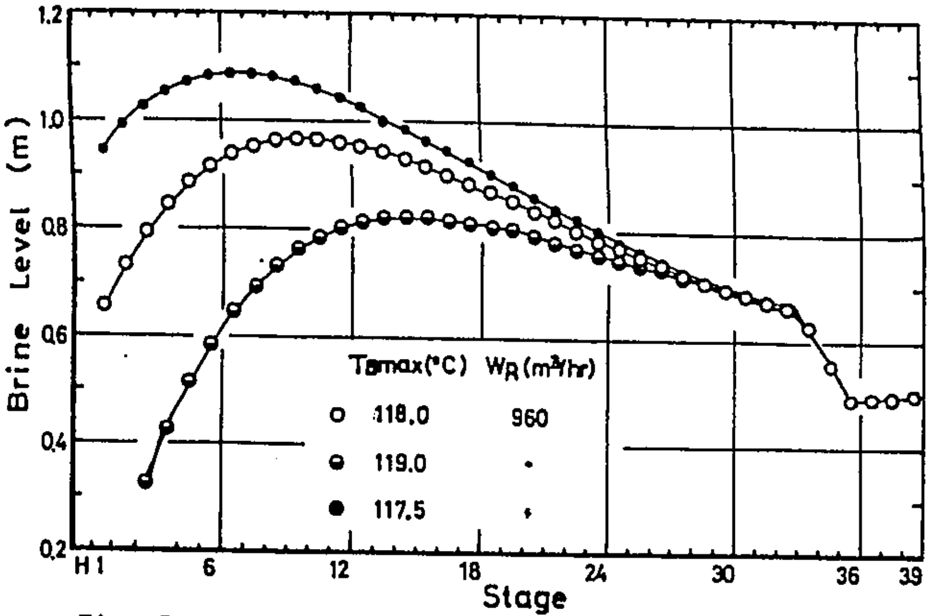


Fig. 5.1-17 Effect of maximum brine temperature for distributions of brine level

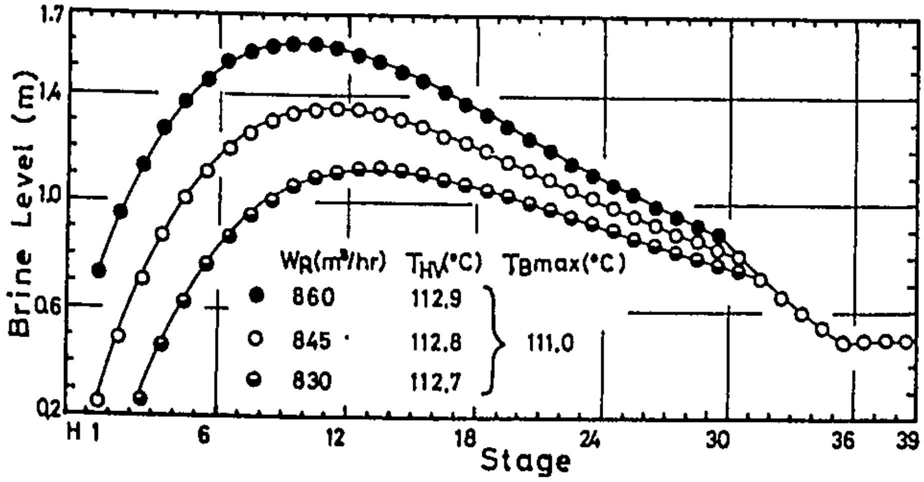


Fig. 5.1-18 Effect of recycle brine flow rate for distributions of brine levels

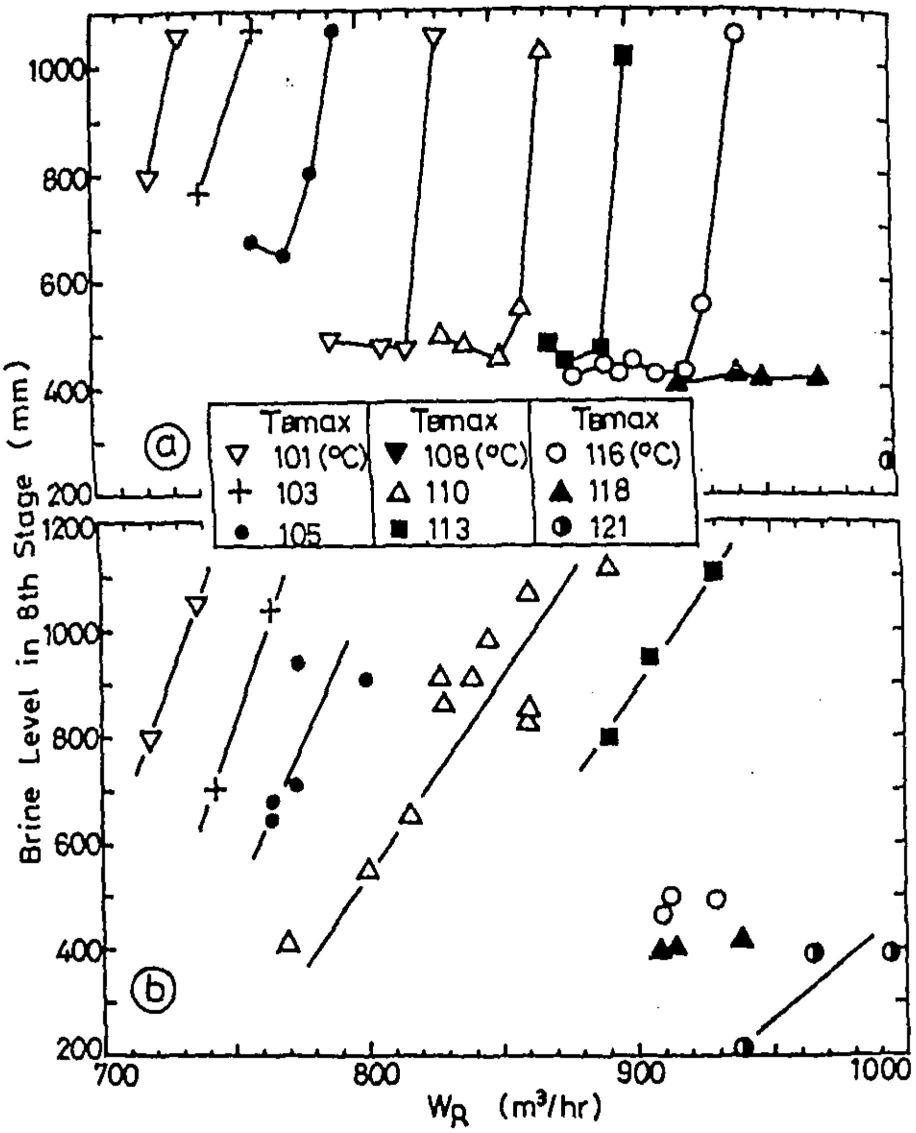


Fig. 5.1-19 Variation of brine levels in 8th stage

a ; Case of increase by step of T_{Bmax} or W_R
b ; Case of decrease by step of T_{Bmax} or W_R

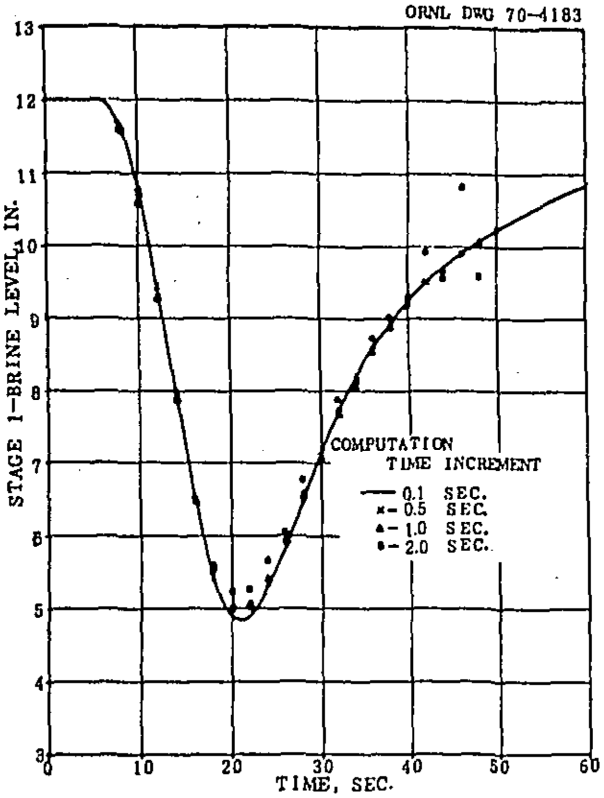


Fig. 5.1-20 Effects of Computation Time Increment on Accuracy: Stage 1 Brine Level Versus Time Following a 10 sec, 10°F Pulse Increase in Brine Heater Steam Temperature.21)

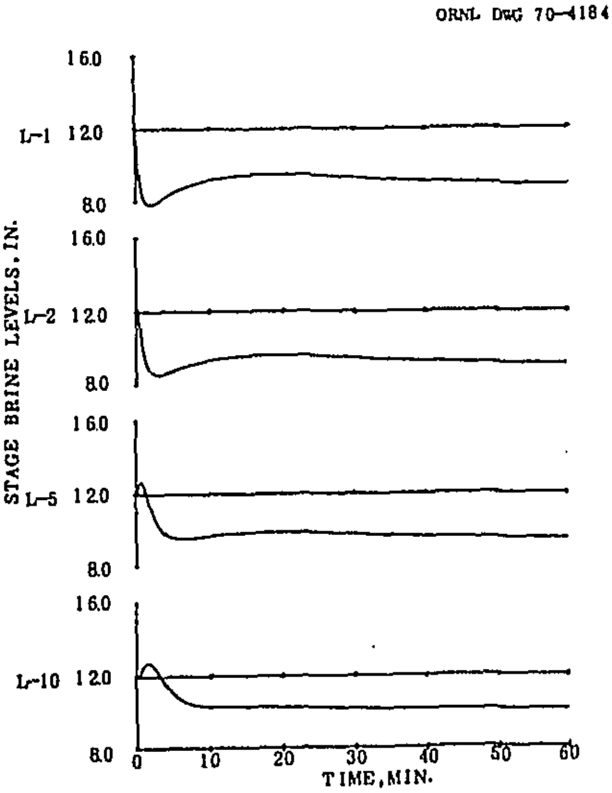


Fig. 5.1-21 Response of Tray Brine Levels to a 2°F Step Increase in Brine Heater Steam Temperature.

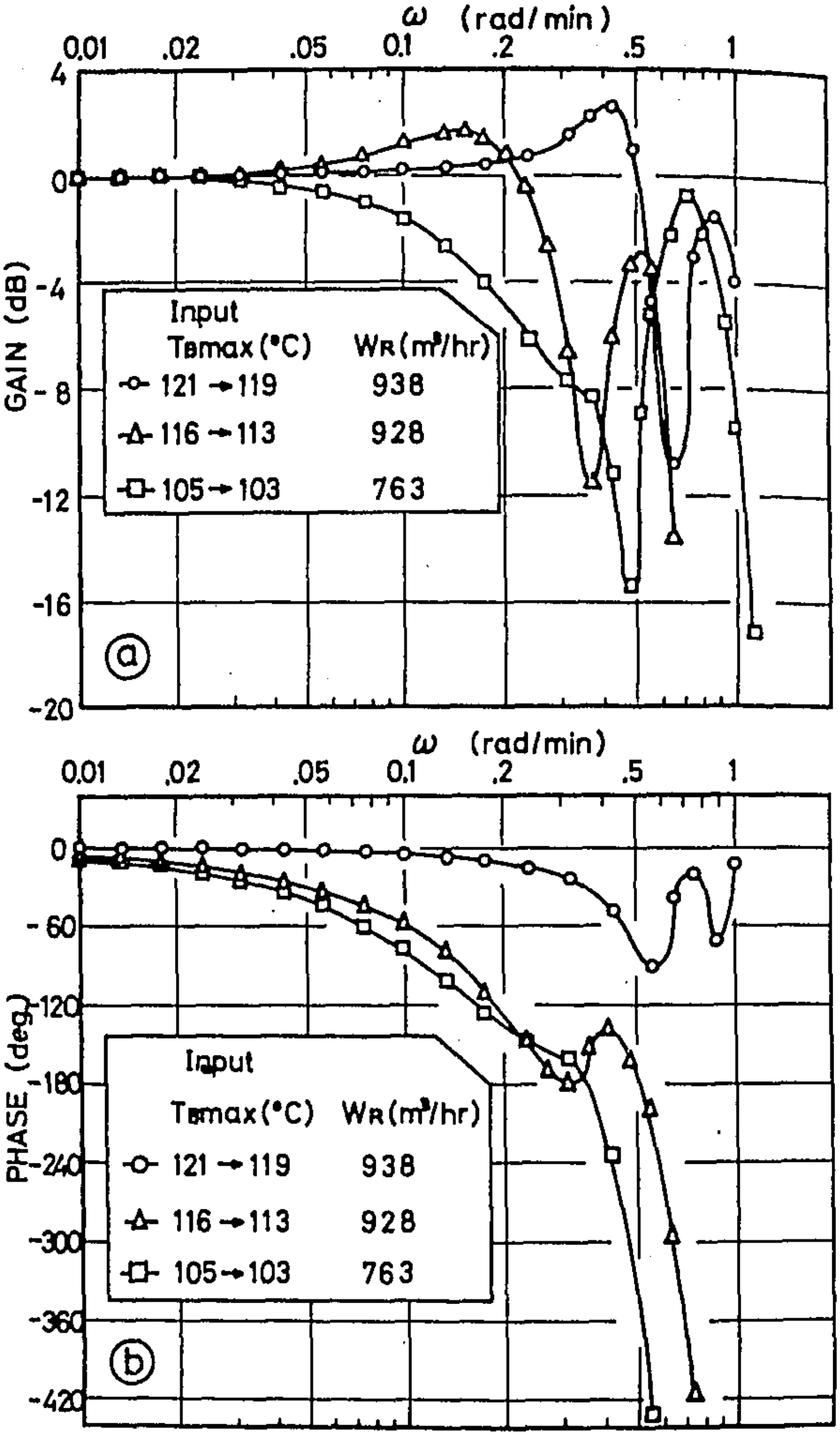


Fig. 5.1-23 Transfer-function (a ; gain, b ; phase) of brine level in 8th stage obtained for step down T_{Bmax} change

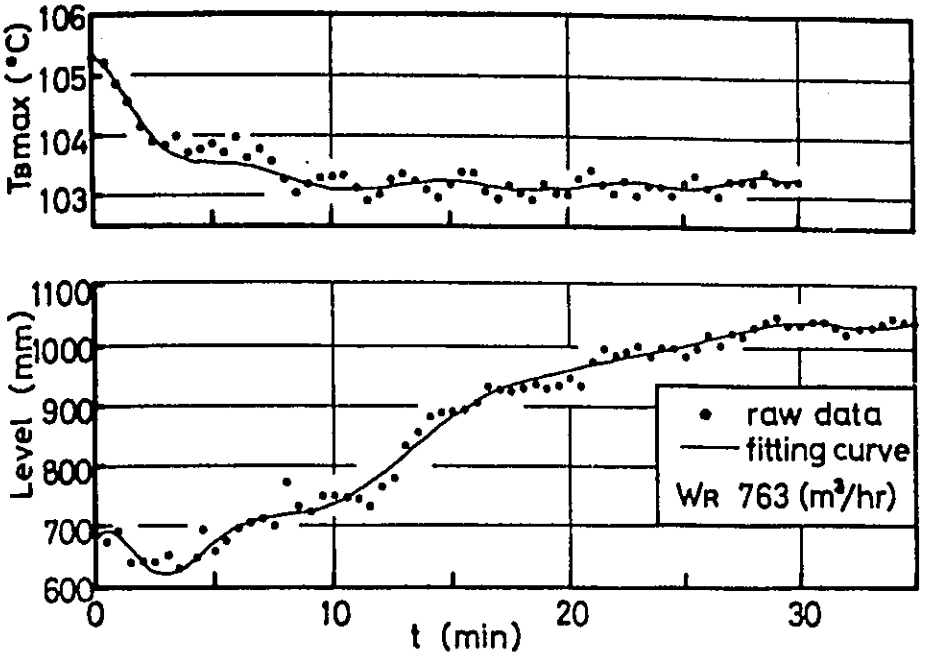


Fig. 5.1-22 Experimental response example of the level of 8th stage

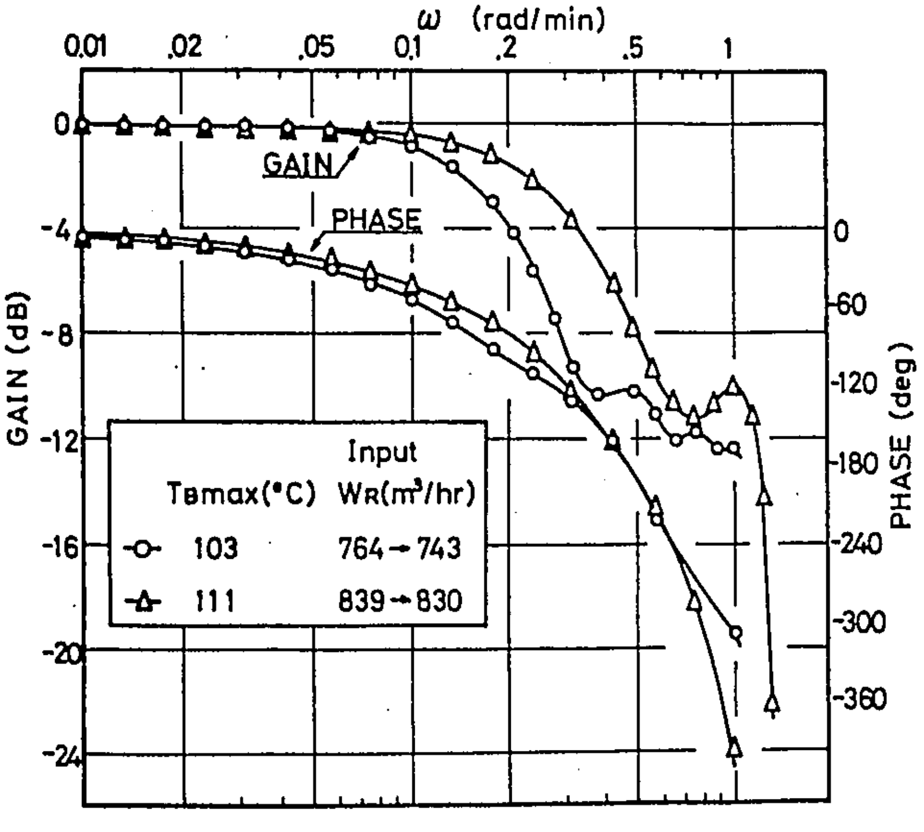


Fig. 5.1-24 Transfer-function of brine level in 8th stage obtained for step down W_R change

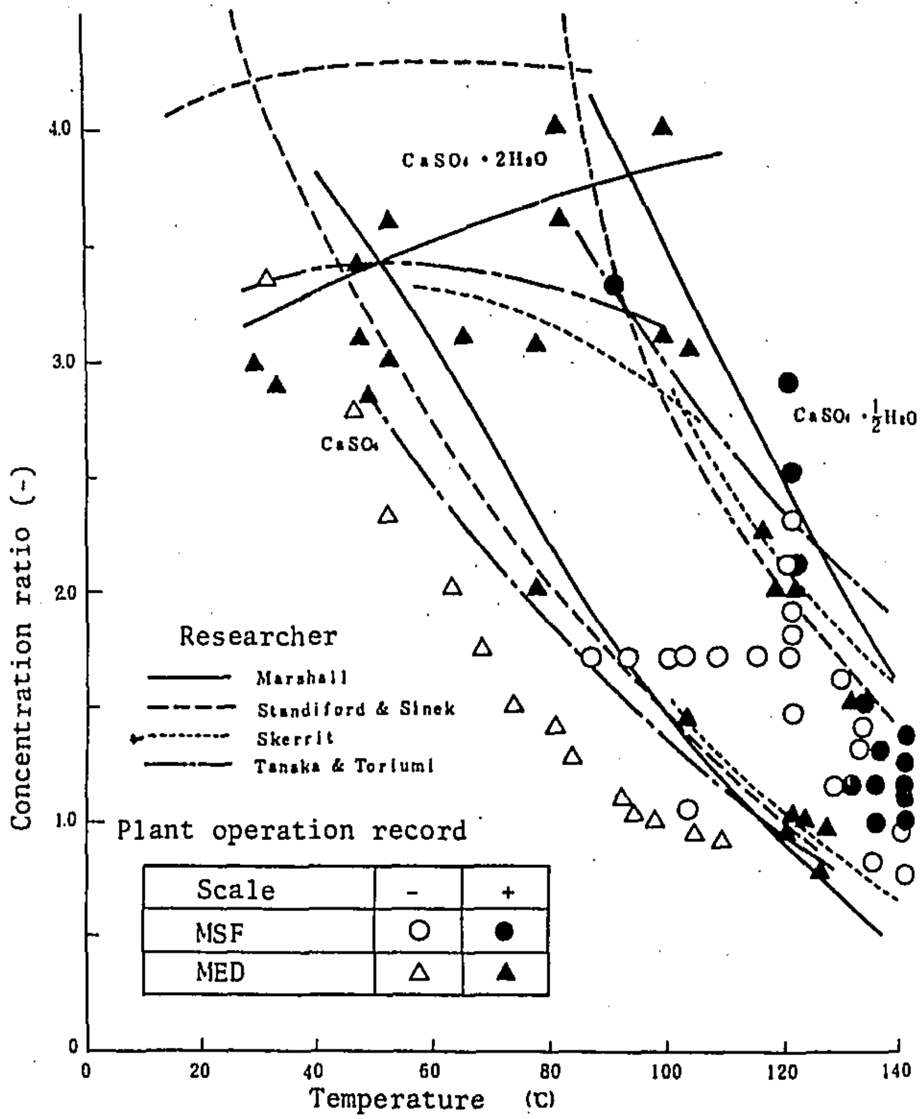


Fig. 5.2-1 Limit of separation for sulfuric calcium in seawater

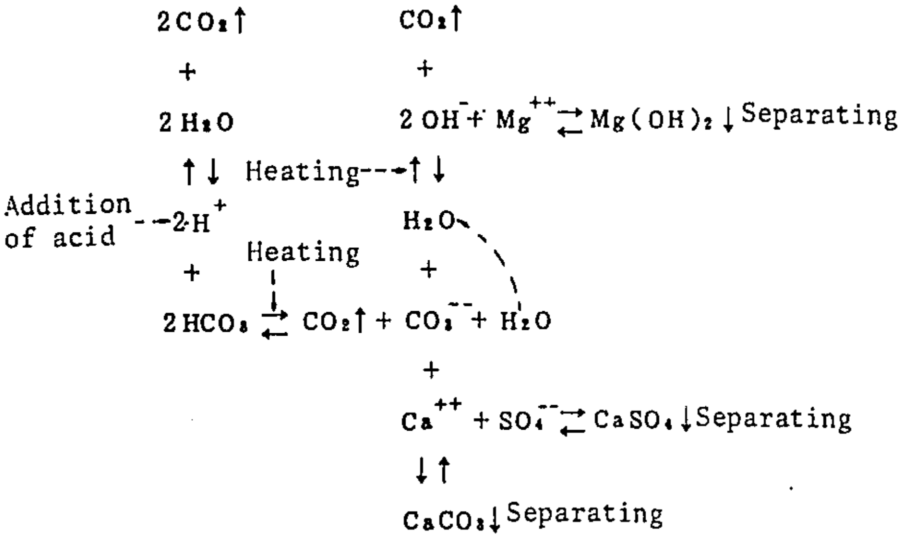


Fig. 5.2-2 Chemical equilibrium of scale component

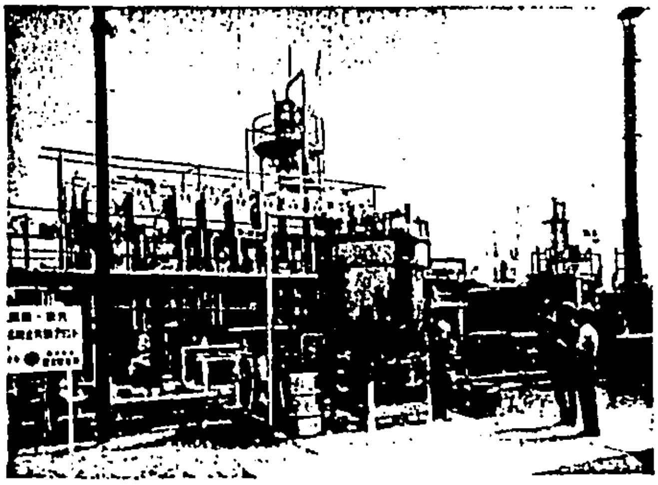


Fig. 5.2-3 Profile of scale preventing test device (100 m^3/day - 10 stage evaporation)

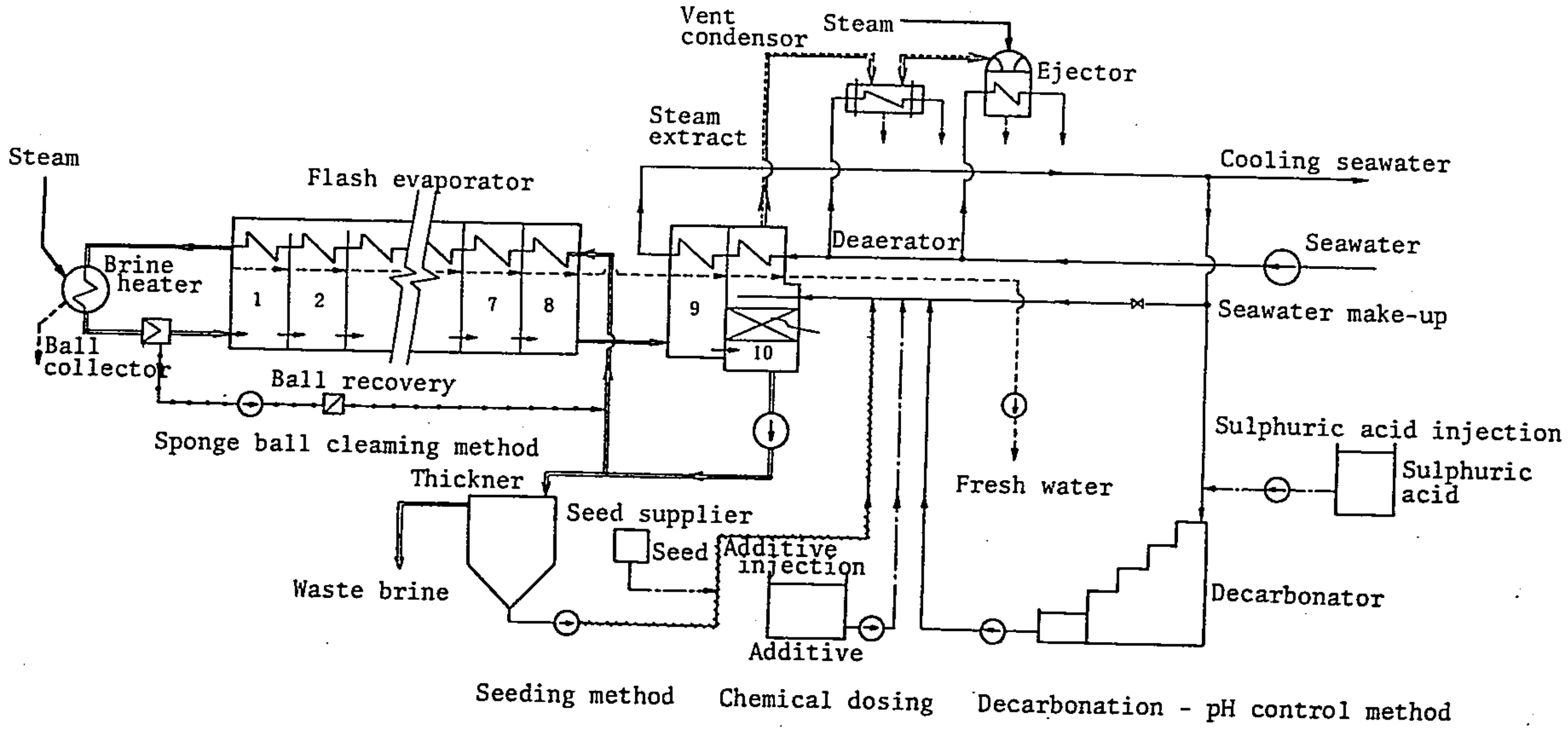


Fig. 5.2-4   Process diagram of scale prevention trial plant

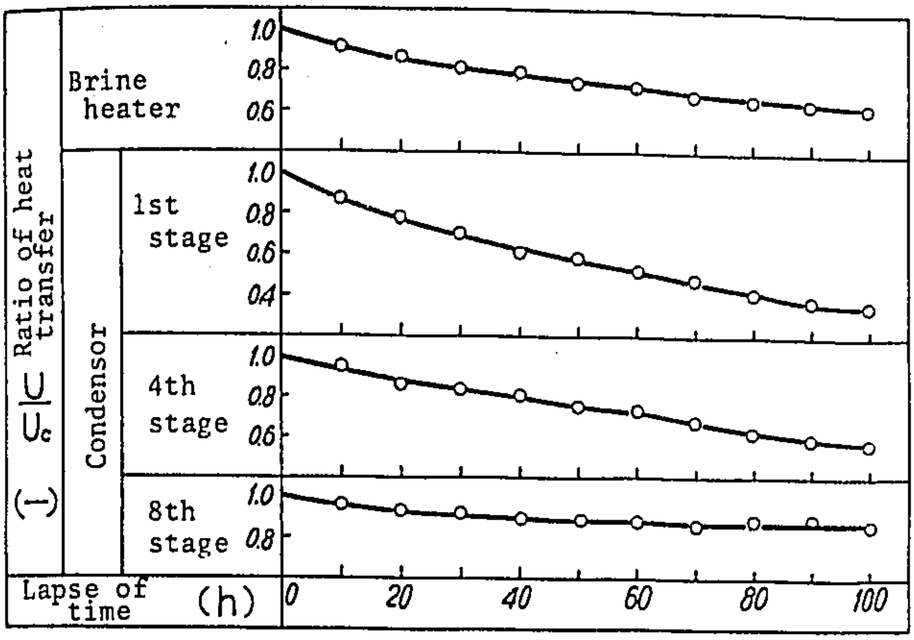


Fig. 5.2-5 Change of heat transfer coefficient

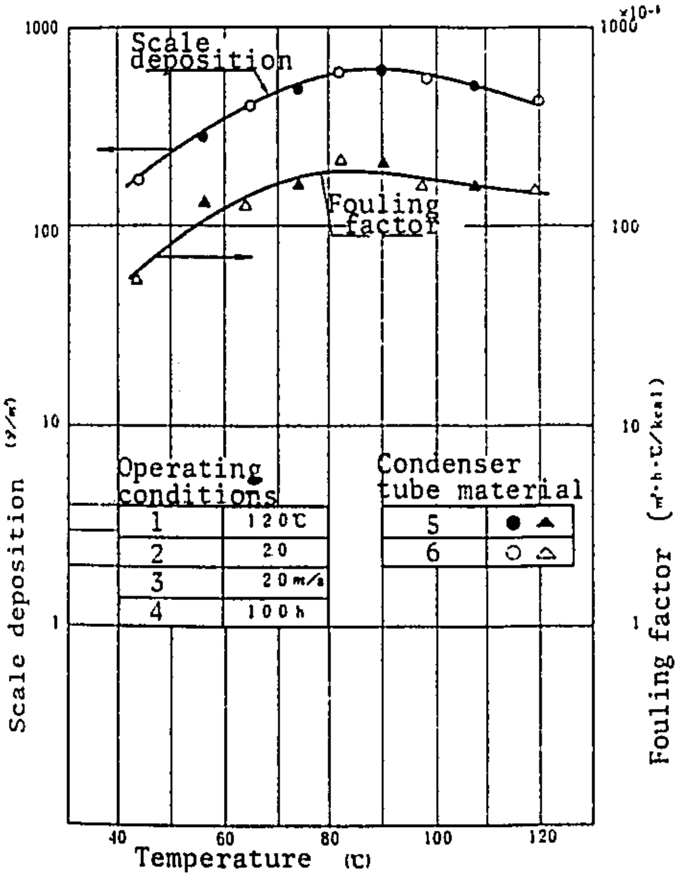


Fig. 5.2-6

Notes:

1. Max. temperature
2. Concentration factor
3. Flow rate
4. Operating hours
5. Alminium brass
6. Titanium

Fig. 5.2-6 Scale deposition and fouling factor in non-treatment test

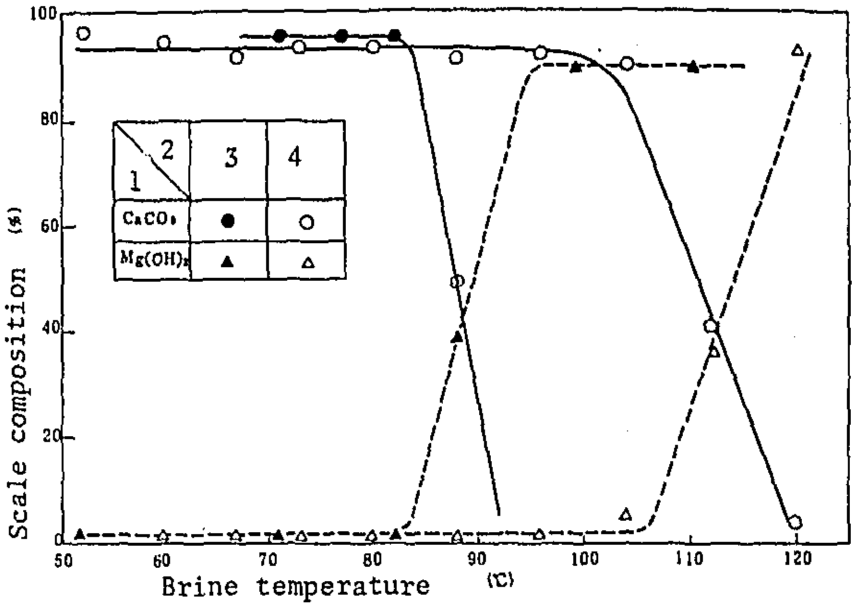
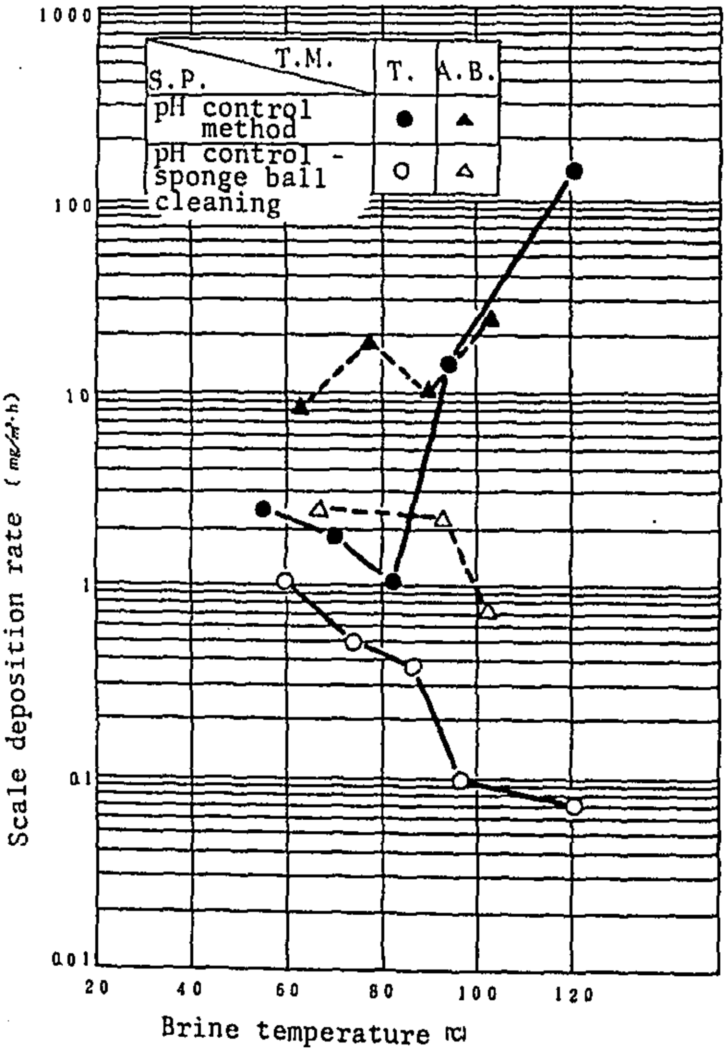


Fig. 5.2-7
Notes:
1. Scale
2. System
3. Submerged tu
type evapora
4. Flash evapor

Fig. 5.2-7 Alkaline scale composition at various temperature



Note:
S.P.: Scale prevention
T.M.: Tube material
T.: Titanium
A.B.: Alminium brass

Fig. 5.2-8 Scale deposition rate by long-term (2,000 hr) continuous operation

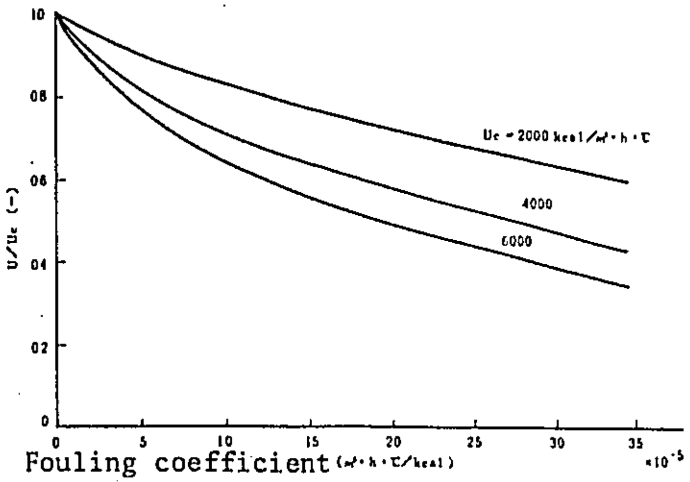


Fig. 5.2-9 Fouling factor and heat transfer coefficient

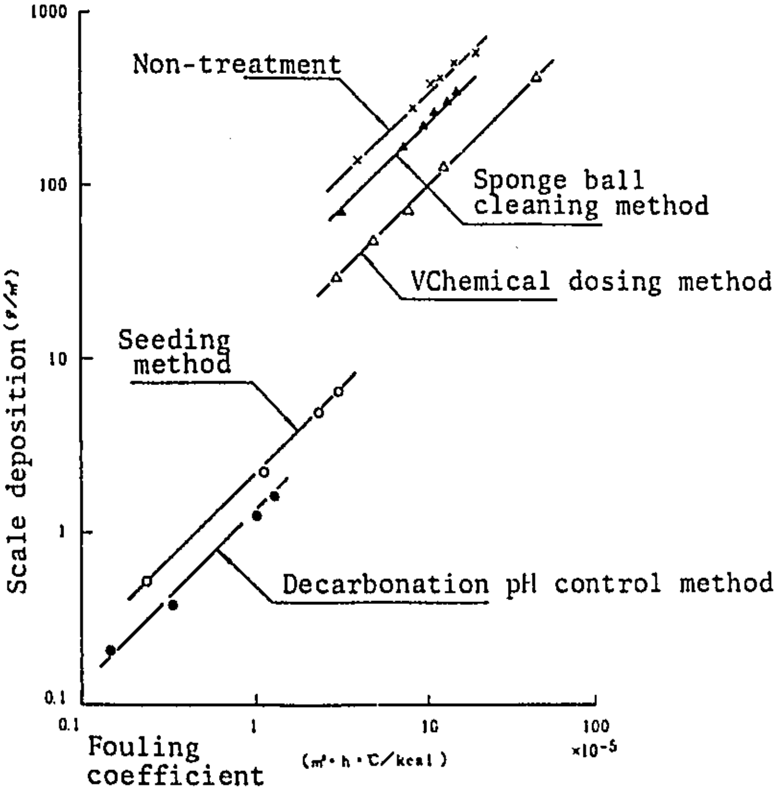
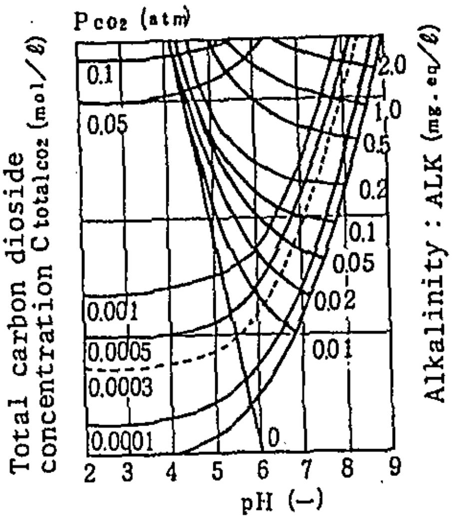


Fig. 5.2-10 Scale deposition vs. fouling factor



Temperature: Ts = 40℃
Cℓ⁻concentration : Cℓ = 19‰

Fig. 5.2-11 Correlation between total carbon dioxide and pH in seawater

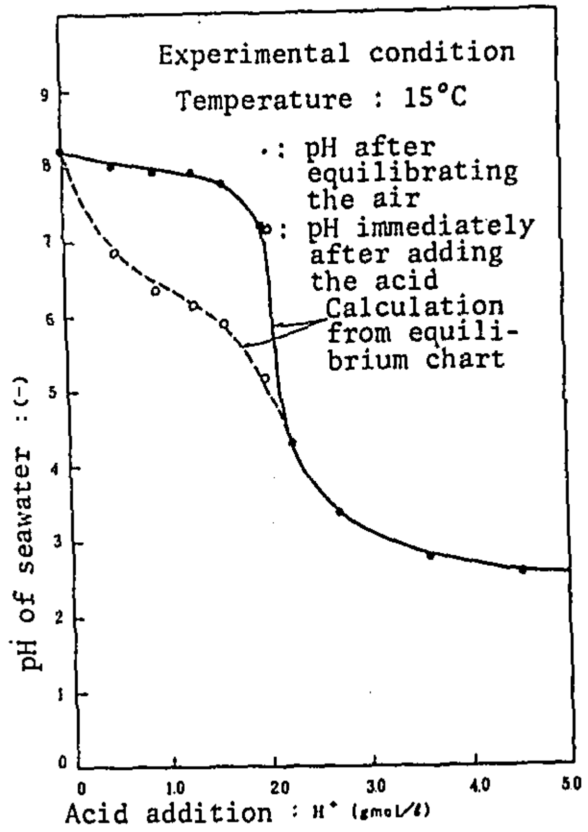
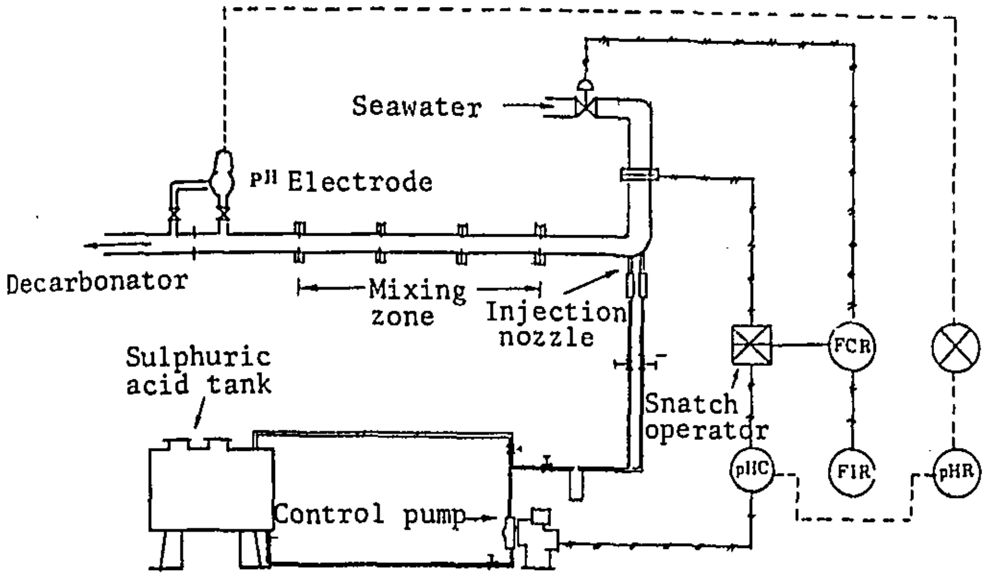
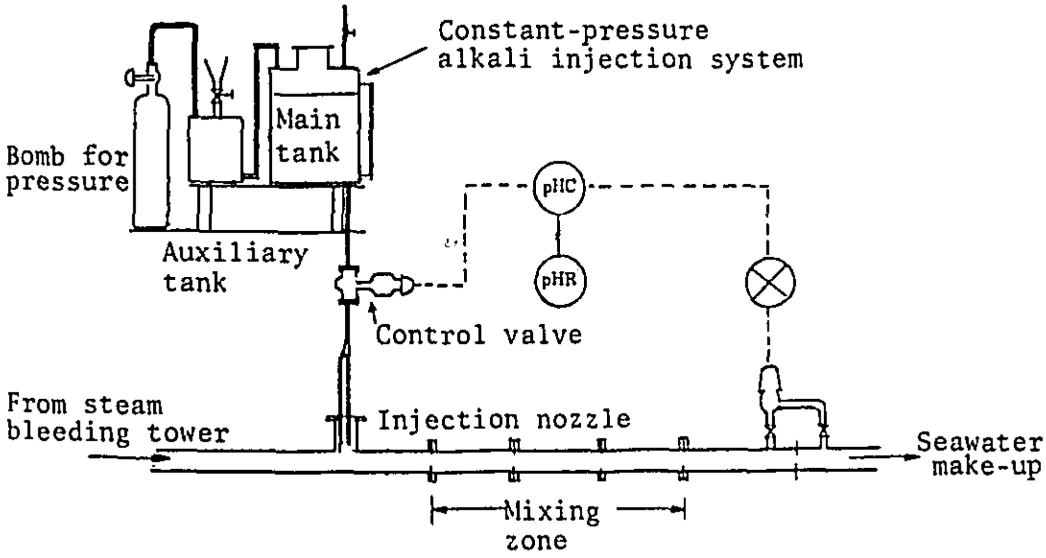


Fig. 5.2-12 Seawater pH vs. equilibrium value by addition of acid

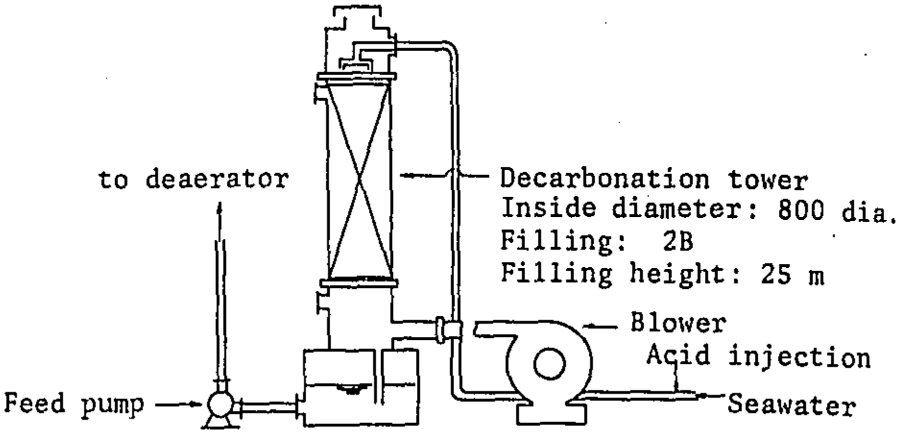


(a) Acid injection control

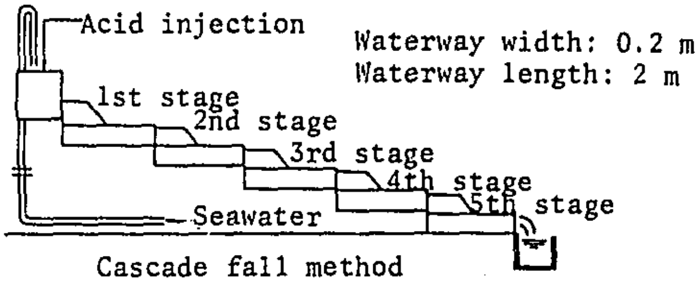


(b) Alkali control system

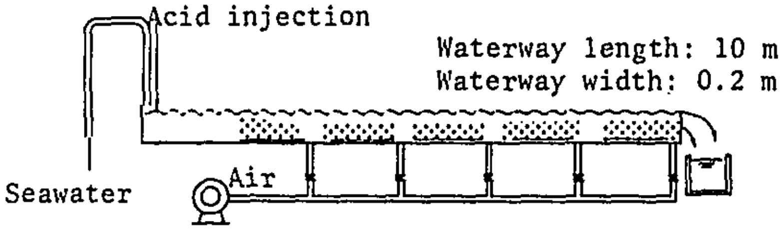
Fig. 5.2-13 Process diagram of pH control system



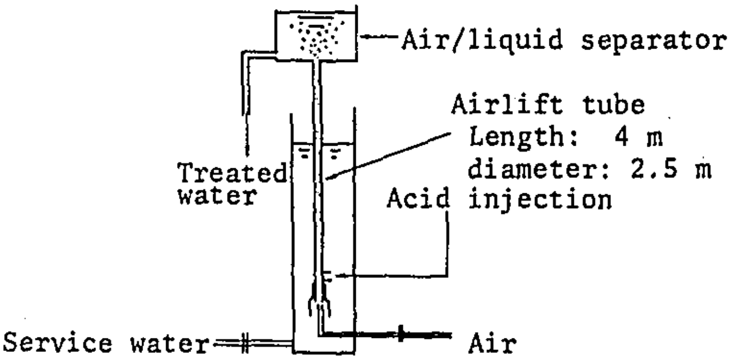
Packed tower



Cascade fall method



Air bubbling method



Airlift method

Fig. 5.2-14 Diagram of some decarbonation systems

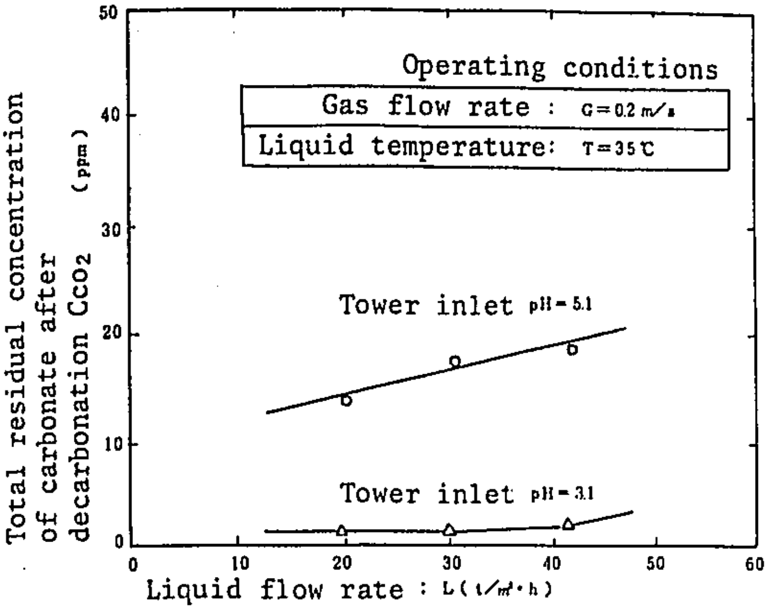


Fig. 5.2-15 Decarbonation performance of packed tower

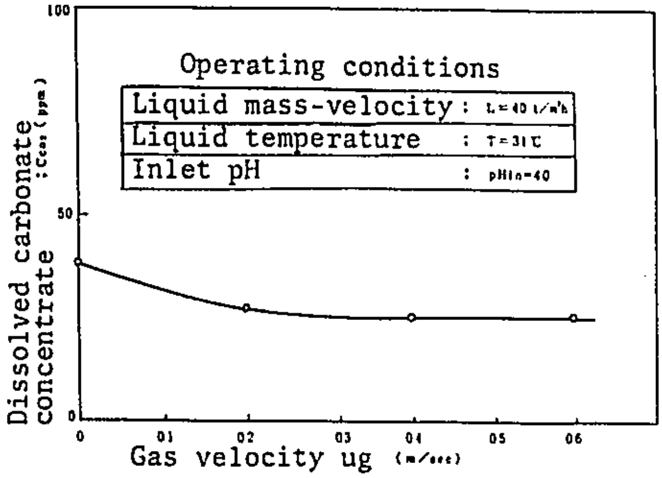


Fig. 5.2-16 Decarbonation performance of spray tower

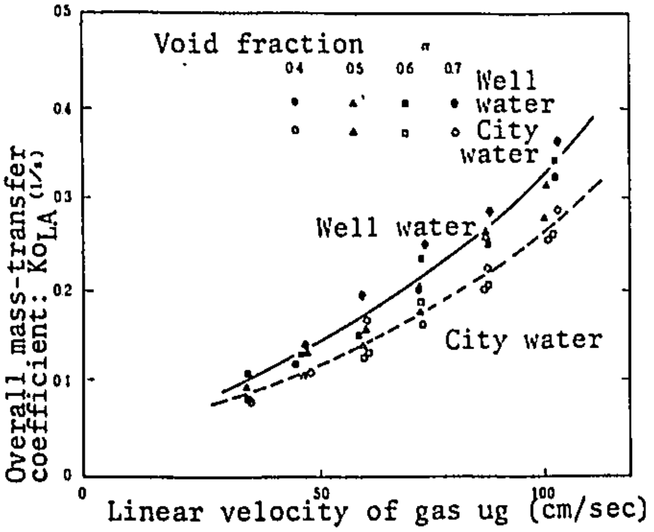


Fig. 5.2-17 Decarbonation performance of airlift method

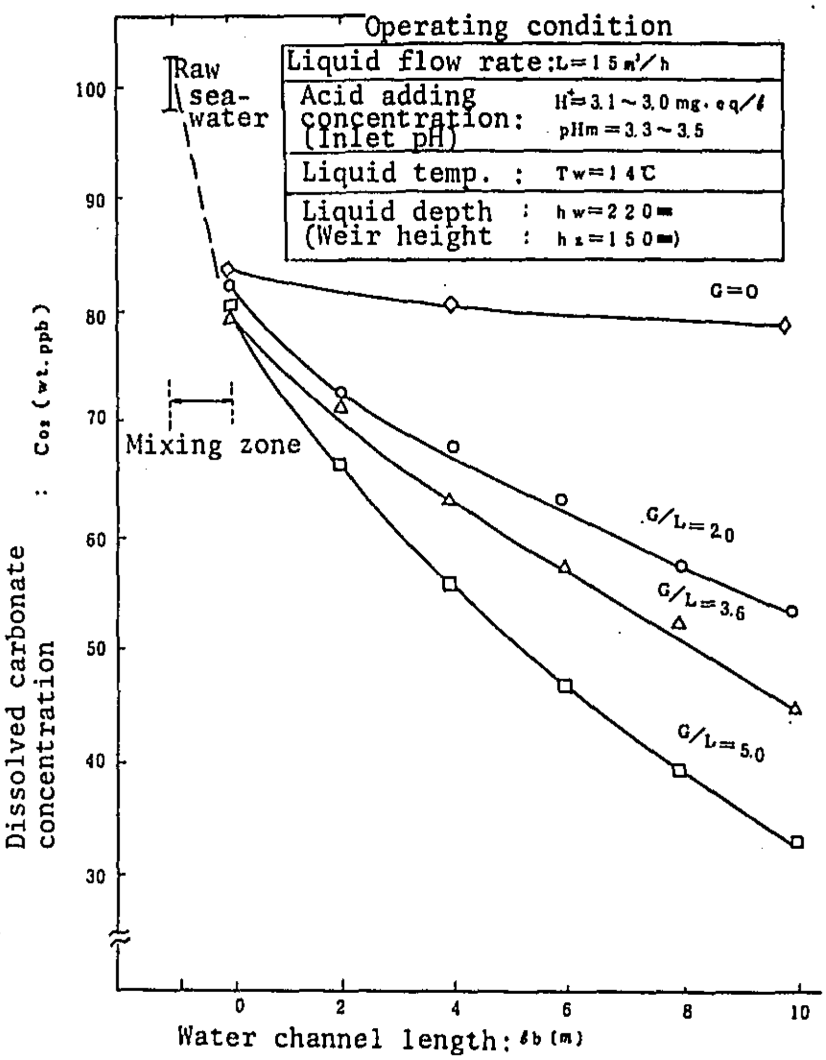
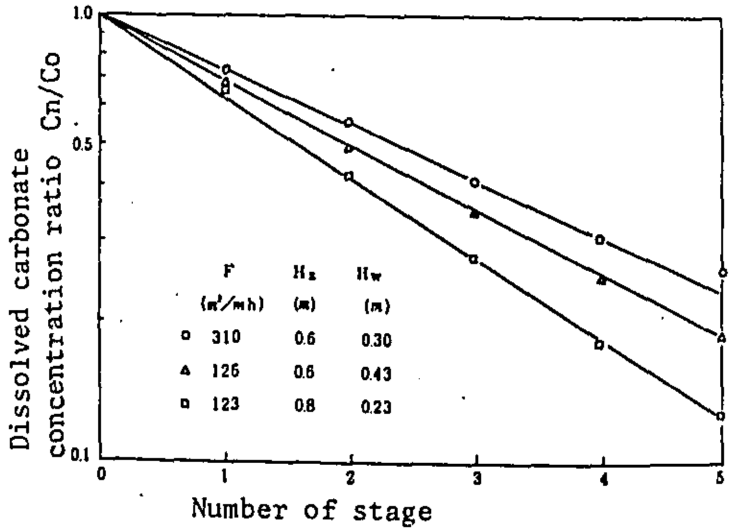


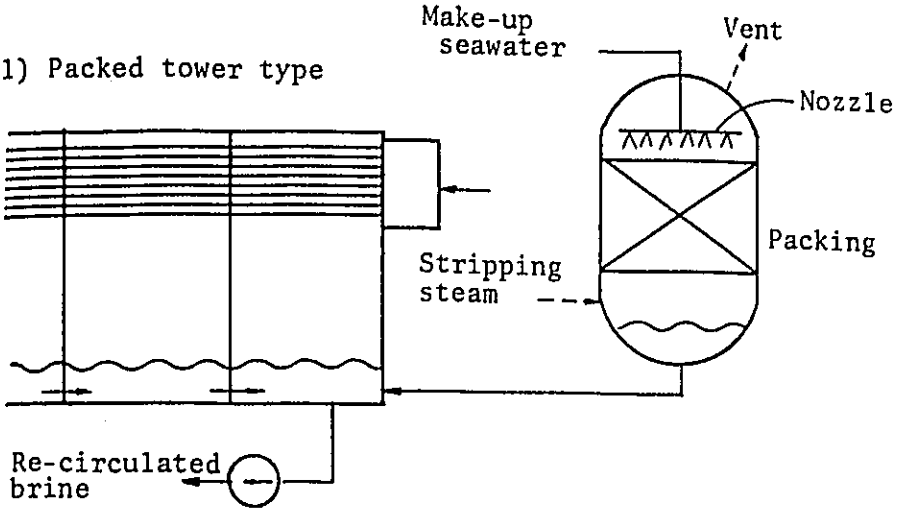
Fig. 5.2-18 Decarbonation performance of air bubblin method



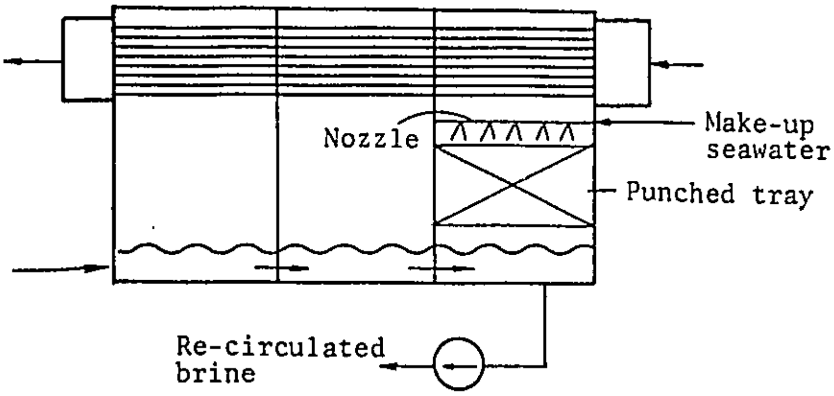
F ; Flow rate per unit width
Hs; Spacing between individual stages
Hw; Liquid depth
C ; Dissolved carbonate concentration

Fig. 5.2-19 Decarbonation performance of multi-stage cascade fall method

(1) Packed tower type



(2) Final-stage deaeration type



(3) Heat rejection stages deaeration type

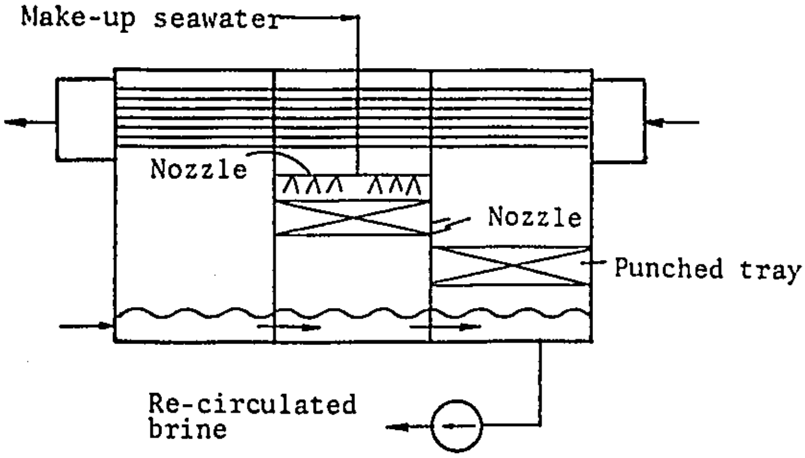
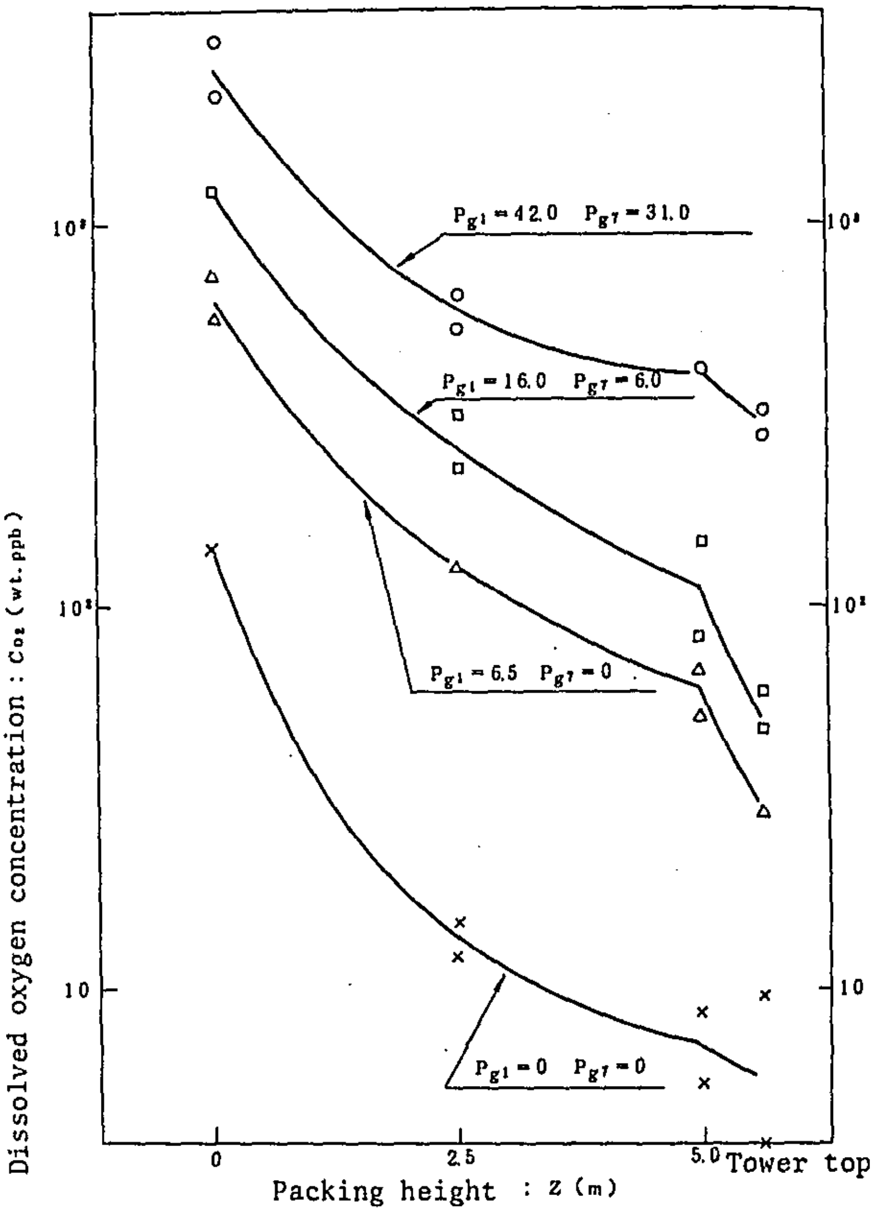


Fig. 5.2-20 Schematic diagram of deaeration test plant



Operating conditions

Liquid mass-velocity : L = 40t/m^{2}h
Liquid temperature : Te$_1$ = 33 45°C
Stripping steam:
Steam: G = 90 170 kg/h
Partial-pressure in tower : Pg
Pg$_1$ = Partial pressure at the top of tower
Pg$_7$ = Partial pressure at the bottom of tower

Fig. 5.2-21 Deaeration performance of packed tower

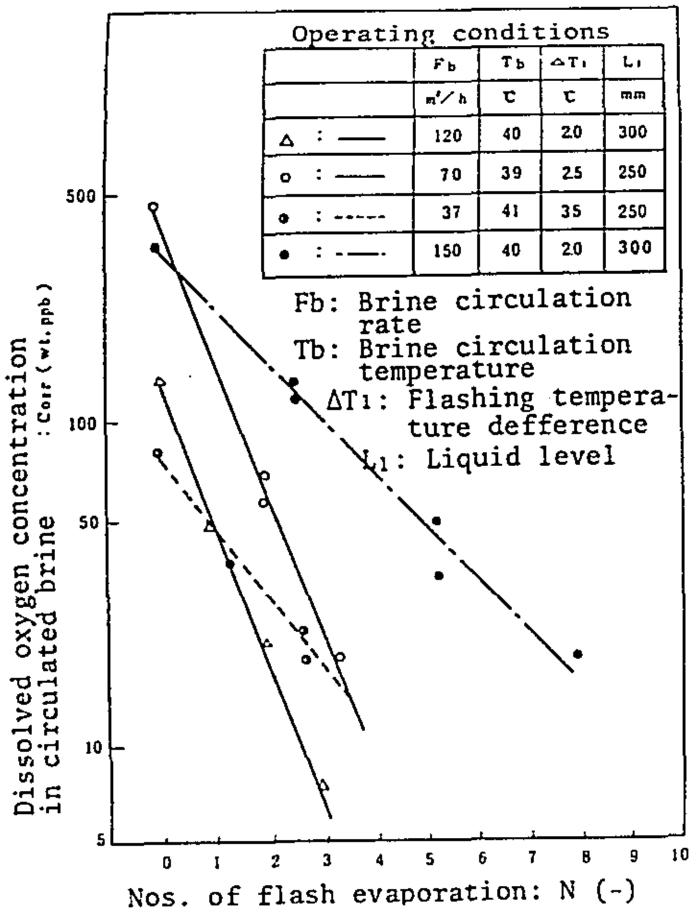


Fig. 5.2-23 Deaeration performance of heat rejection stages

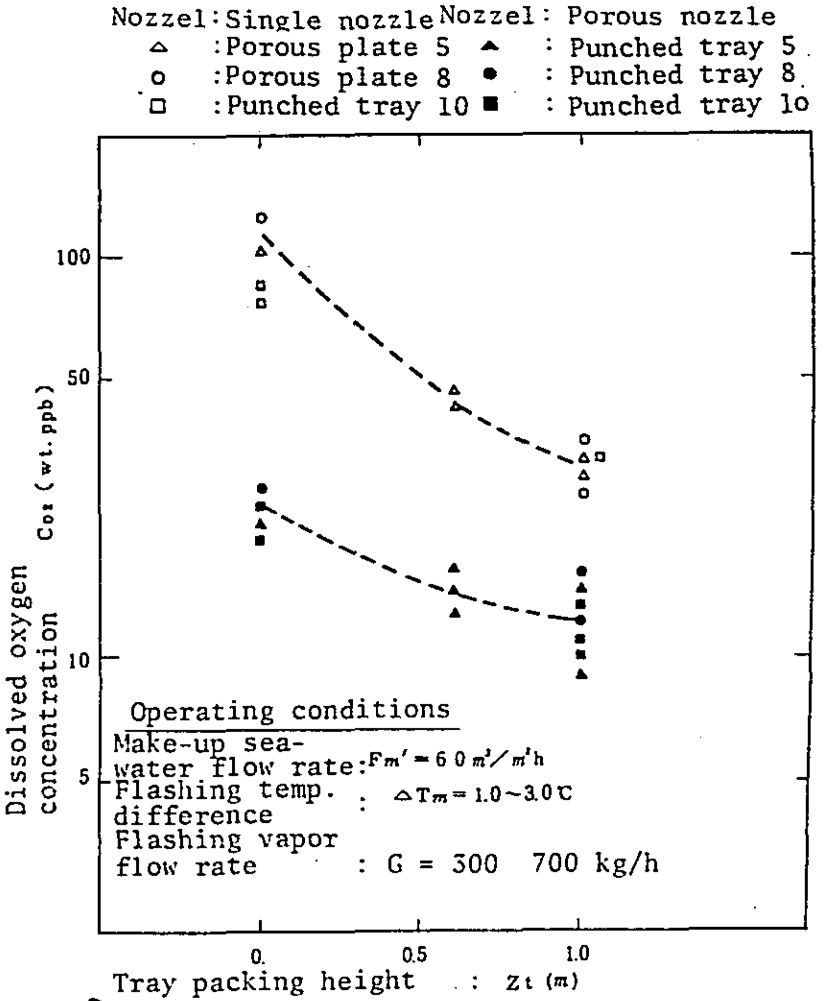
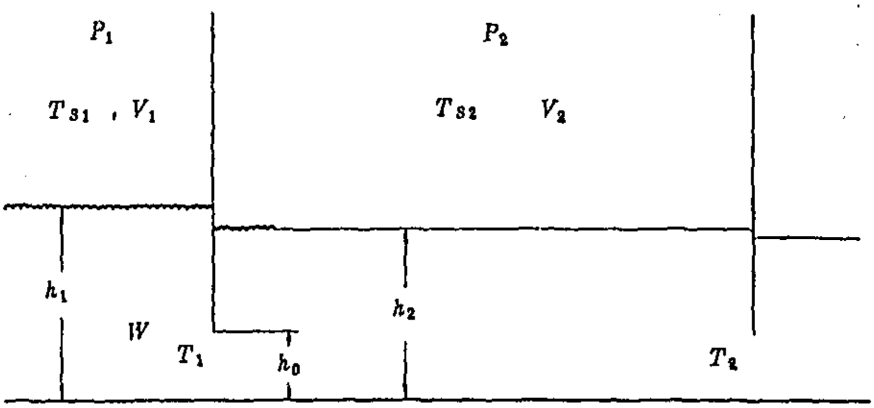


Fig. 5.2-22 Deaeration performance of final stage



h :Height (m)

P : Pressure (Kg/m^2)

T :Brine temperature(°C)

T_S: Steam temperature (°C)

V :Specific volume of steam (m^3/Kg)

ρ : Brine density (Kg/m^3)

Fig. 5.3-1 State of brine and steam between stages and explanation of symbols

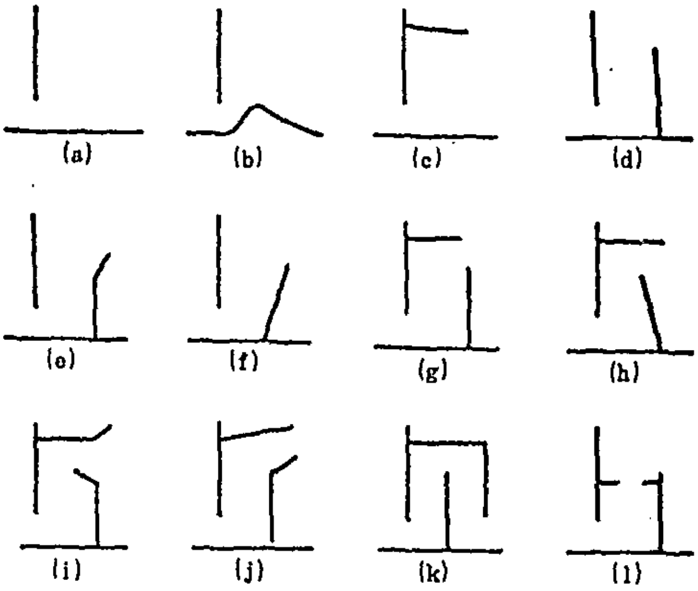


Fig. 5.3-2 Various configurations of orifice section at flash chamber inlet

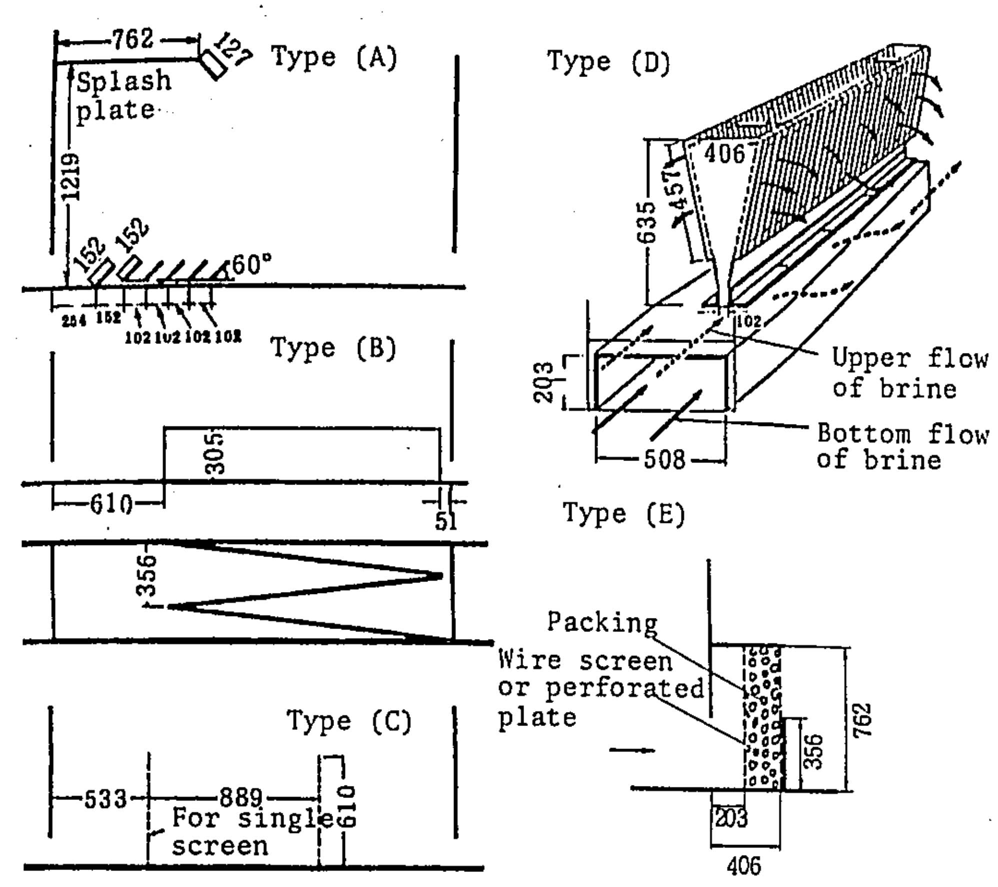


Fig. 5.3-3 Various types of evaporation booster

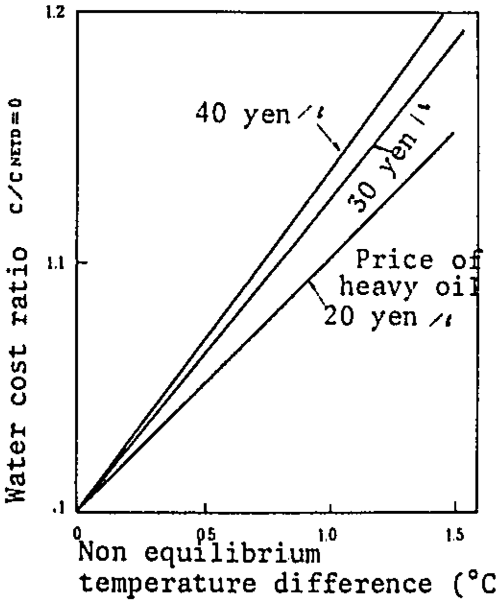


Fig. 5.3-4 Effect of non equilibrium temperature difference on water cost

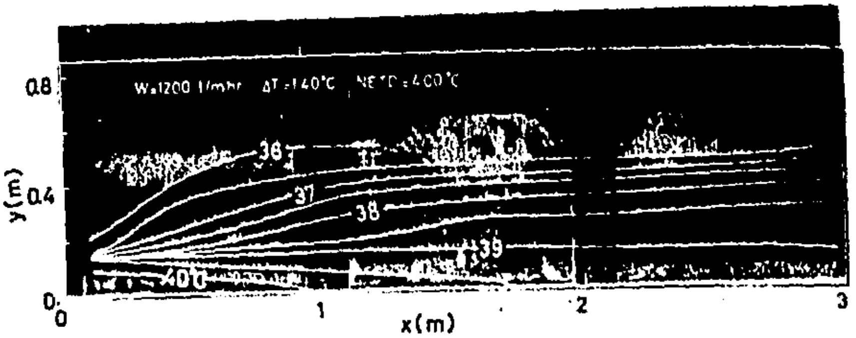


Fig. 5.3-5 State of flash evaporation and temperature distribution

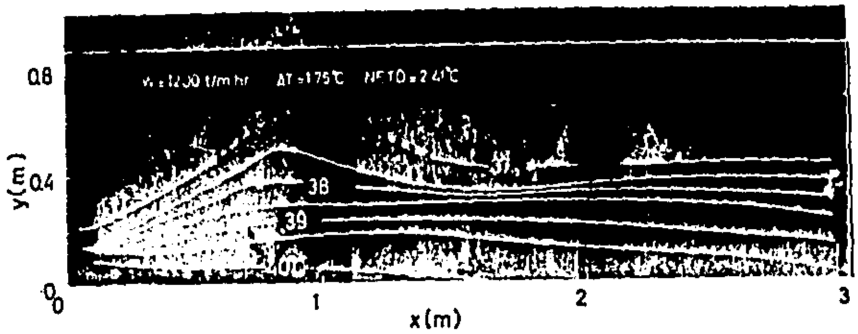


Fig. 5.3-6 State of flash evaporation and temperature distribution

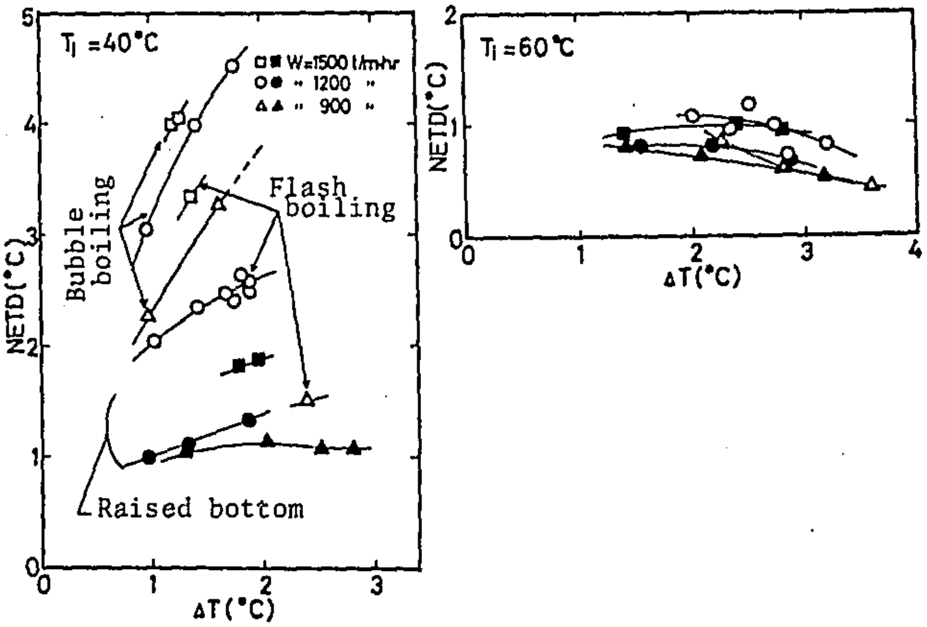


Fig. 5.3-7 Correlation between non equilibrium temperature difference and flash temperature difference

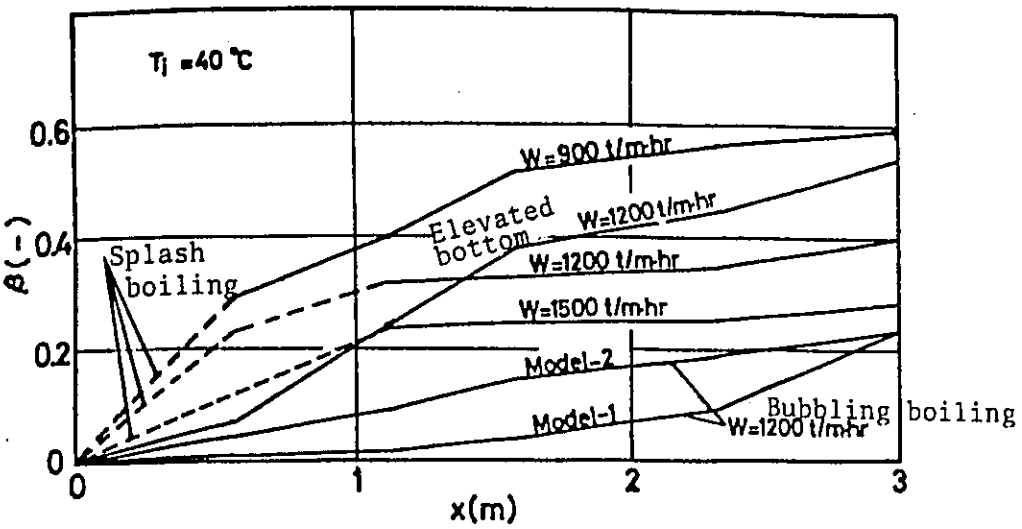


Fig. 5.3-8 Variations inllocal equilibrium attainment of flow direction

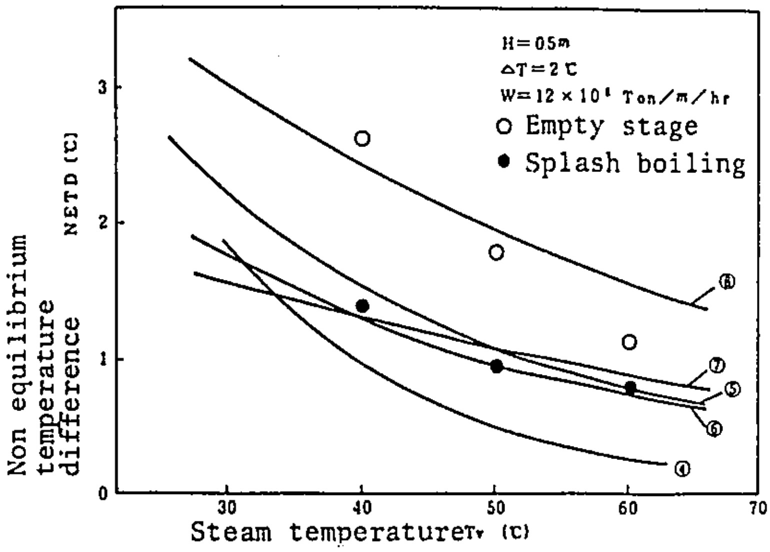


Fig. 5.3-9 A comparison of empirical formulae of non equilibrium temperature difference

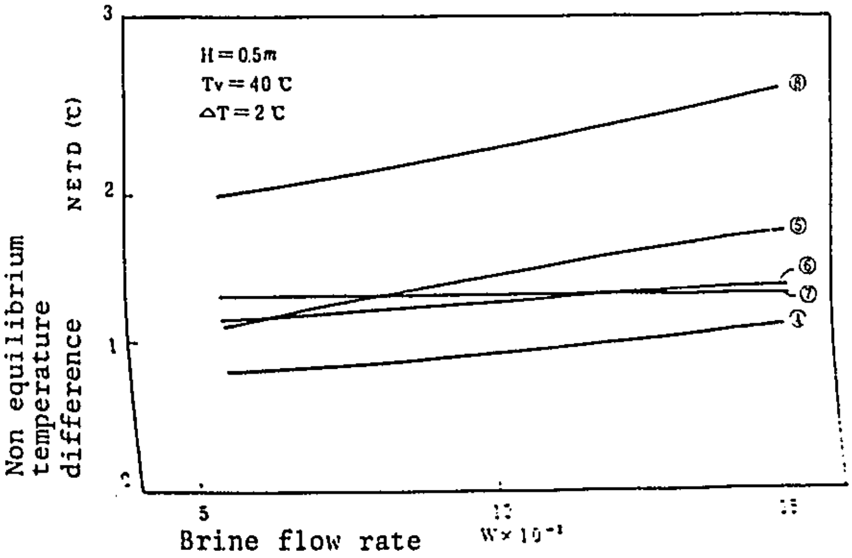
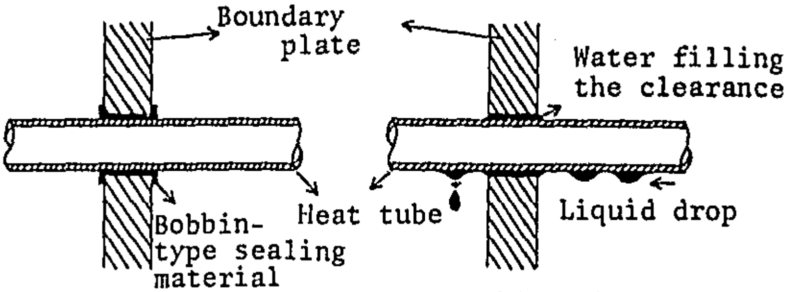
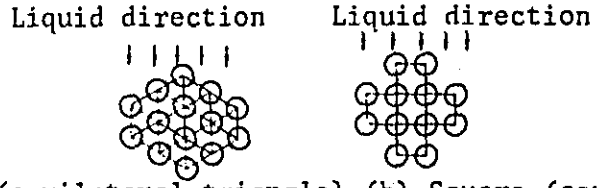


Fig. 5.3-10 A comparison of empirical formulae of non equilibrium temperature difference

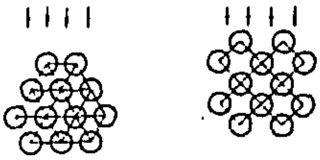


Bobbin-type seal Liquid seal

Fig. 5.3-11 Interstage boundary seal, diagram



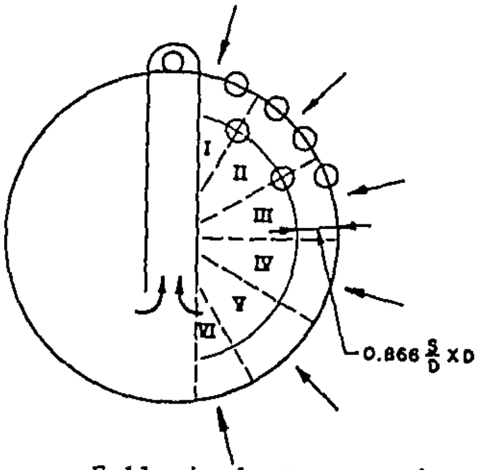
(a) Triangle (equilateral triangle) configuration (series type) (b) Square (square) configuration(series type)

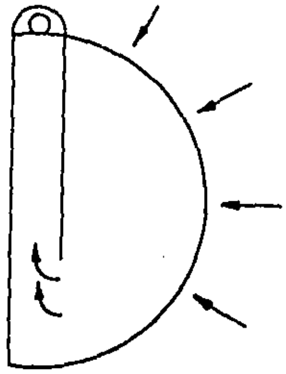


(c) Triangle (equilateral triangle) configuration (d) Square (square) configuration

Note: (c), (d) are staggered tube toward liquid direction

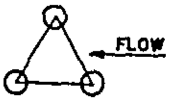
Fig. 5.3-12 Tube assortment and pitch (unit: mm)



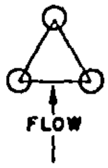


Semi-circular type condensor

Full-circle type condensor

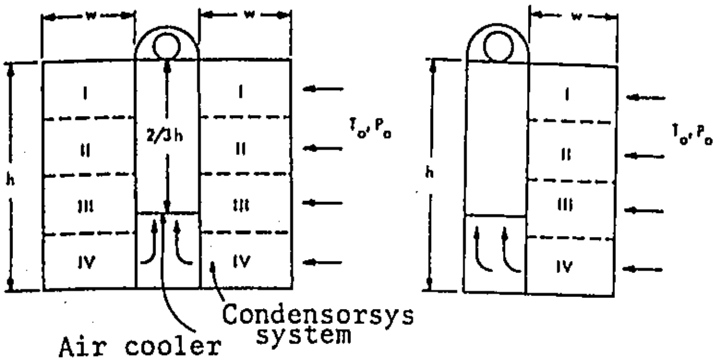


Typical flow of condensor system



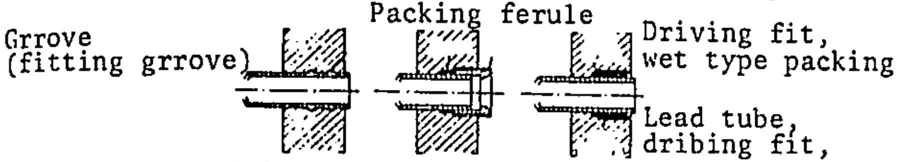
Typical flow of air cooler system

Fig. 5.3-13 Circular type condensor, nest of tubes

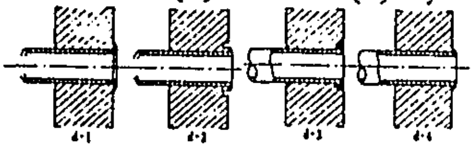


Full-rectangular type condensor   Semi-rectangular type condensor

Fig. 5.3-14 Nest of tubes, of rectangular type condensor



(a) Expanded tube (b) Ferule (c) By Y-type packing



(d) By welding

Fig. 5.3-15 Different installation of heat tubes

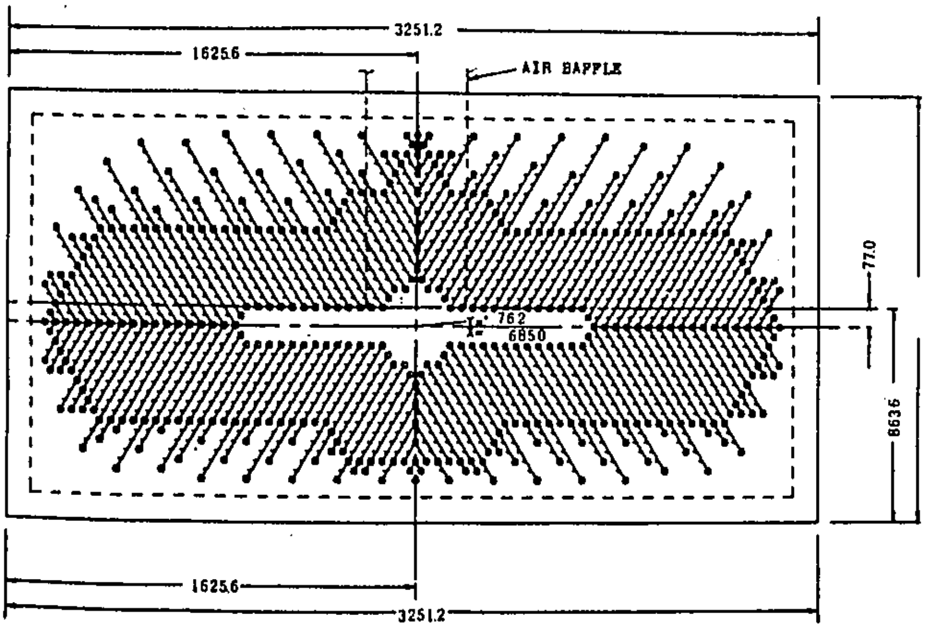


Fig. 5.3-16 Tube plate arrangement of preheater of VTE/MSF desalting plant with a capacity of 757,000 m^3/day (35 stages, 1.375" in outside diameter, 0.028" wall thickness, flow rate of 7.68 ft/s, effective length of 22.75 ft, titanium tube)

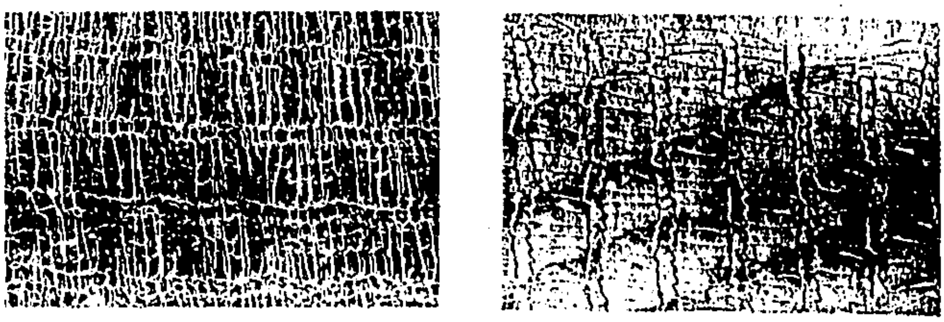


Fig. 5.3-17 Mesh elminator (ε = 0.99, Df = 0.25 mm)

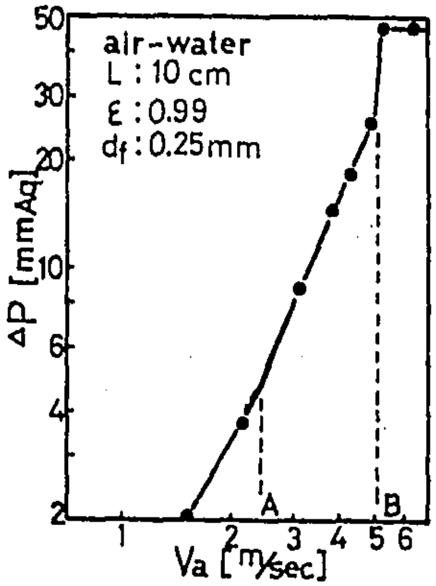


Fig. 5.3-18 Relation between gas velocity and pressure drop in series of water - air

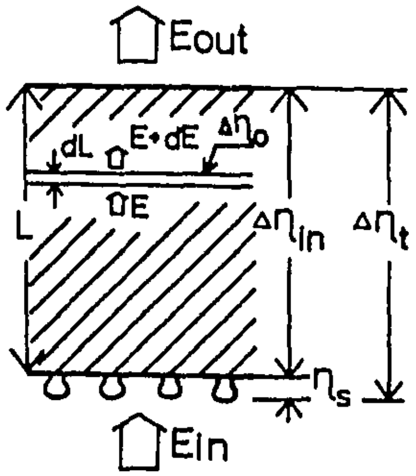


Fig. 5.3-19 A model for column efficiency of mesh eliminator

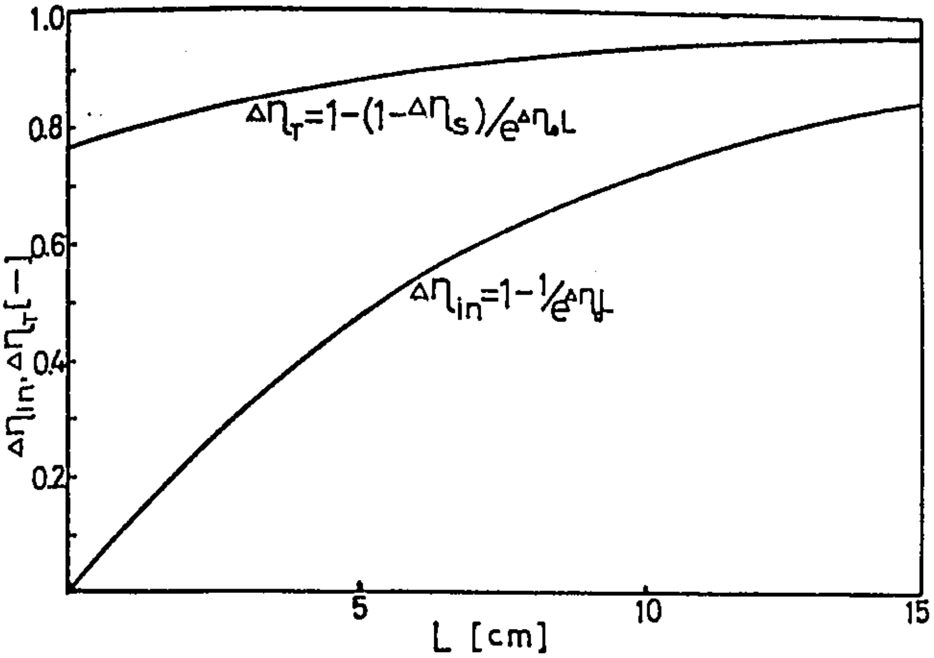


Fig. 5.3-20 Correlation between column efficiency and thickness of mesh eliminator with and without consideration of column efficiency underneath the eliminator

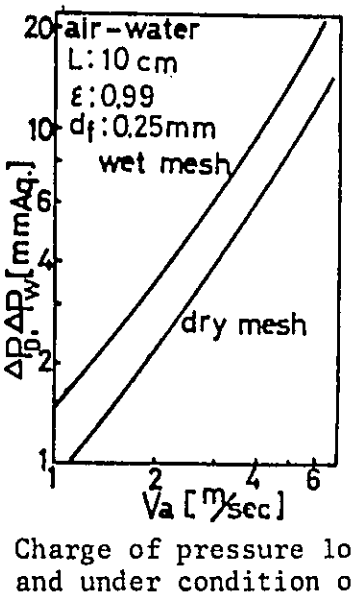
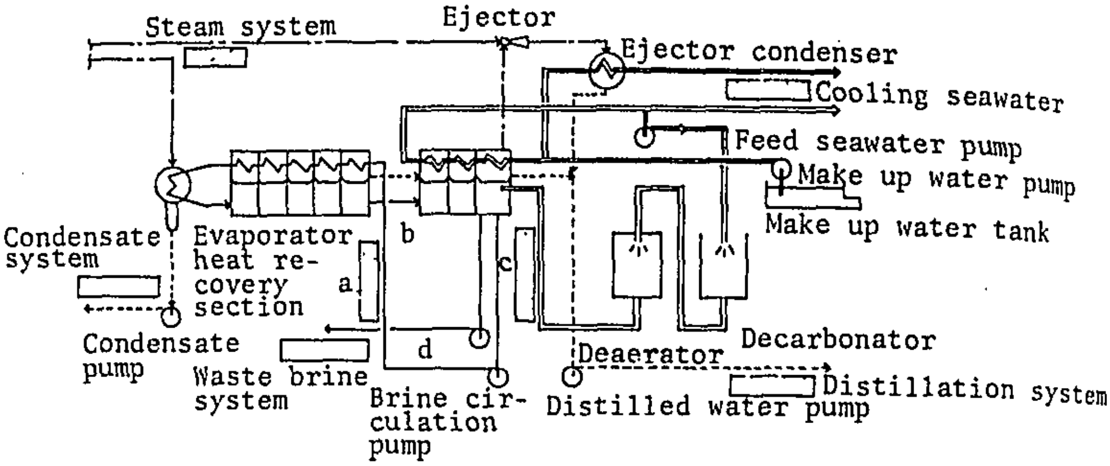


Fig. 5.3-21 Charge of pressure loss under dry condition and under condition of mist adhesion



Note: a: Circulating brine system
b: Evaporator heat rejection section
c: Waste brine system
d: Waste brine pump

Fig. 5.3-22 Distribution diagram of desalting plant using multistage flash distillation process

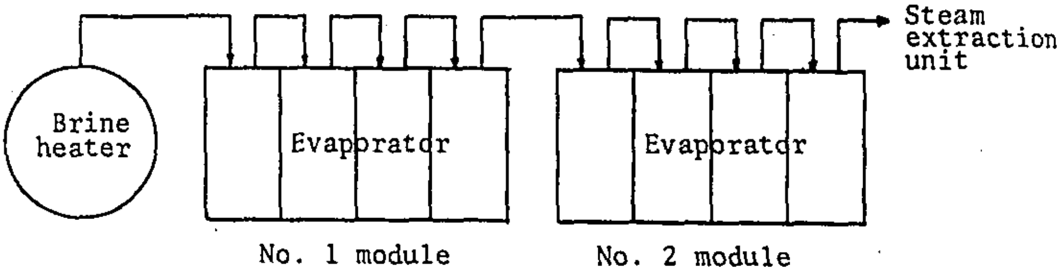


Fig. 5.3-23 Cascade steam extraction

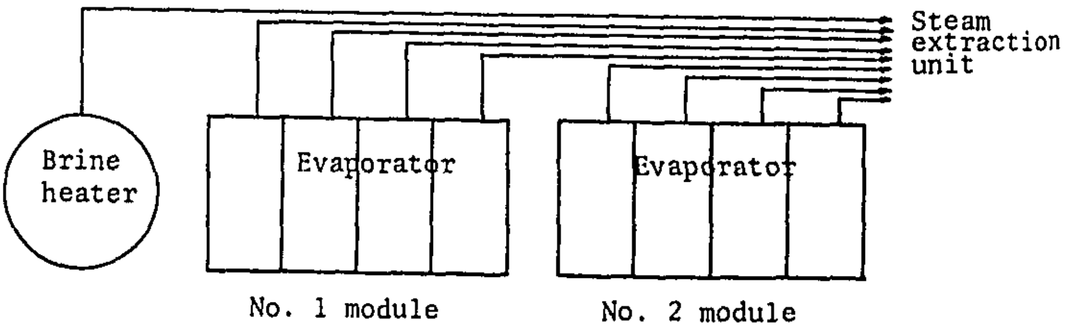


Fig. 5.3-24 Parallel steam extraction

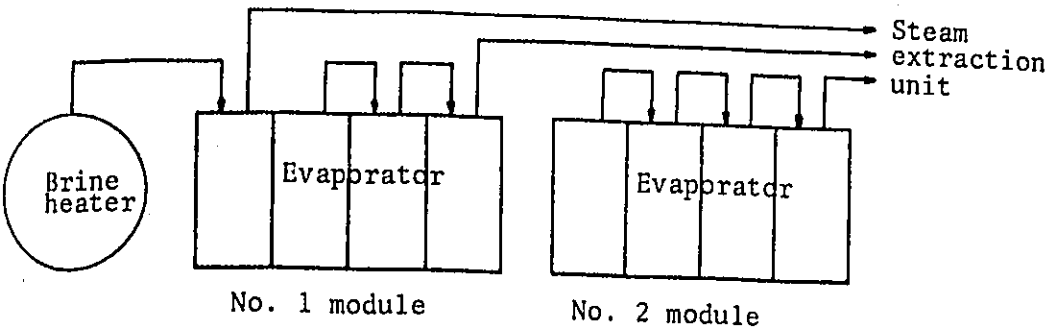


Fig. 5.3-25 Combined steam extraction

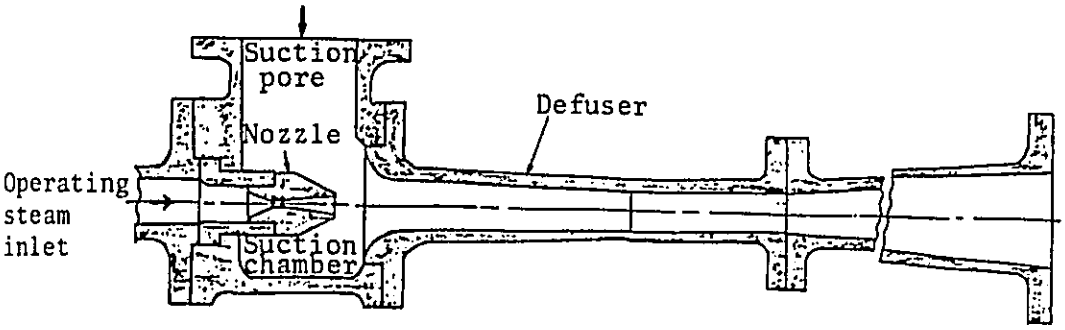


Fig. 5.3-26 Steam ejector

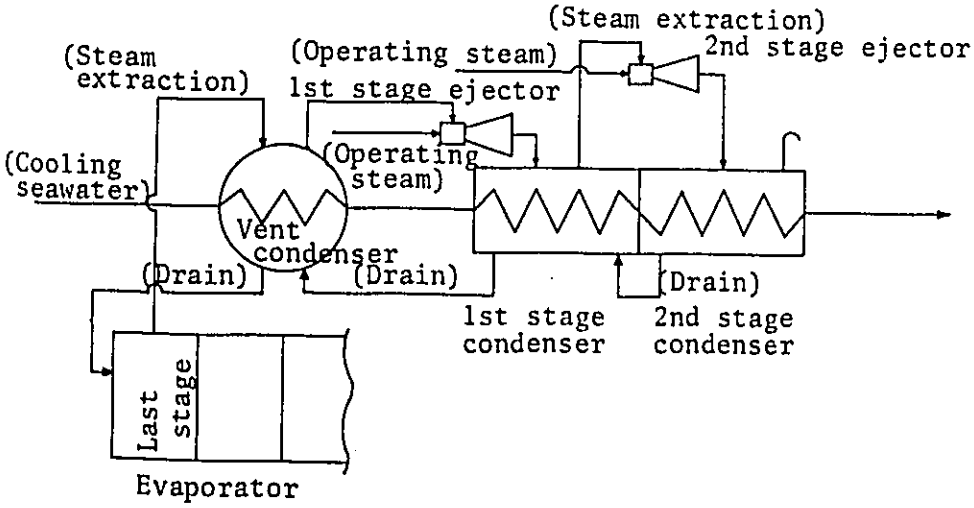


Fig. 5.3-27 Flow chart of vent condenser

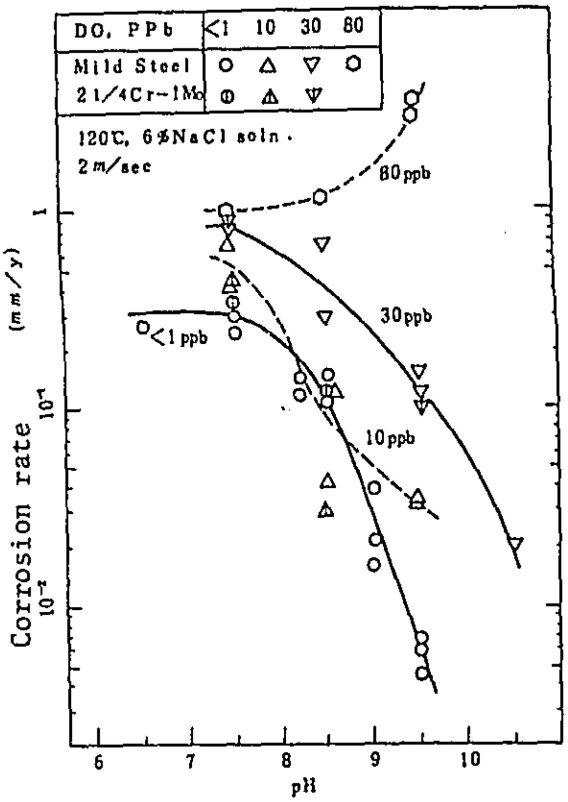


Fig. 5.4-1 Effect of pH and dissolved oxygen on corrosion rate

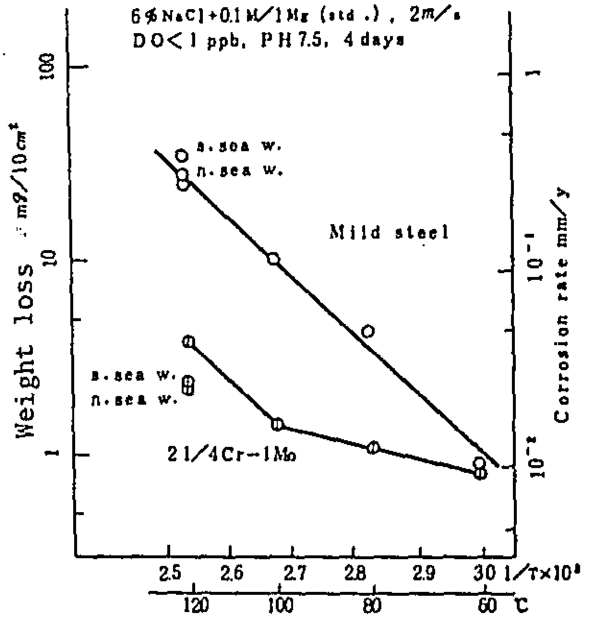
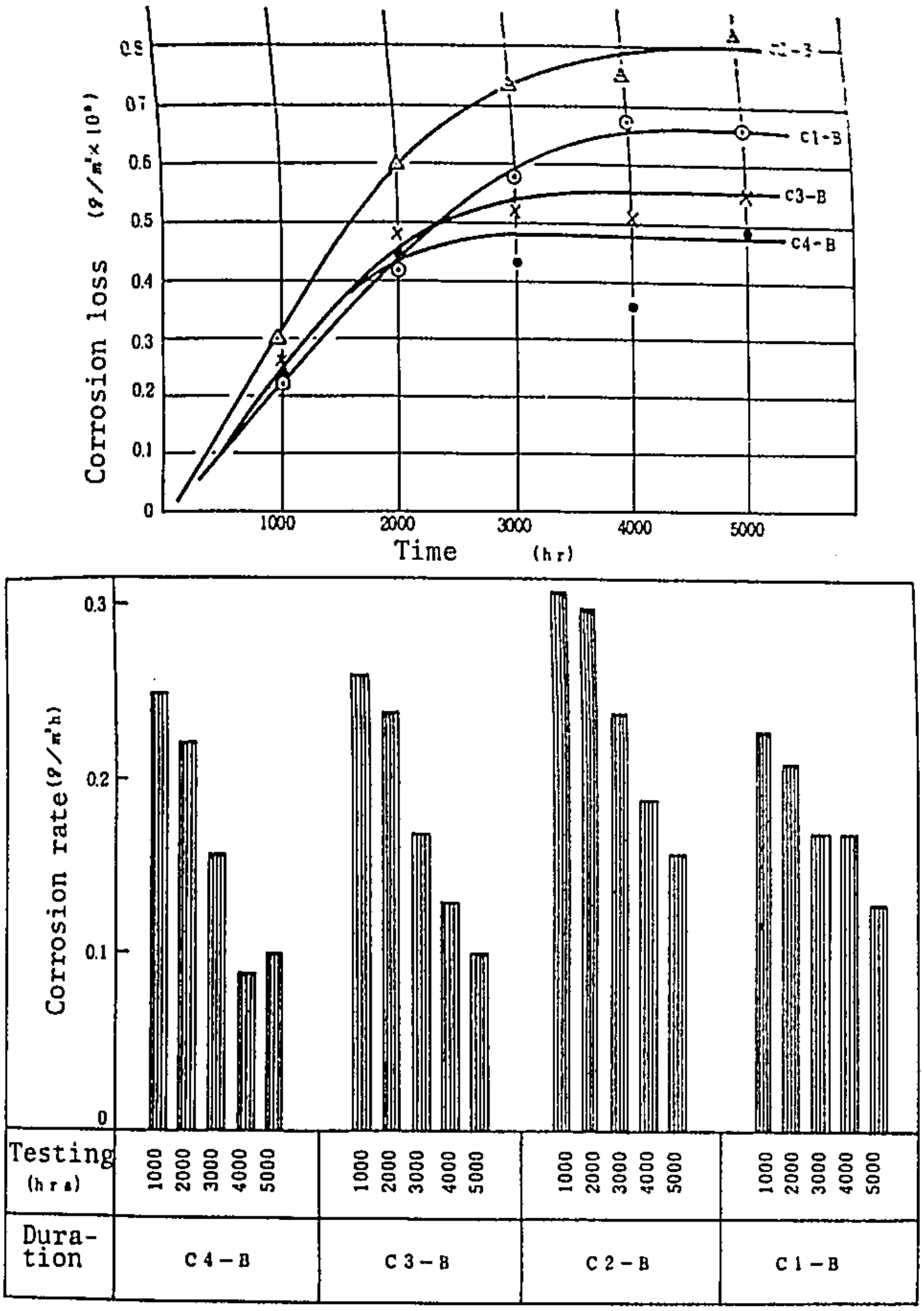


Fig. 5.4-2 Effect of temperature on corrosion of steel

s.sea w.: Synthetic sea water n.sea w.: Natural sea water



Run No.	Temp.	D.O	C.F.	PH	Flow rate	Time
6-B	90~55℃	<10ppb	1.7~1.8	7	1.8m/sec	1000hr ~5000hr

C.F.: Concentration factor

Fig. 5.4-3 Results of 5,000 hr field test for 3Cr-1 Al steel

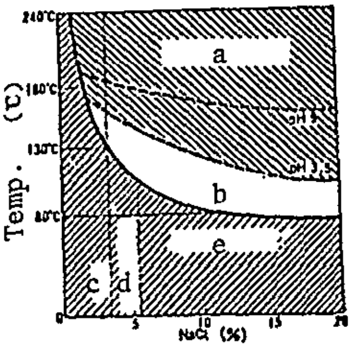


Fig. 5.4-4 Corrosion behavior of titanium for concentration and temperature of NaCl solution

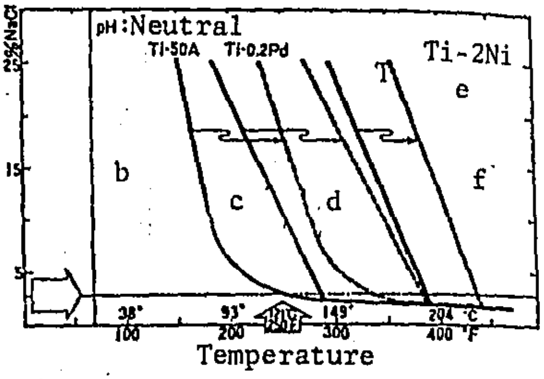


Fig. 5.4-5 Corrosion resistance of titanium, Ti-0.2Pd and Ti-2Ni in NaCl solution

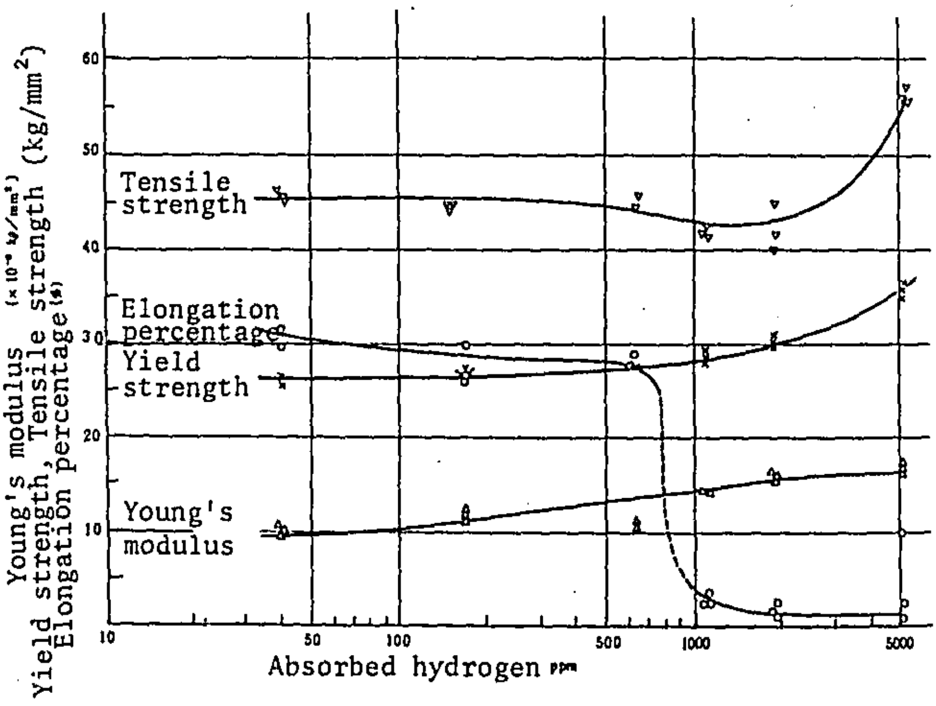


Fig. 5.4-6 Concentration of absorbed hydrogen and tensile characteristic of Titanium

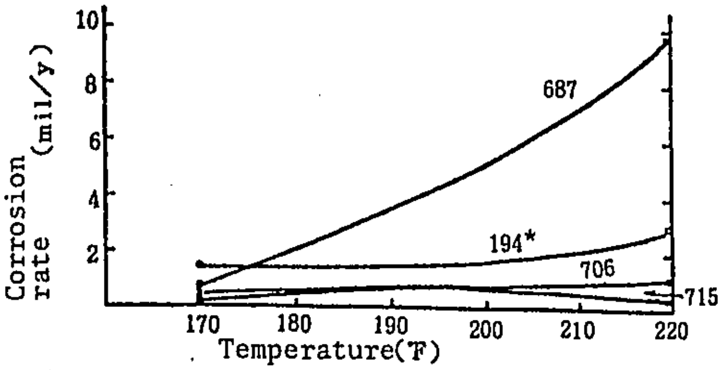


Fig. 5.4-7 Effect of temperature on corrosion rate of copper alloys DO200ppb, 90 days

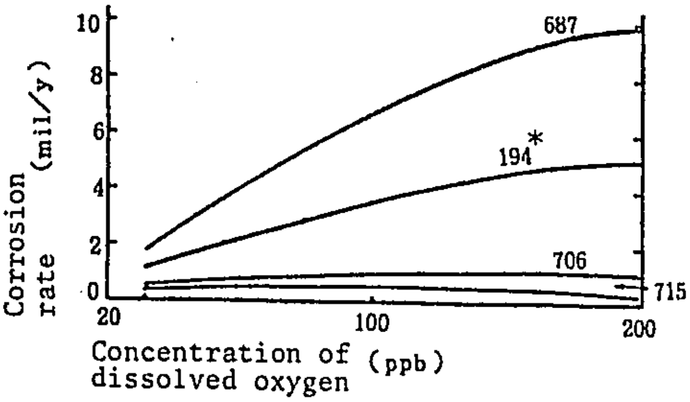
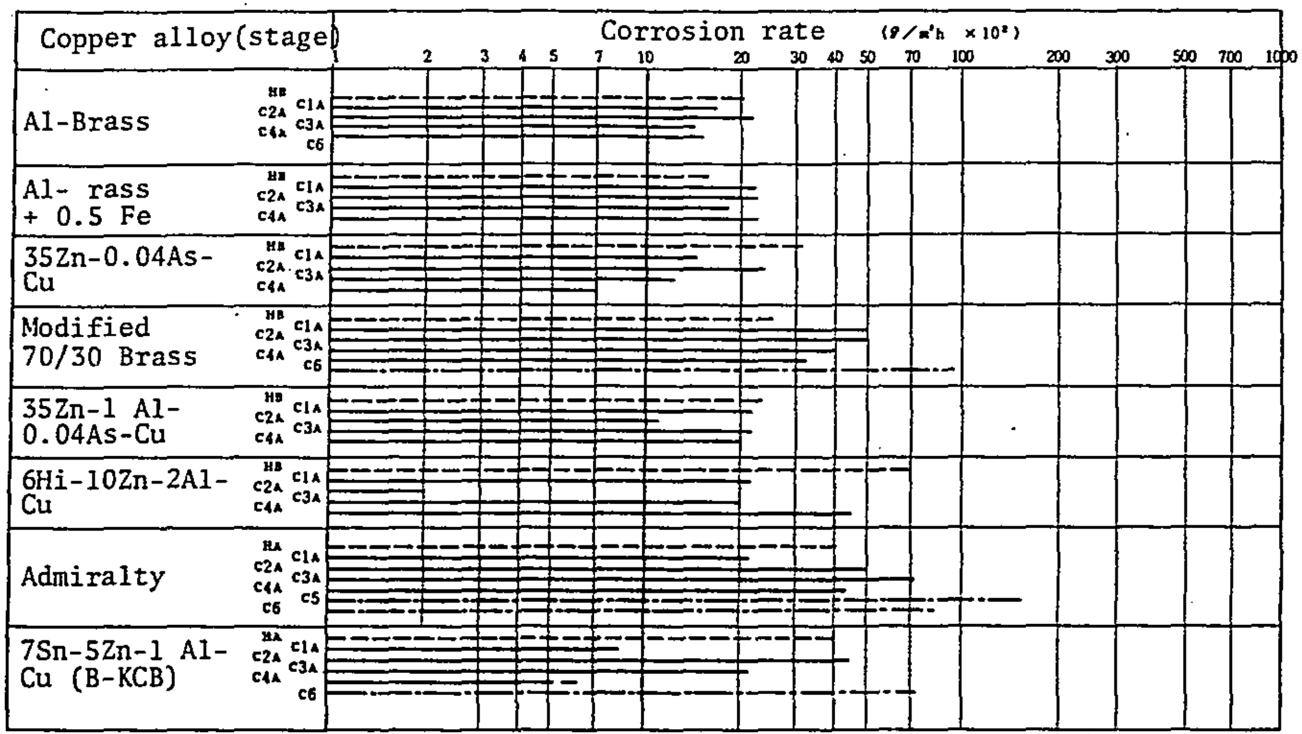


Fig. 5.4-8 Effect of dissolved oxygen on corrosion rate of copper alloys 114°C, 90 days

* CDA 194: Cu-2.3Fe



Note: 1) Concentration factor 1.66～1.70 DO 1ppb pH 6.6～7.6 2,000hrs (20m /d test plant)

2) Temperature	Outlet of heater A	Outlet of C1-A	Inlet of C4-A	Outlet of C5	Inlet of C6
	115 - 120°C	96 - 98°C	56 - 60°C	47 - 52°C	15 - 22°C

3) - - - Brine heater —— Heat recovery —·— Heat rejection

Fig. 5.4-9 Corrosion rate of copper alloys obtained from 2,000 hr field test

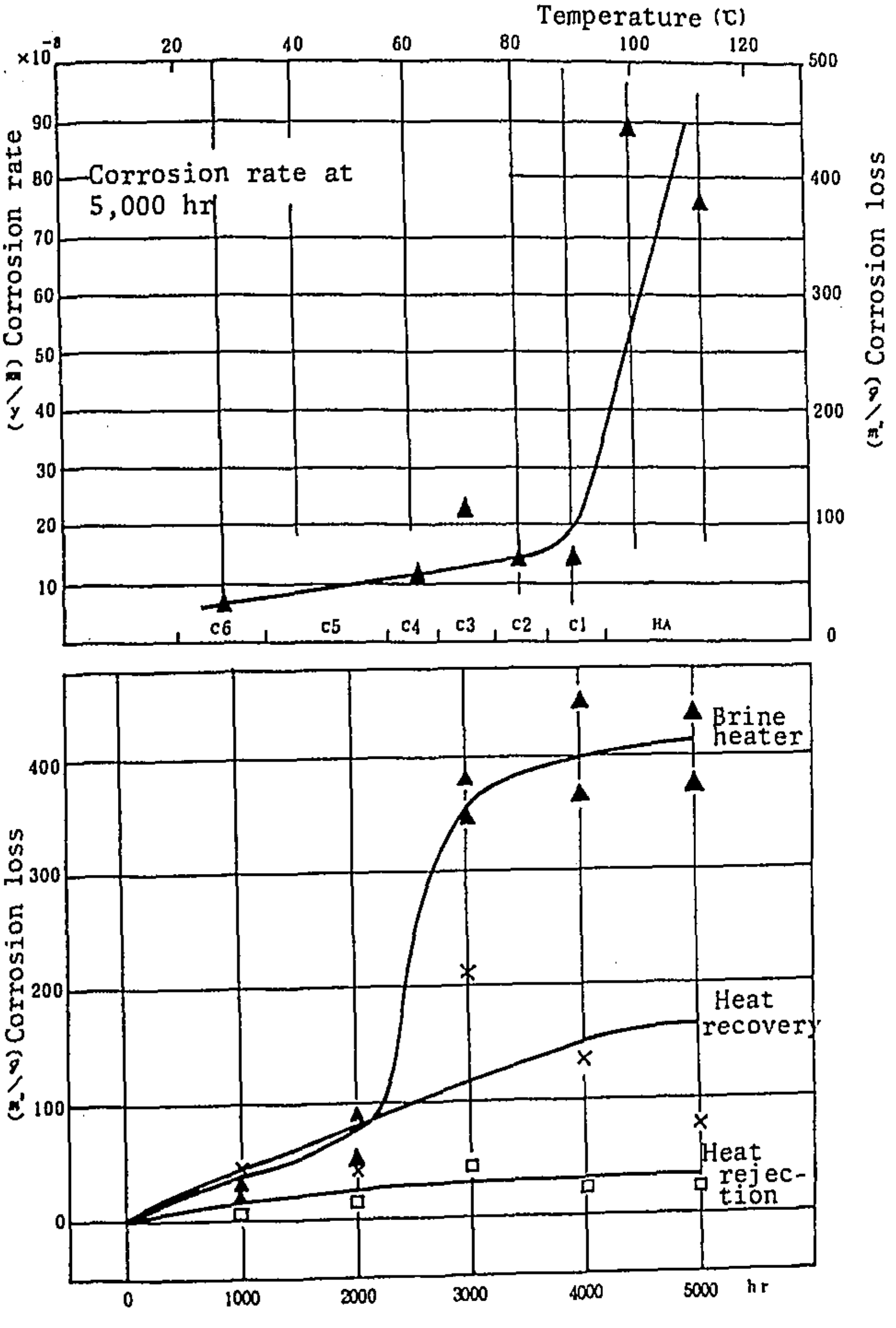


Fig. 5.4-10 Effect of section (temperature) and test duration on corrosion of aluminum brass

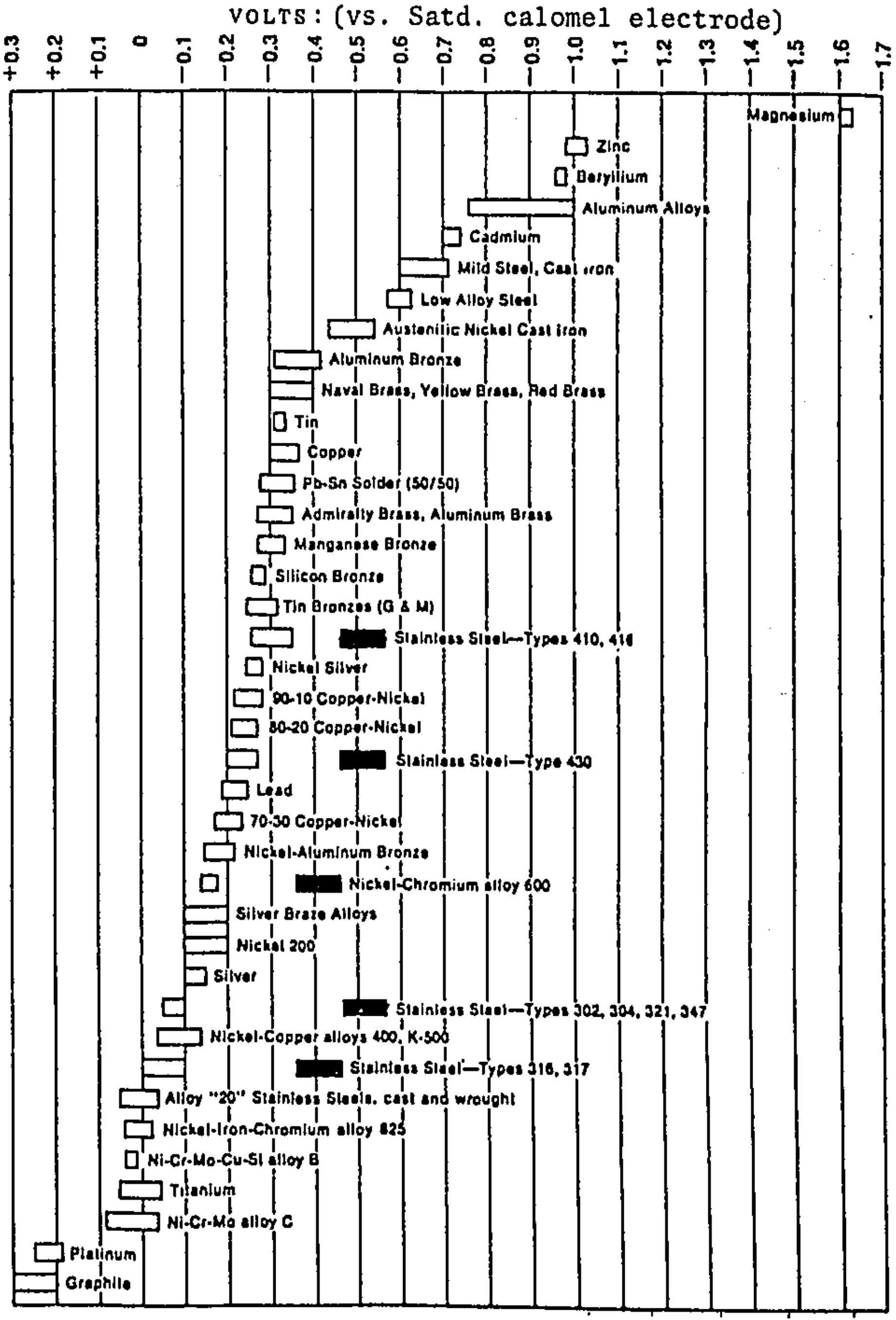


Fig. 5.4-11 Galvanic series in flowing seawater (INCO)

Flow rate 2.5 - 4m/s, Temperature 10°C - 27°C

■ : Certain alloys indicated by this symbol in low-velocity or poorly aerated water, and in shielded areas, may become active and exhibit a potential near -0.5 volt.

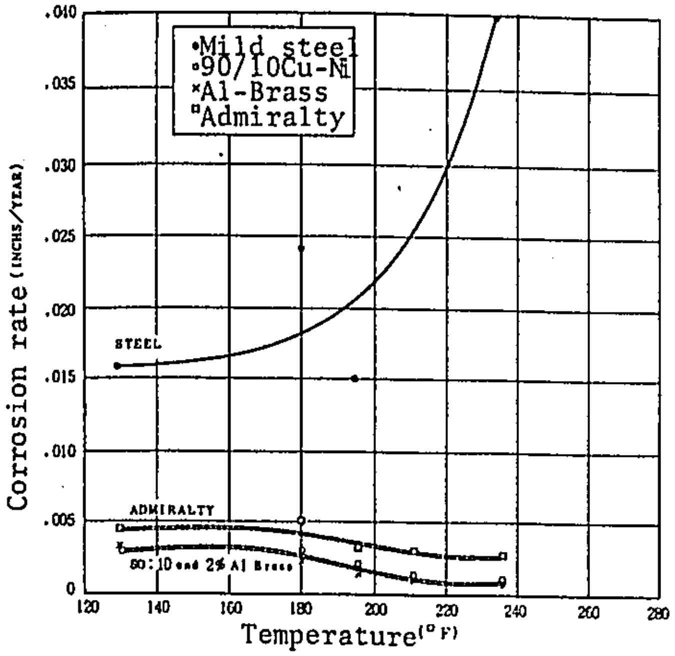
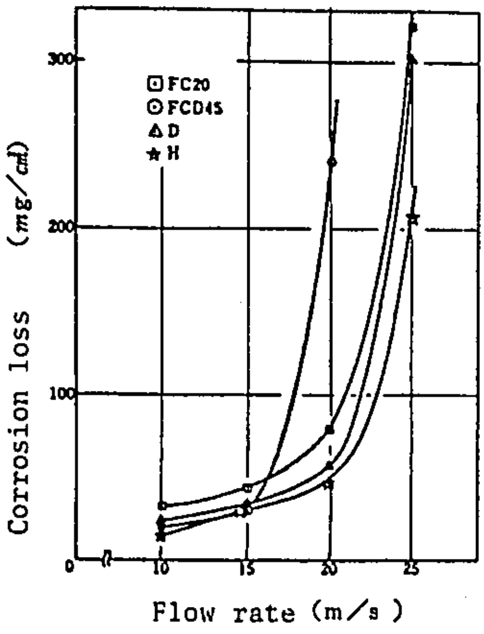


Fig. 5.4-12 Corrosion rate of metal in hot seawater



△D Elaky graphite type low Cr cast iron

☆H Spheroidal graphite type low Cr cast iron

Fig. 5.4-13 Corrosion loss of cast iron after 15 days

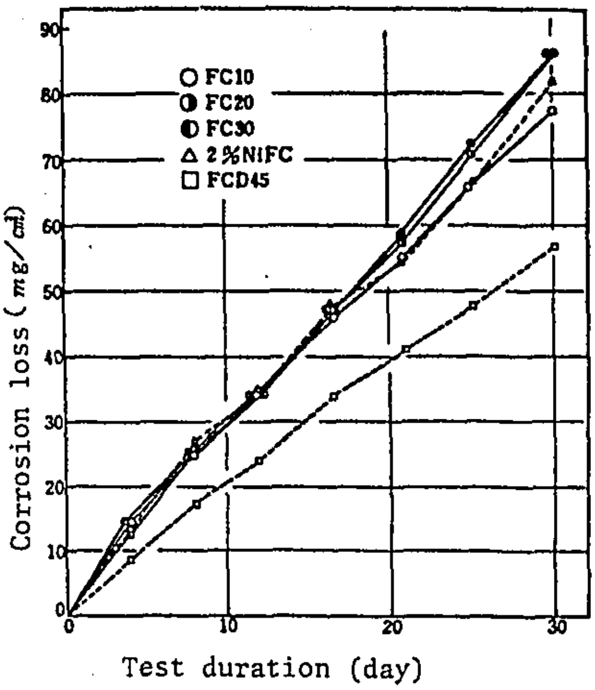


Fig. 5.4-14 Corrosion loss of cast iron in seawater at flow rate of 15 m/s

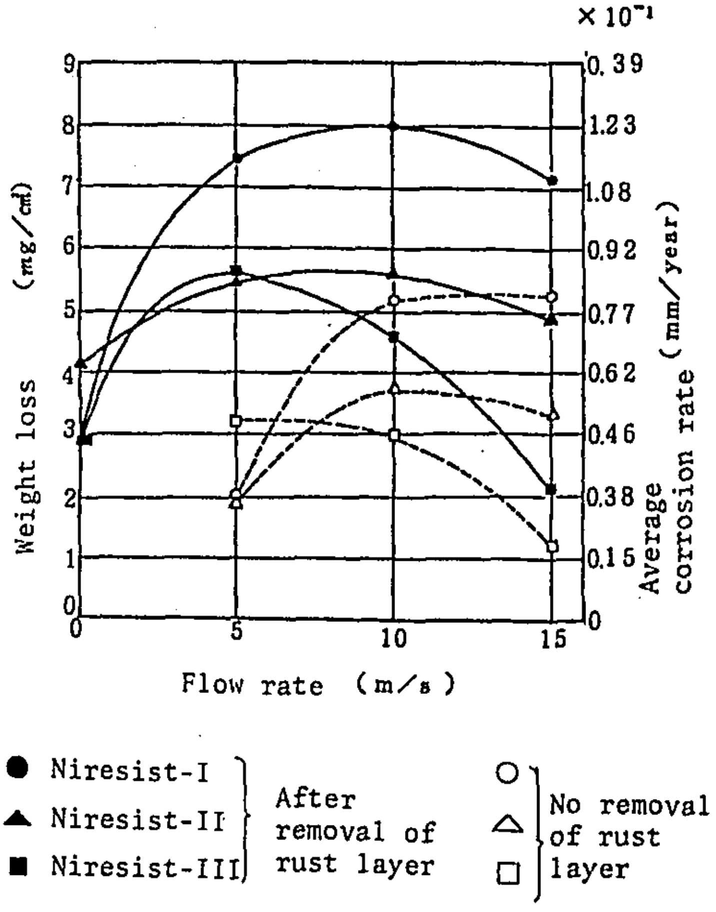


Fig. 5.4-15 Weight loss of niresist cast iron in seawater (30 days test)

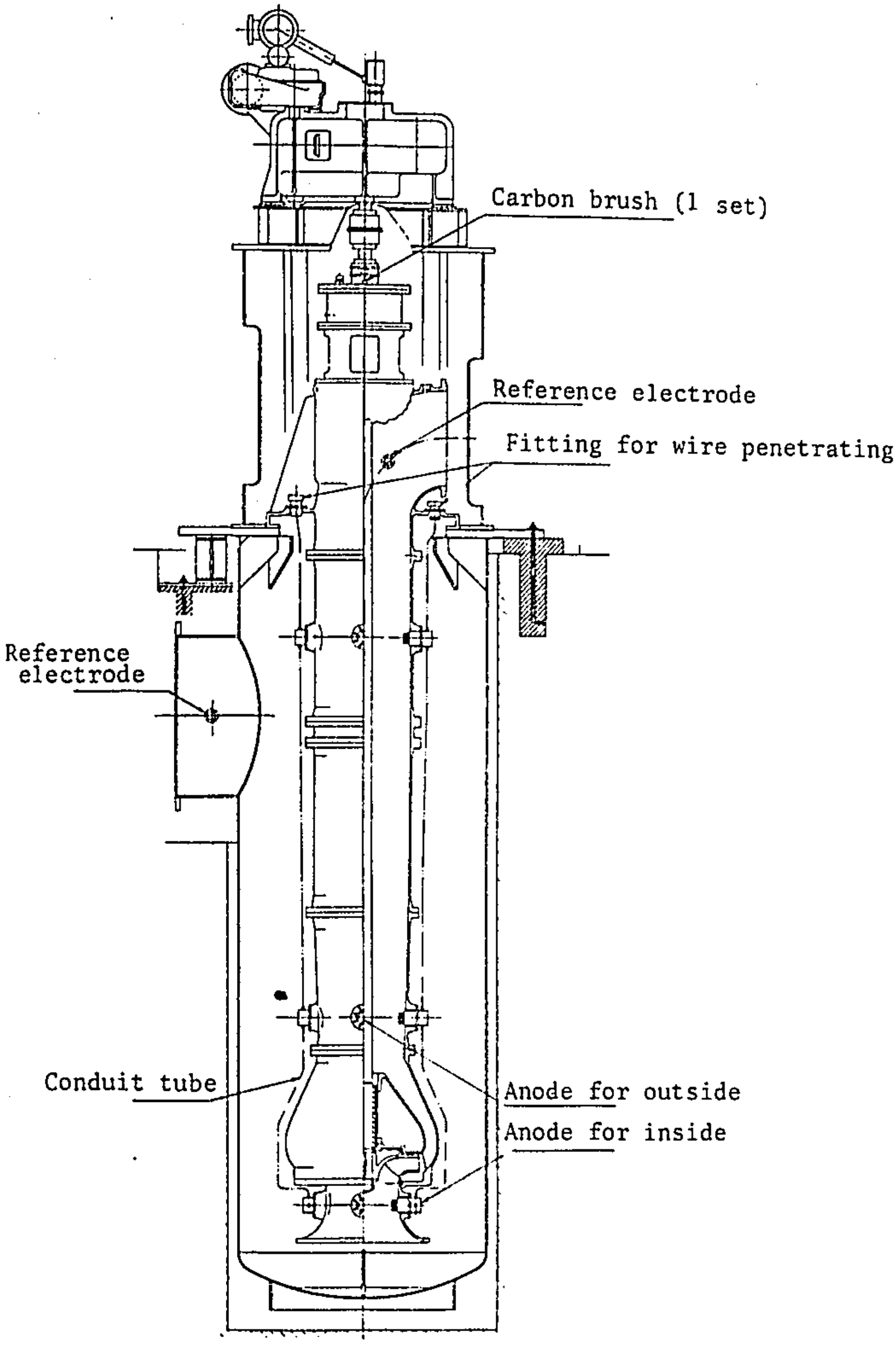


Fig. 5.4-16 Brine circulating pump and arrangement of cathodic protection (Pit-barrel type, Vertical mixed flow pump)

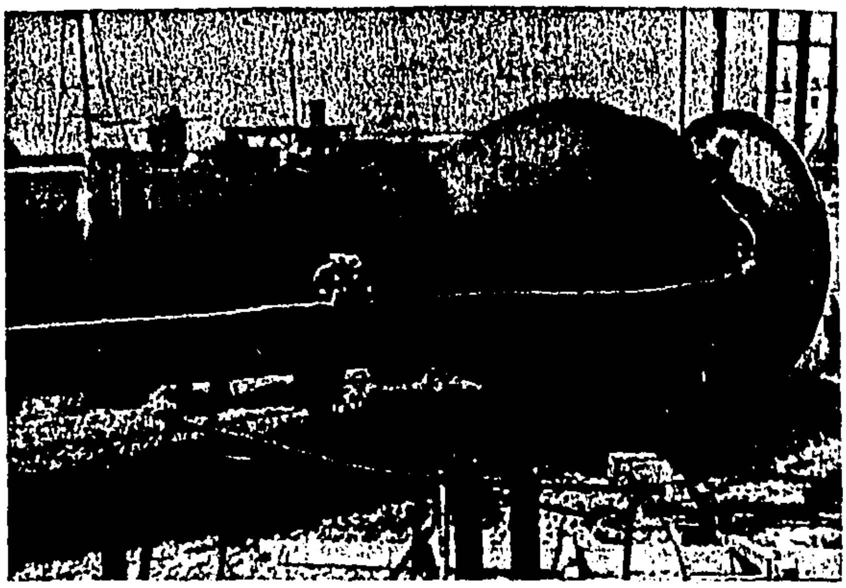


Fig. 5.4-17 Arrangement of cathodic protection for pump (External power supply system)

(a) Before use

(b) After use

Fig. 5.4-18 High silicon cast iron electrode

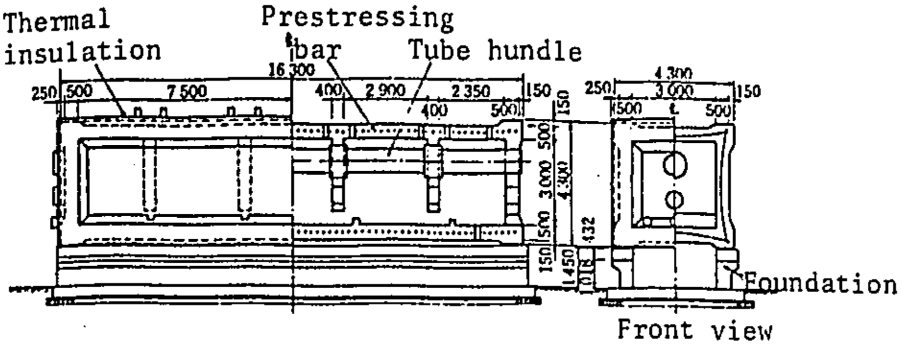


Fig. 5.5-1 Profile of prestressed concrete evaporator shell

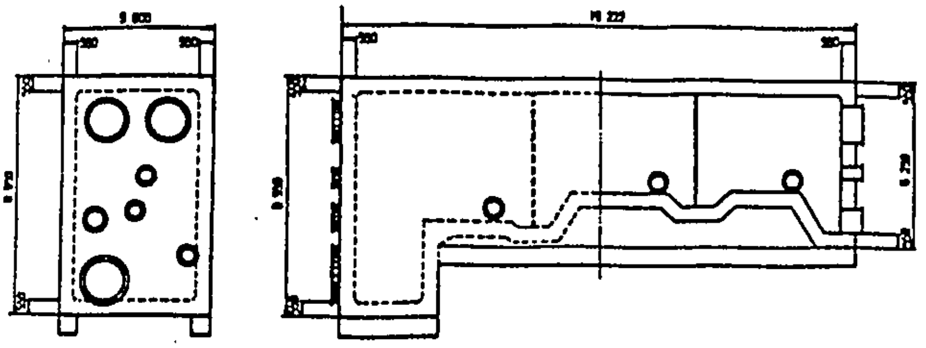


Fig. 5.5-2 Profile of reinforced concrete evaporator shell (heat rejection section)

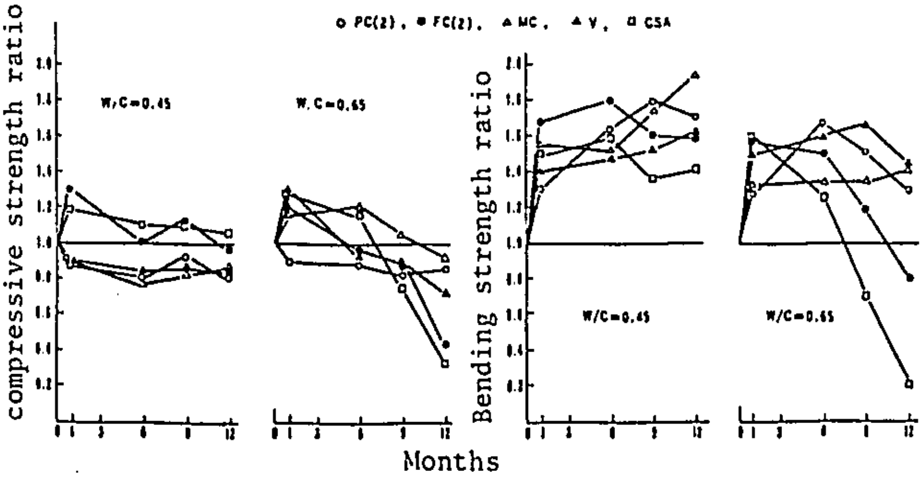


Fig. 5.5-3 Strength-ratios of the mortar specimens exposed to sea water at 120°C to those of controls

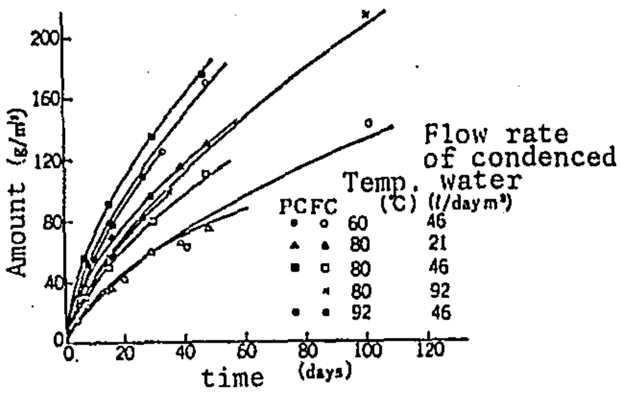


Fig. 5.5-4 The amount of dissolved matter vs leaching period

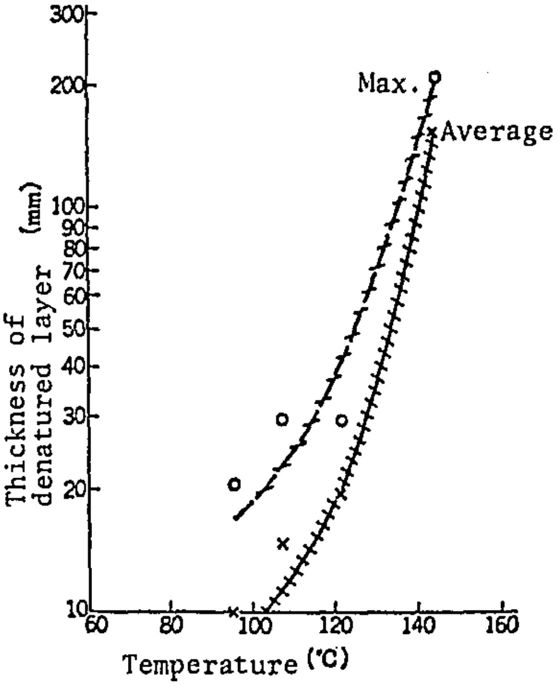
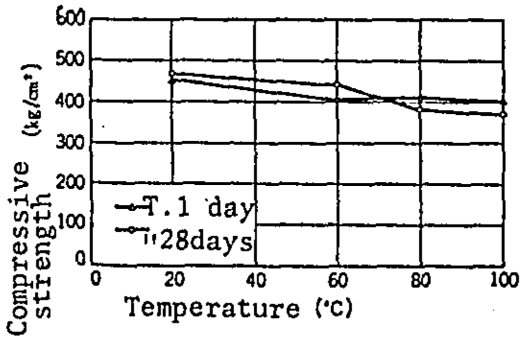


Fig. 5.5-5 Penetration at 30 years compared with temperature



Note:
T.1 day: Time since heating 1 day
" 28days: Time since heating 28 days

Fig. 5.5-6 Relation between compressive strength and temperature

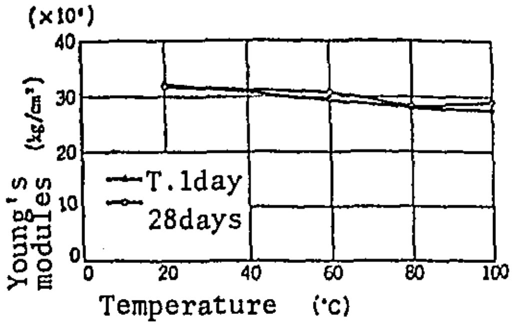


Fig. 5.5-7 Relation between young's modules and temperature

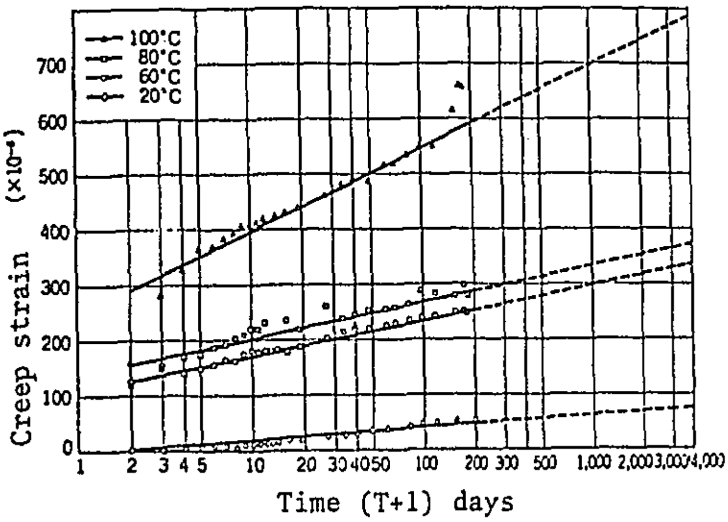


Fig. 5.5-8 Change of creep strain vs time

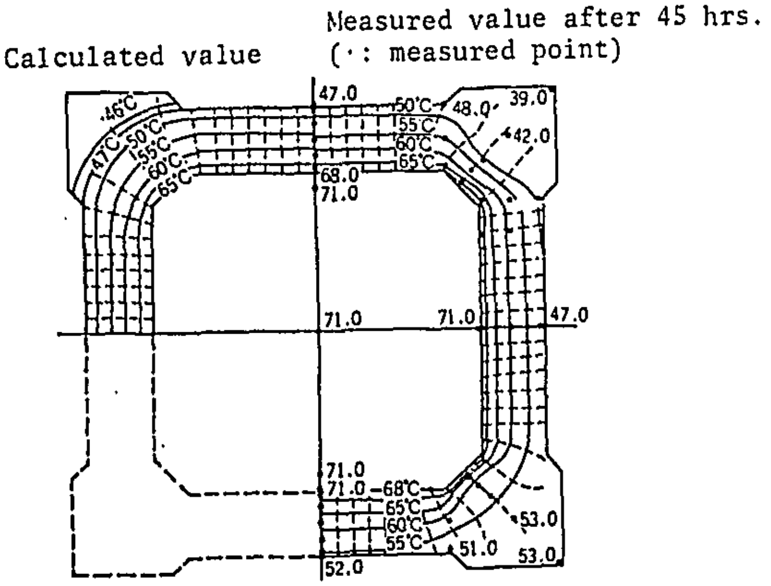


Fig. 5.5-9 Temperature distribution at Stationary state

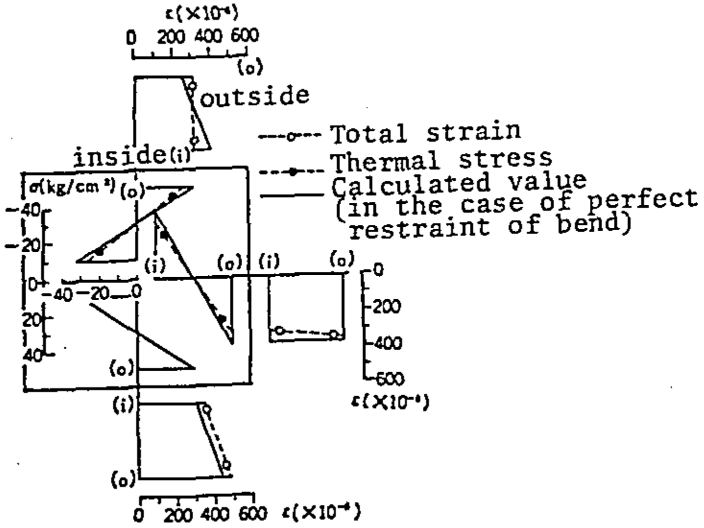


Fig. 5.5-10 Strain and stress in concrete wall (C-section at 45 hours)

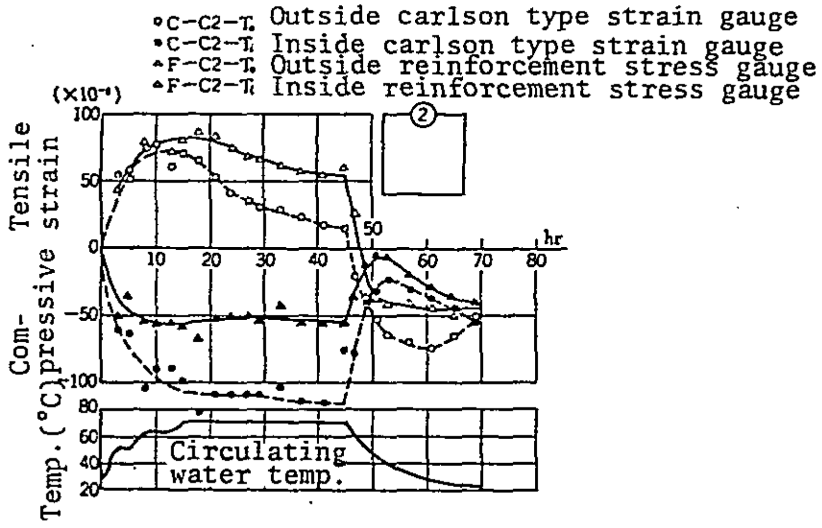


Fig. 5.5-11 Strain vs time

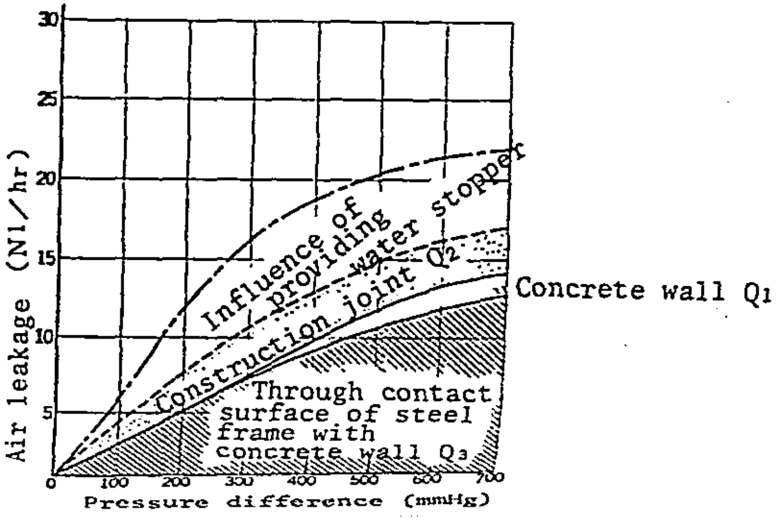
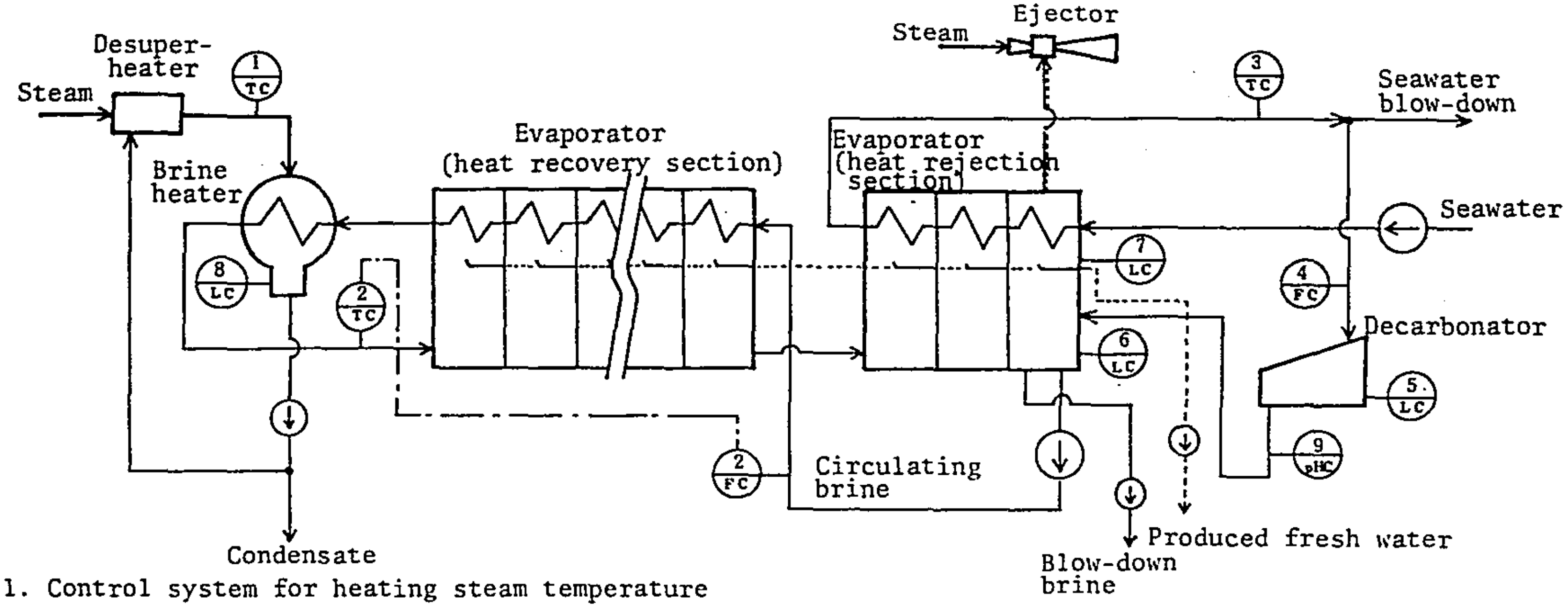


Fig. 5.5-12 Breaking down the total air leakage to allot to the each air-permeable member of concrete test model



1. Control system for heating steam temperature
2. Control system for temperature and flow of circulating brine
3. Control system for seawater temperature
4. Control system for make-up seawater flow
5. Control system for make-up seawater level in decarbonator
6. Control system for brine level in final stage
7. Control system for distillate water level in final stage
8. Control system for condensate level in brine heater
9. PH control system

Fig. 5.6-1 Process control system of MSF

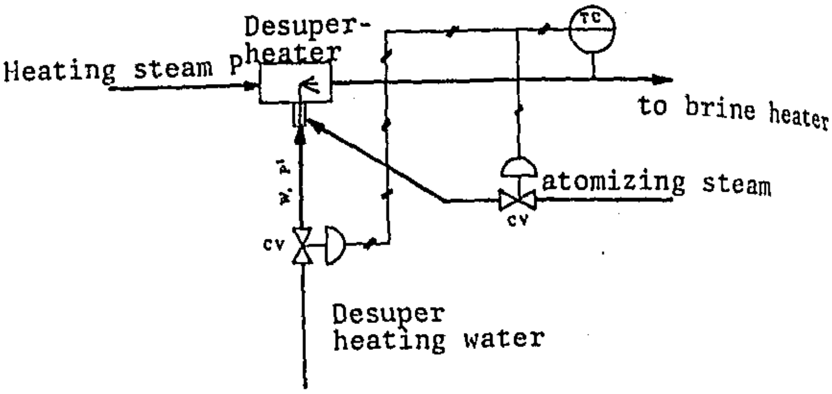


Fig. 5.6-2 Steam atomizing type desuperheater

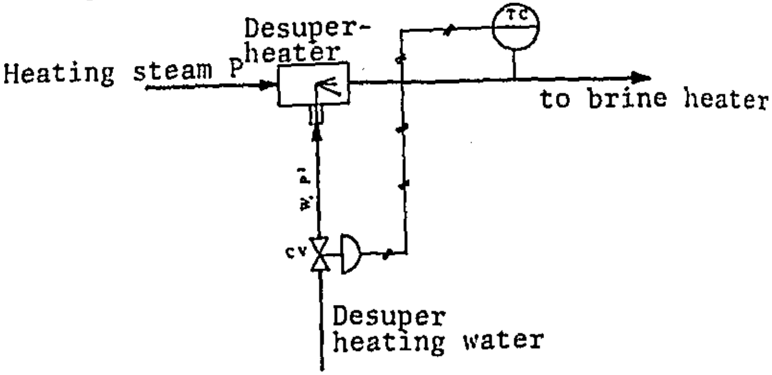


Fig. 5.6-3 Spray type desuperheater

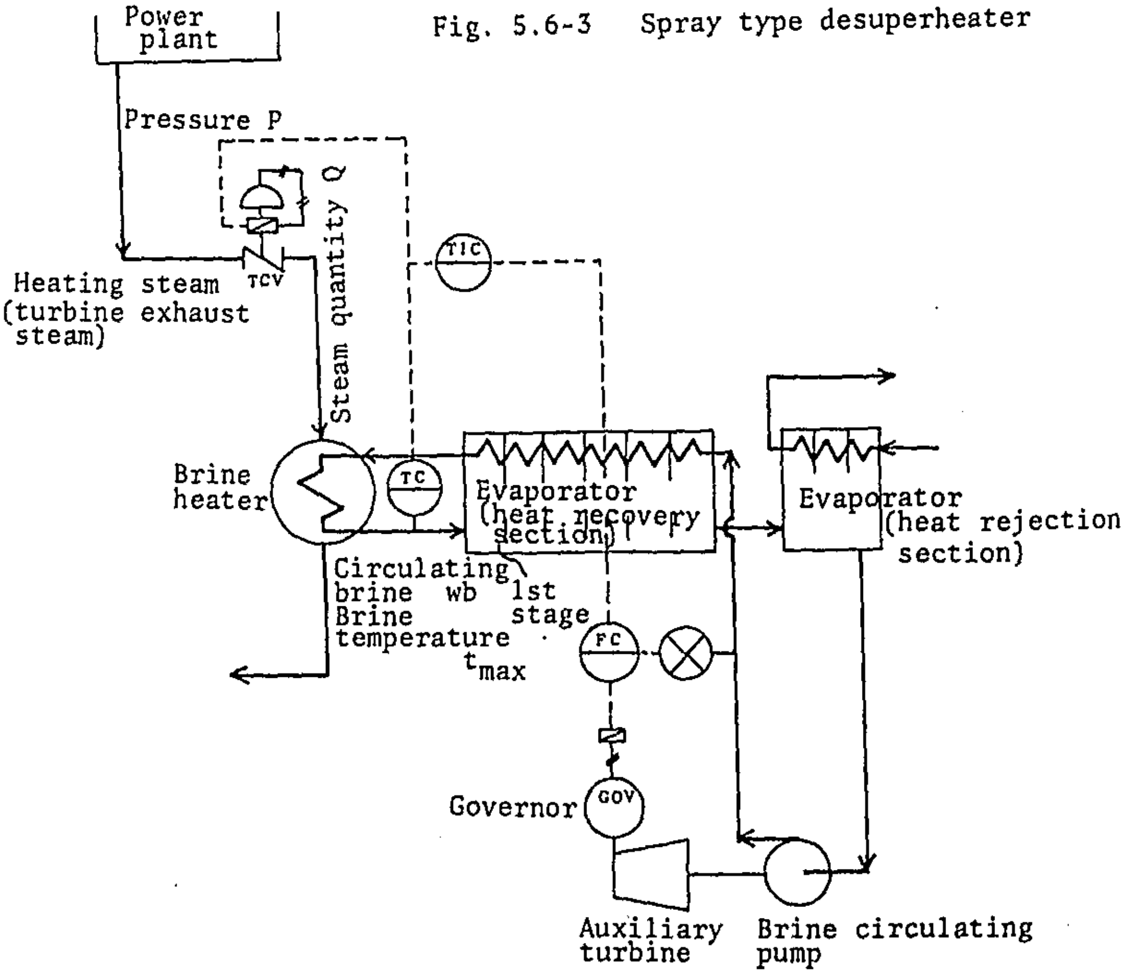


Fig. 5.6-4 Control system for temperature and
flow of circulating brine

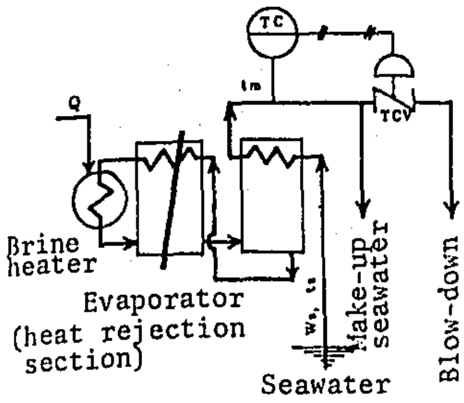


Fig. 5.6-5 Control system for outlet seawater temperature

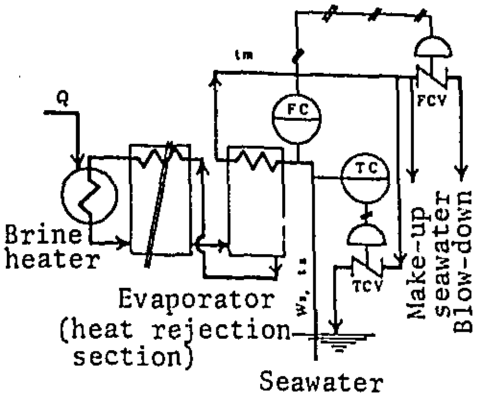


Fig. 5.6-6 Control system for inlet seawater temperature

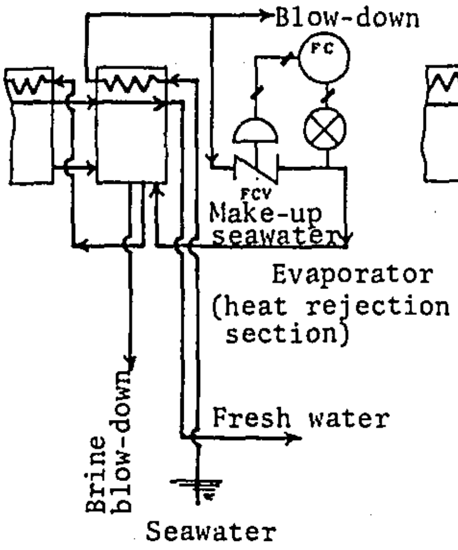


Fig. 5.6-7 Independent control system for make-up seawater flow

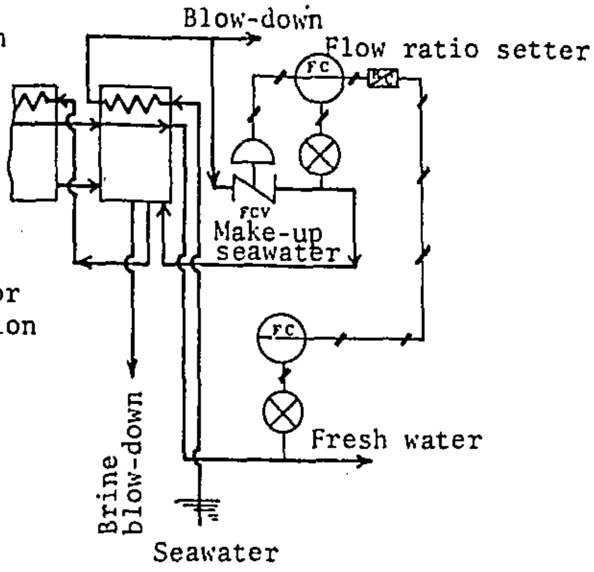


Fig. 5.6-8 Control system for make-up seawater flow proportioning fresh water

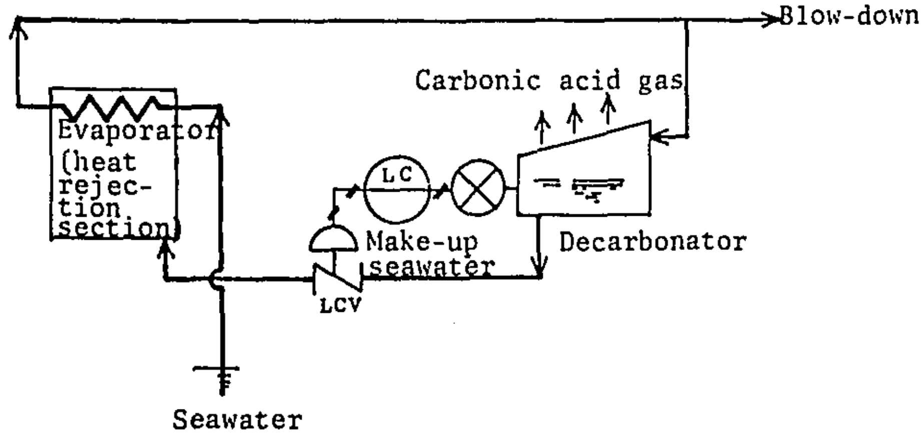


Fig. 5.6-9 Control system for make-up seawater level in decarbonator

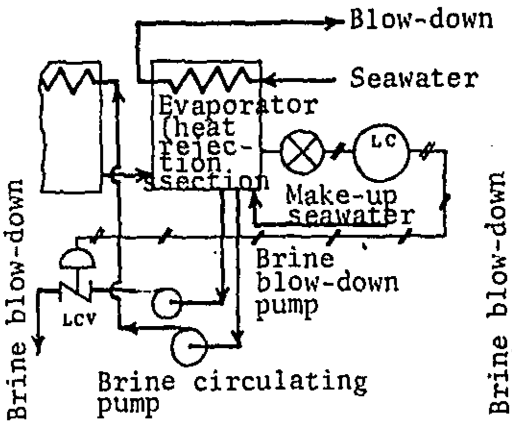


Fig. 5.6-10 Control system for brine level (blow-down from final stage)

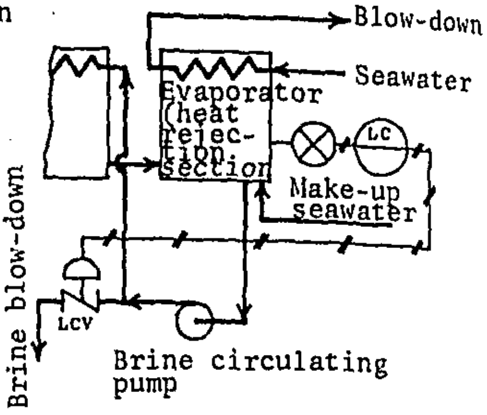


Fig. 5.6-11 Control system for brine level (blow-down from circulating brine line)

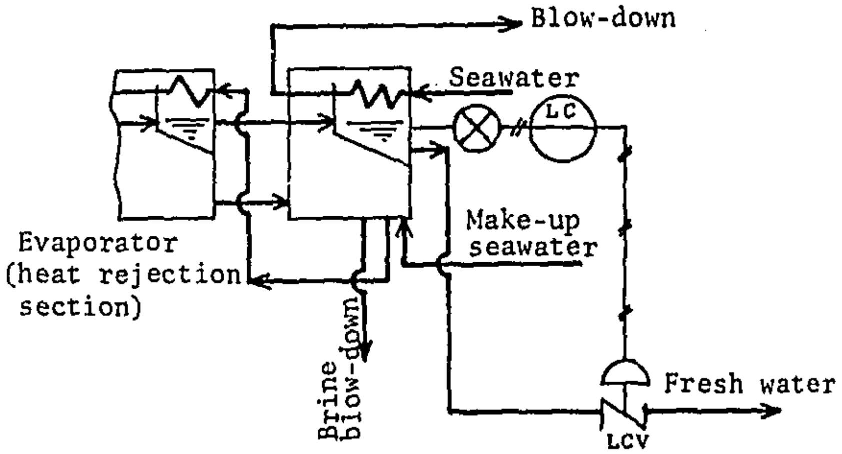


Fig. 5.6-12 Control system for distillate water level

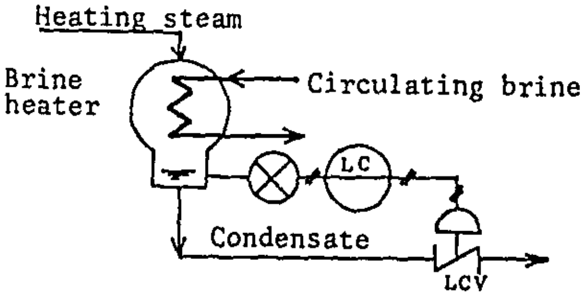


Fig. 5.6-13 Control system for condensate level

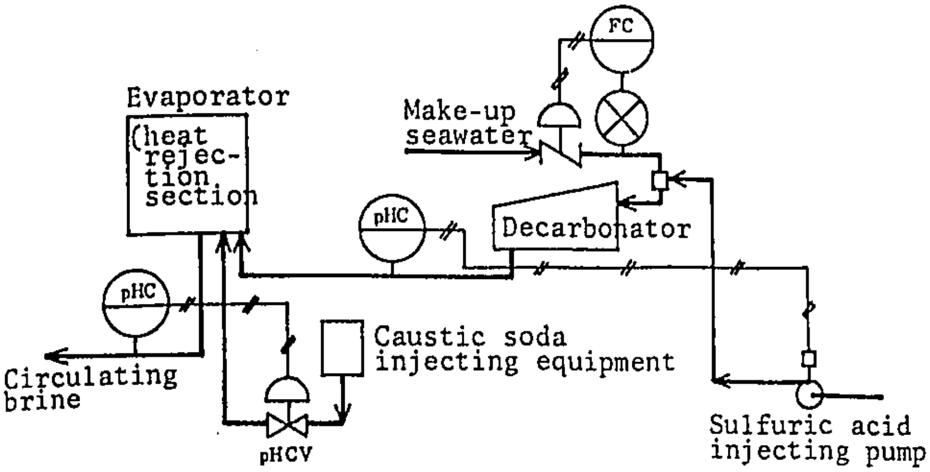


Fig. 5.6-14 pH control system (pH self restoration type)

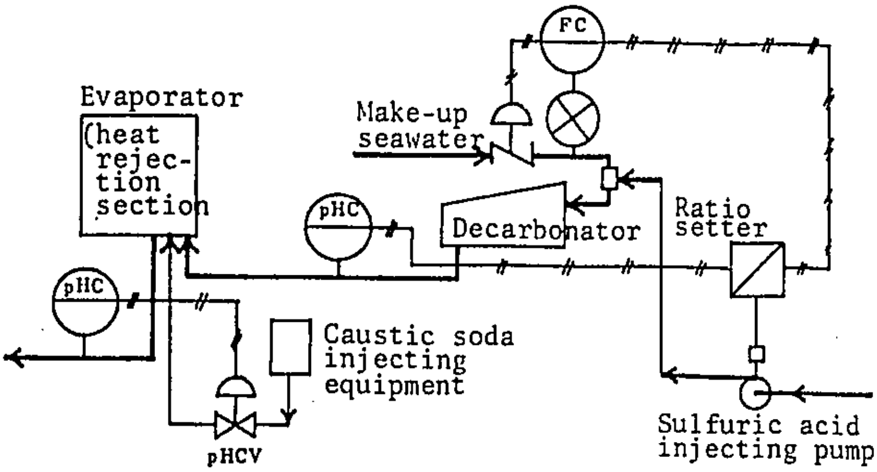


Fig. 5.6-15 pH control system (proportioning to make-up seawater· pH fine adjust type)

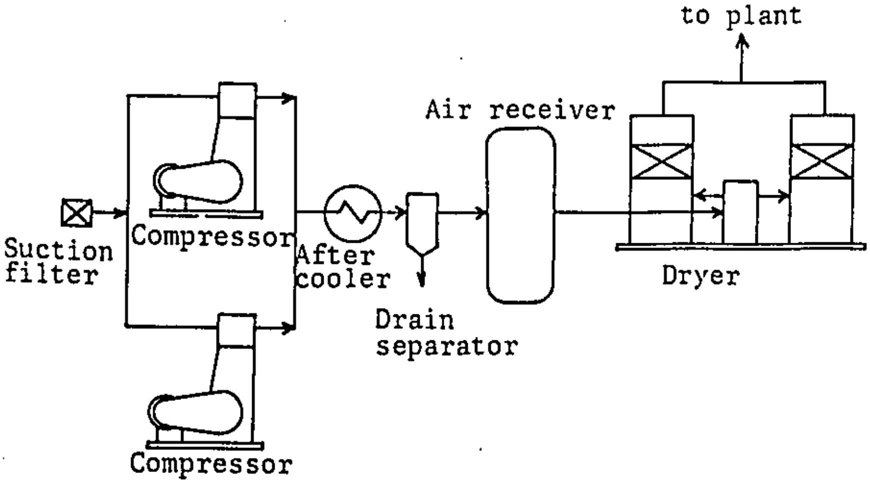


Fig. 5.6-16 Pneumatic source system

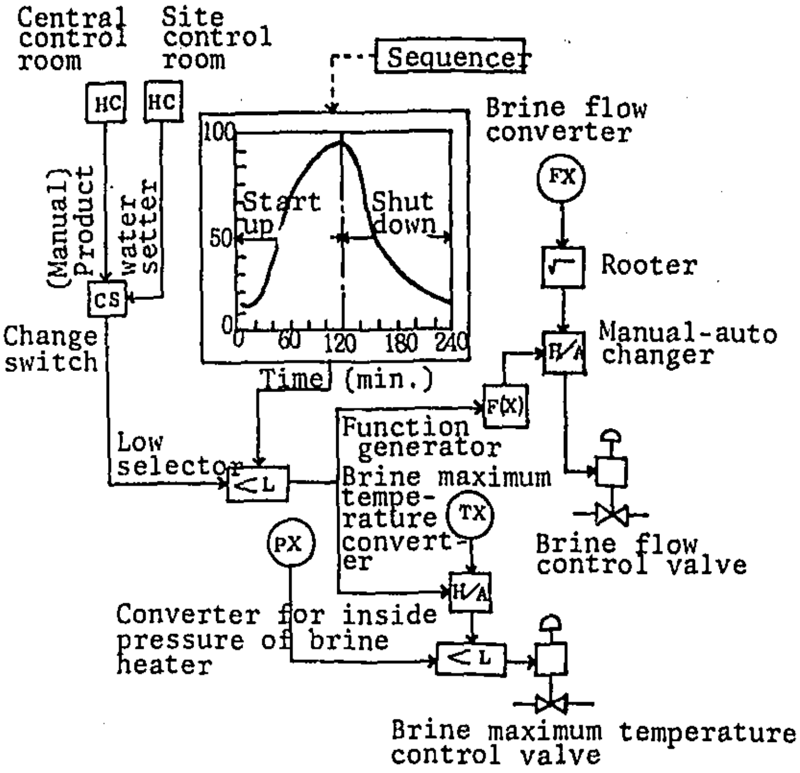


Fig. 5.6-17 Block diagram of start up and shut down control

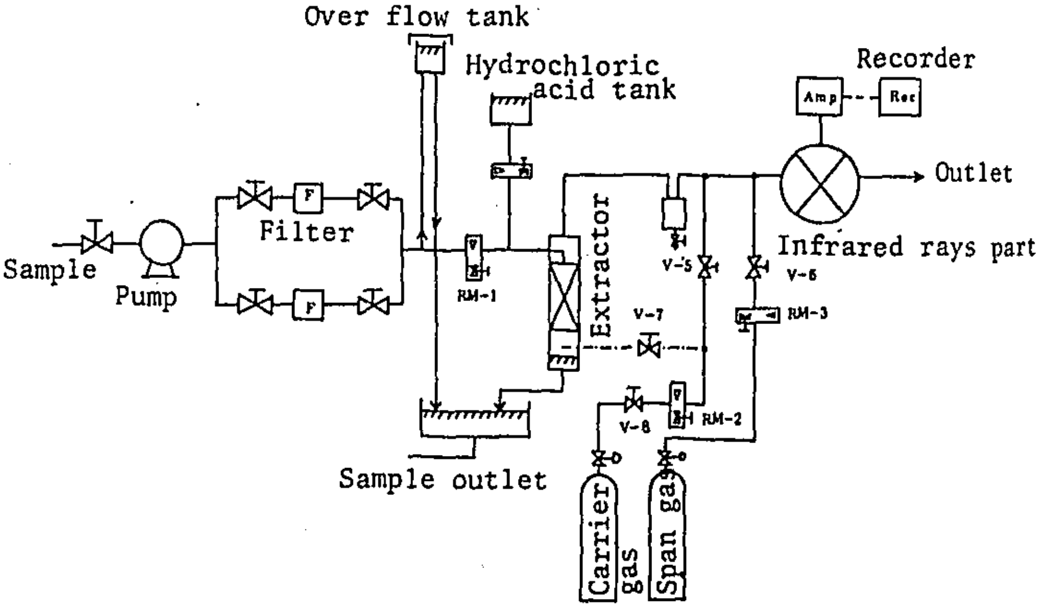


Fig. 5.6-18 Flow diagram of total carbonic acid analyzer

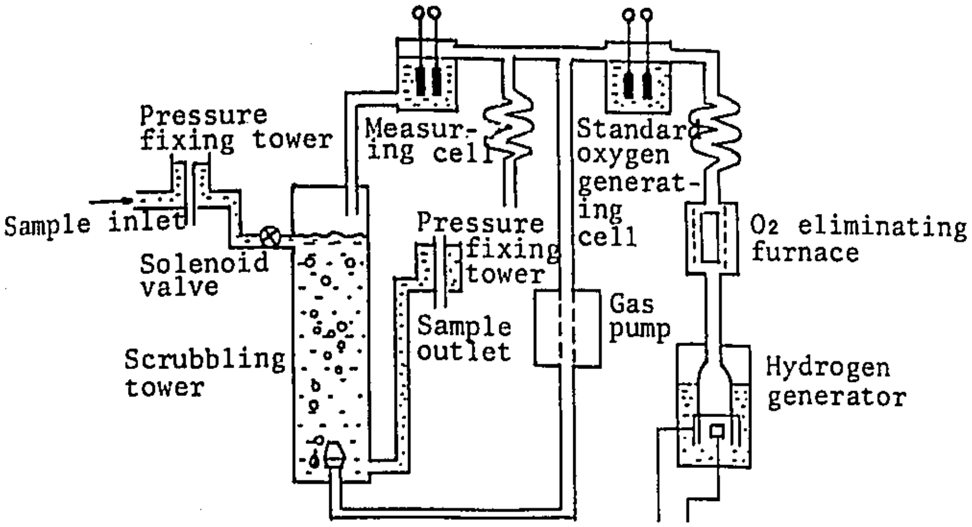


Fig. 5.6-19 Outline of EIL - Cambridge desolved oxygen analyzer

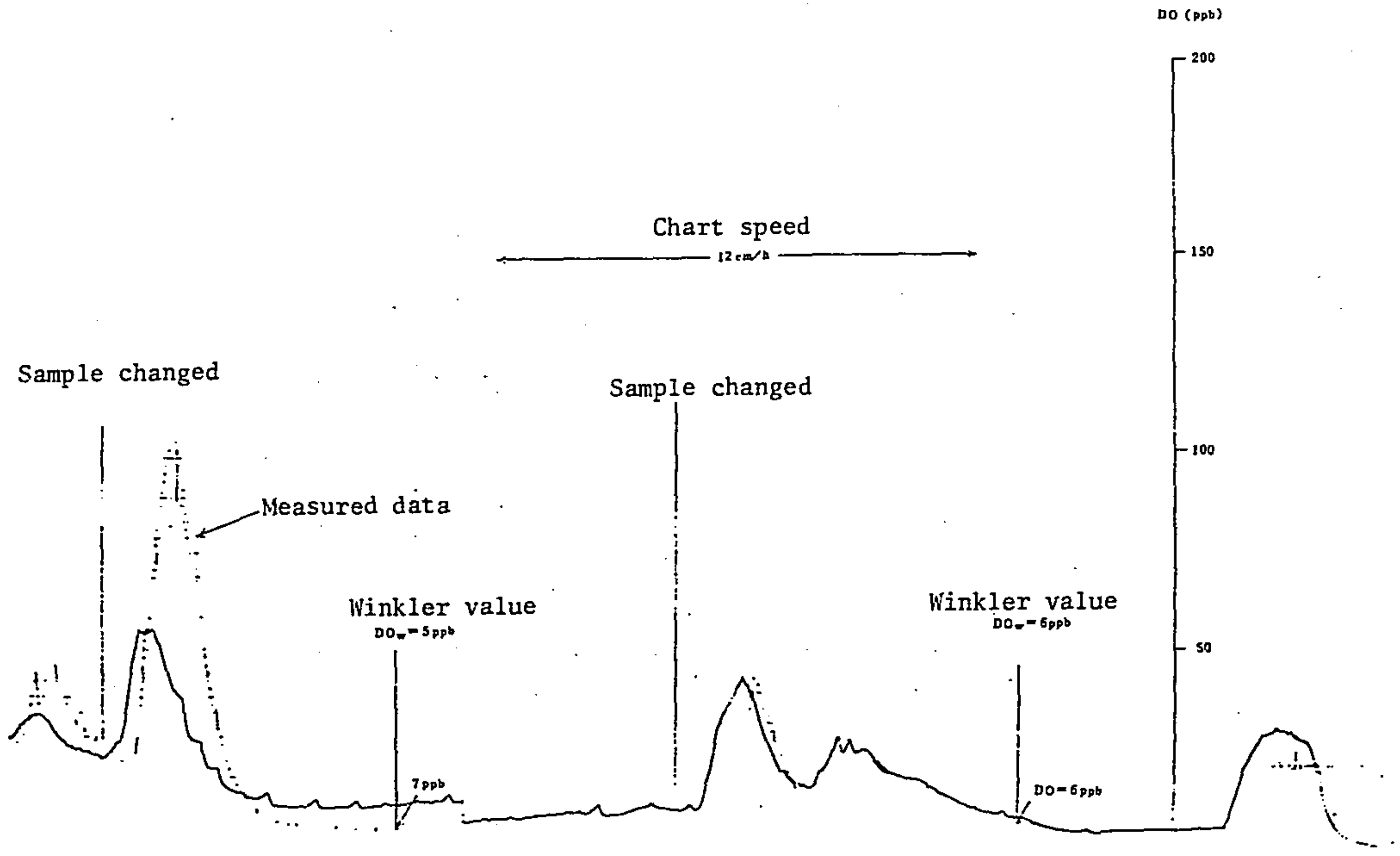


Fig. 5.6-20 Example of EIL - Cambridge disolved oxygen analyzer in operation

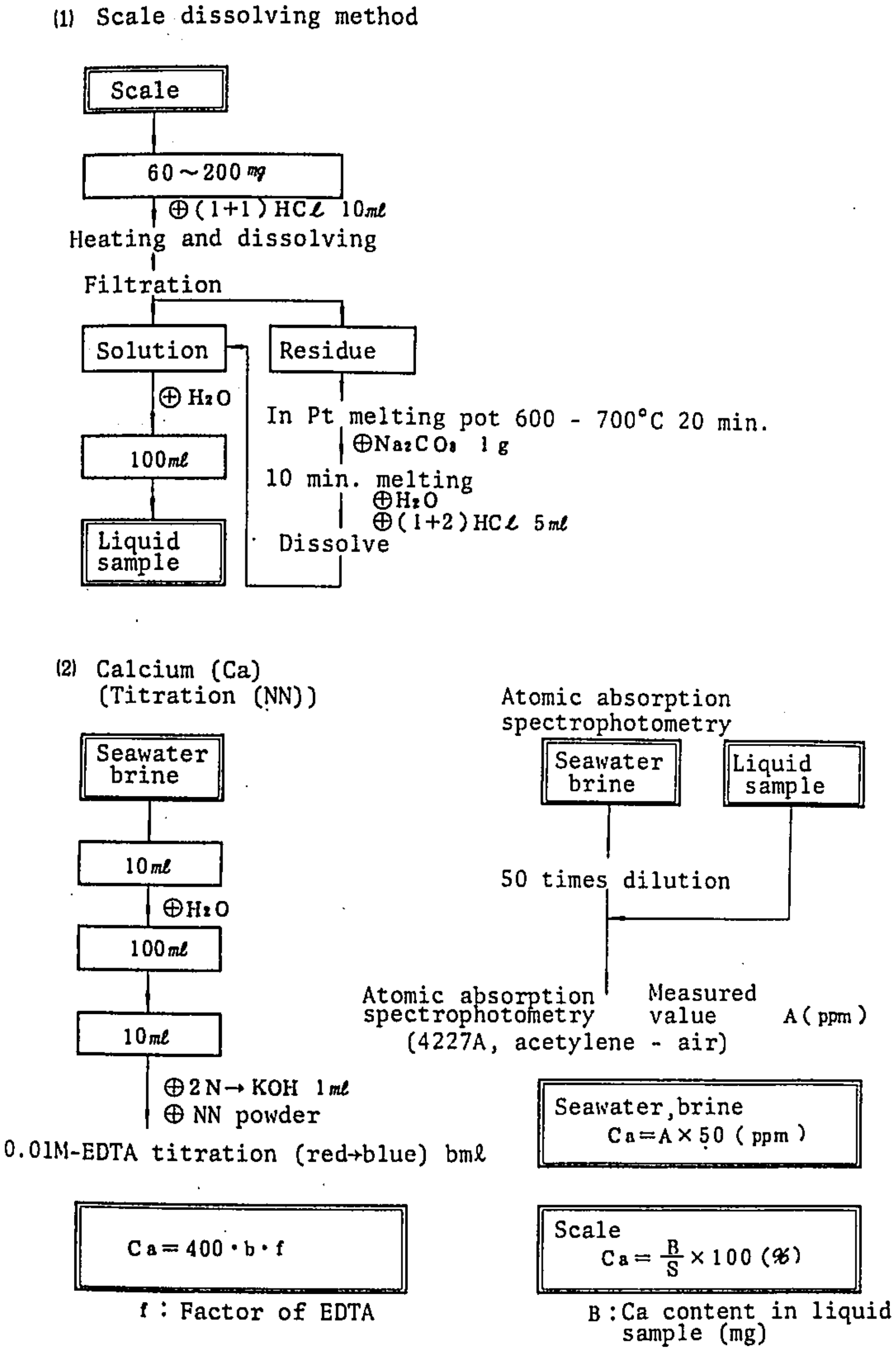
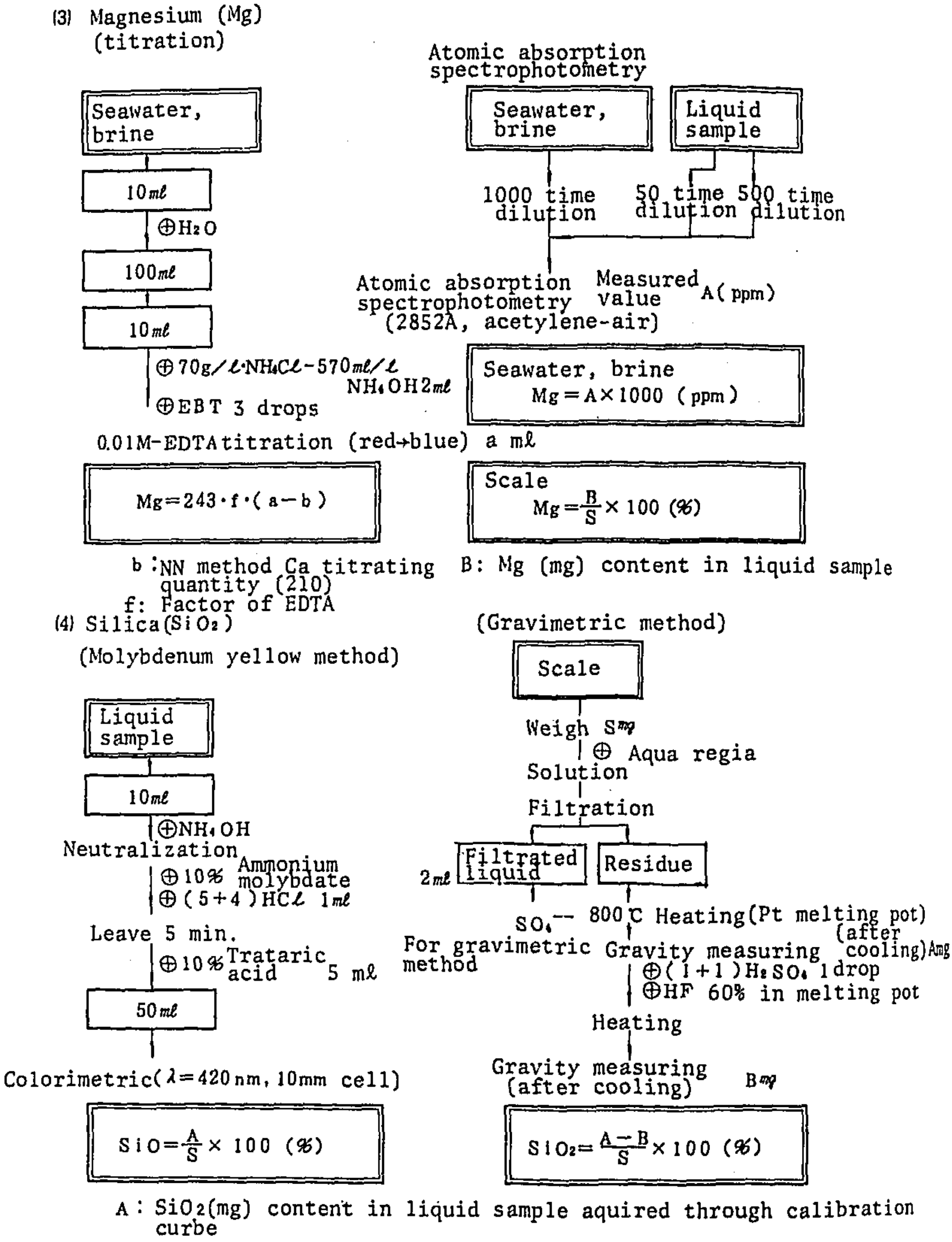


Fig. 5.6-21 Analyzing method

(3) Magnesium (Mg)

(4) Silica (SiO_2)



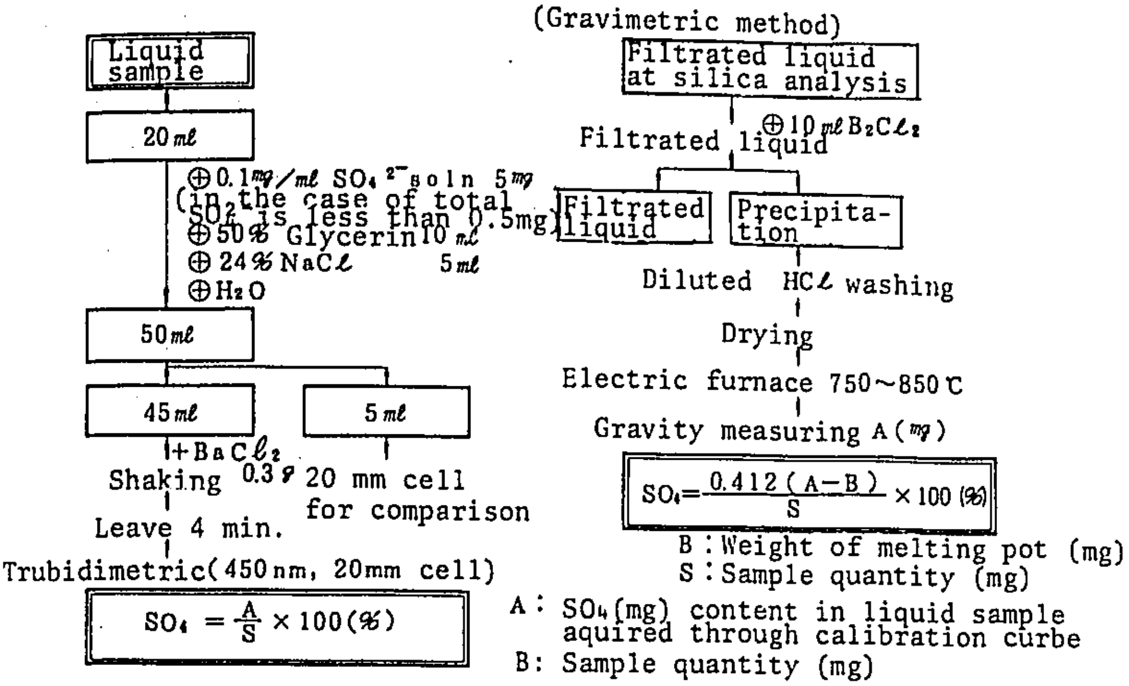
b: NN method Ca titrating quantity (210)
f: Factor of EDTA

B: Mg (mg) content in liquid sample

A: SiO_2(mg) content in liquid sample aquired through calibration curbe

Fig. 5.6-22 Analyzing method

(5) Sulfuric acid radical (SO_4^{2-})
(turbidimetric method)



(6) Iron (Fe)
(Colorimetric method)

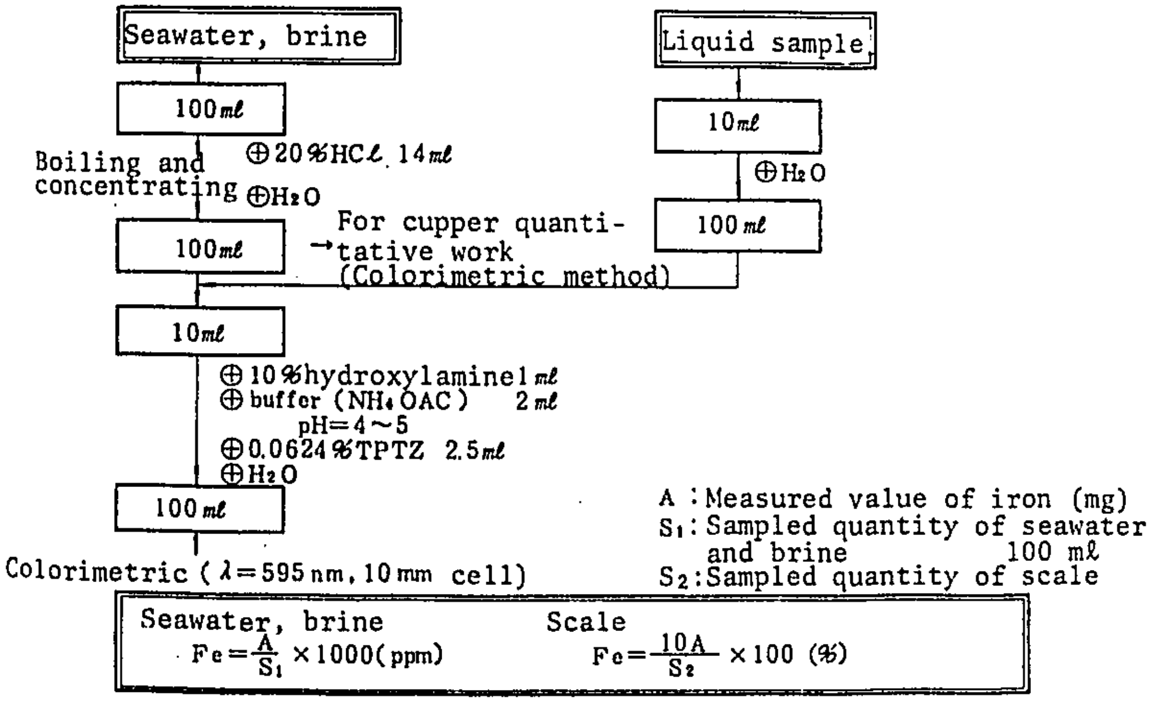
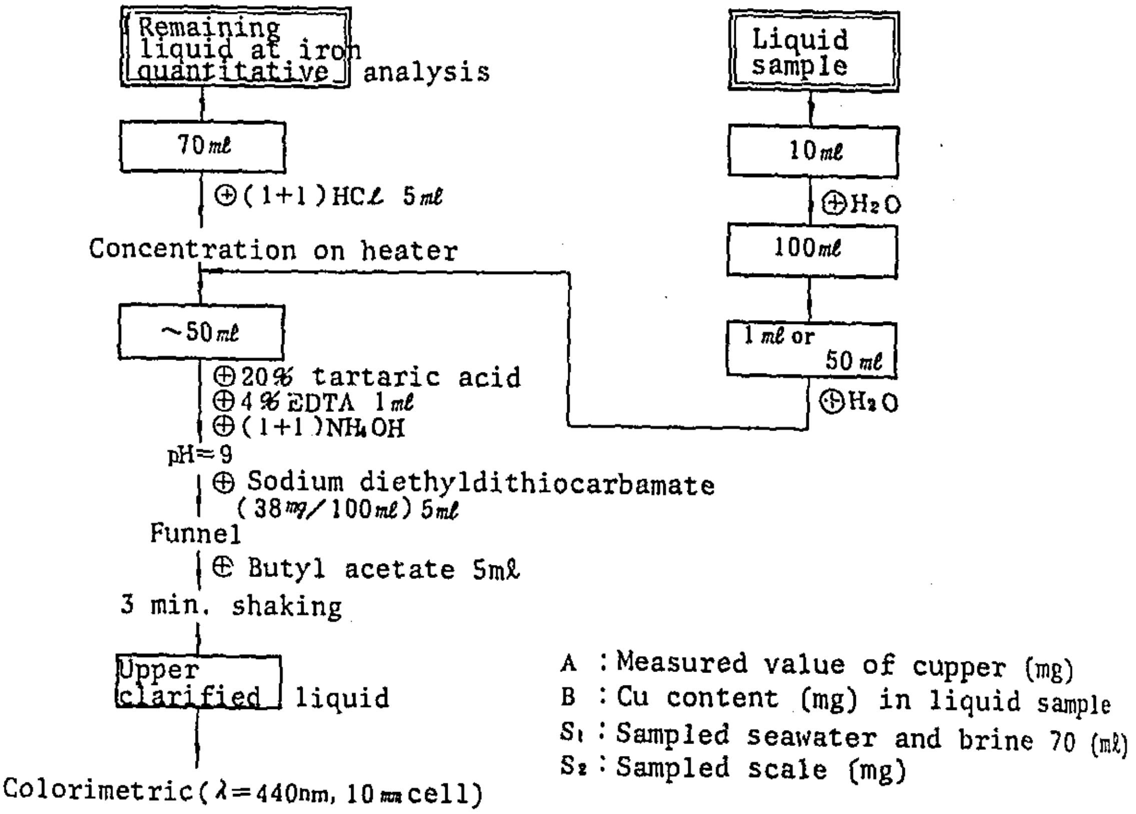


Fig. 5.6-23 Aanlyzing method

(7) Cupper (Cu)
(Colorimetric method)

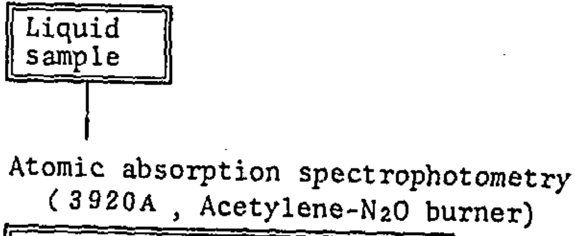


A : Measured value of cupper (mg)
B : Cu content (mg) in liquid sample
S_1 : Sampled seawater and brine 70 (mℓ)
S_2 : Sampled scale (mg)

Seawater, brine $Cu=\frac{A}{S_1}\times 1000$ (ppm)

Scale $Cu=\frac{B}{B_2}\times 100$ (%)

(8) Aluminium (Al)
(Atomic absorption spectrophotometry)



$A\ell=\frac{A}{S}\times 100$ (%)

Fig. 5.6-24 Analyzing method

(9) Titanium (Ti)
(Colorimetric method)

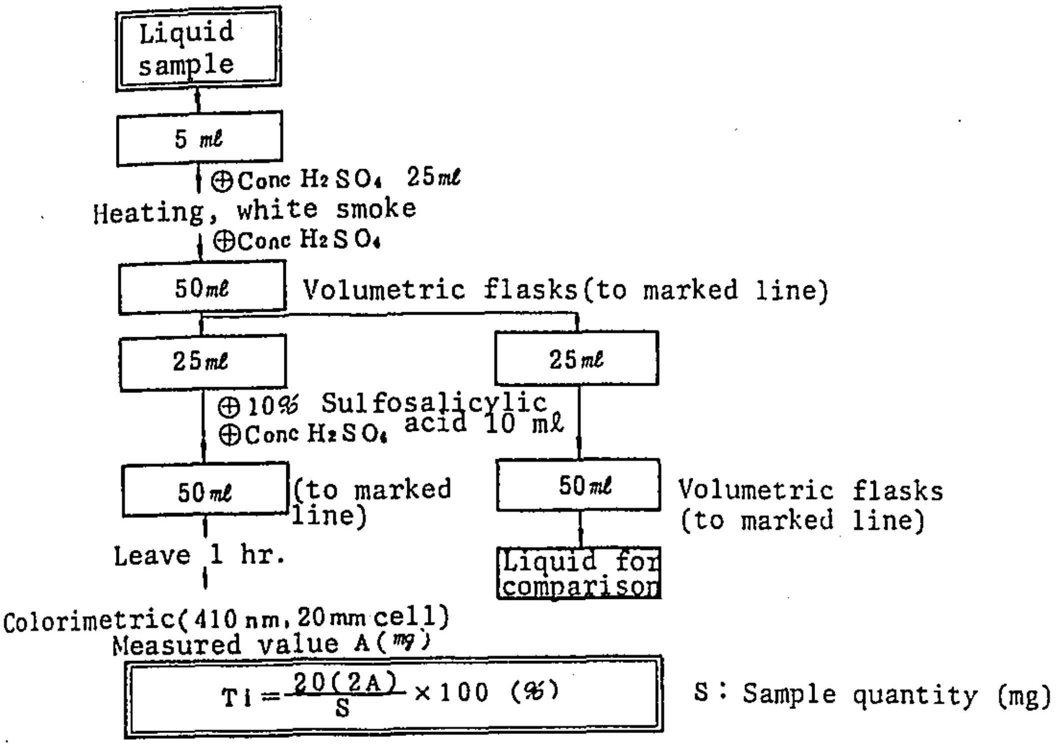
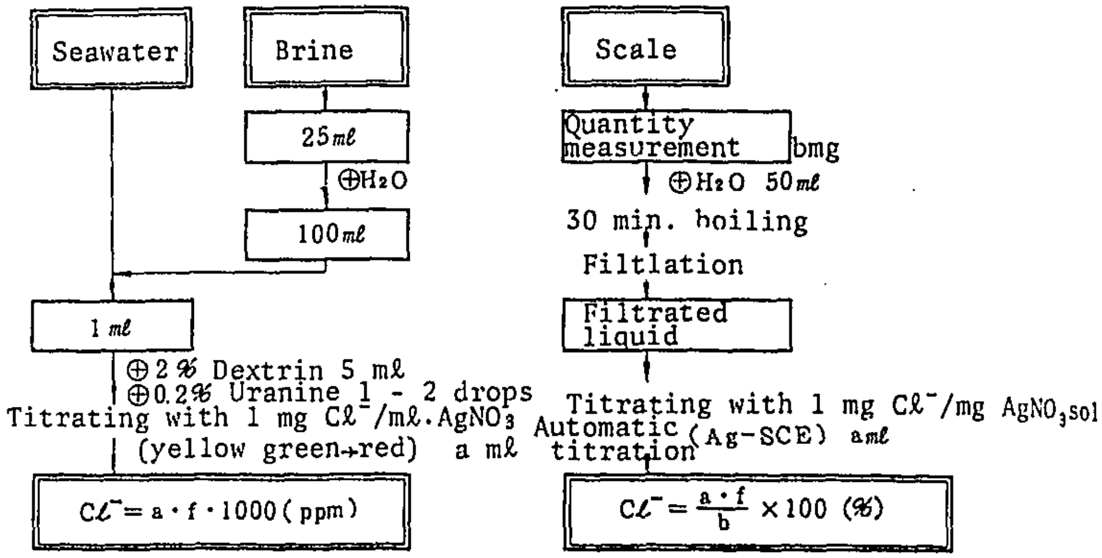


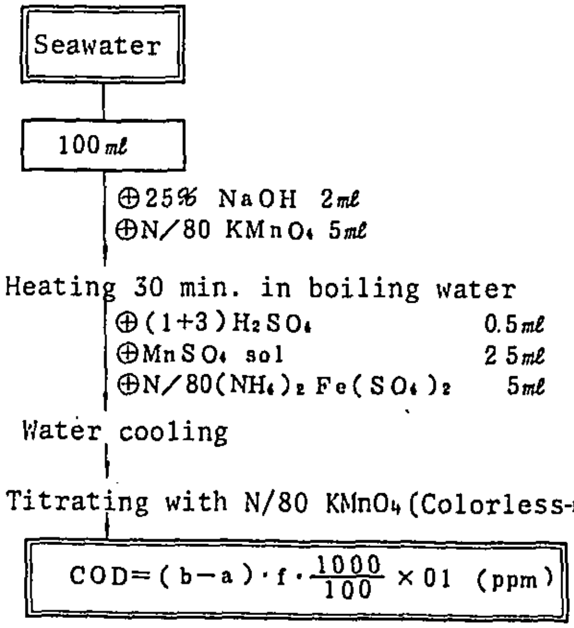
Fig. 5.6-25 Analyzing method

(10) Chlorine (Cl^-)



(f: Factor of $AgNO_3$ sol)

(11) COD (Chemical oxygen demand)



b : Titrating quantity of $KMnO_4$sol (mℓ)
a : Titrating quantity of $KMnO_4$sol in brank test (mℓ)
f: Factor of N/80 $KMnO_4$sol

Fig. 5.6-26 Analyzing method

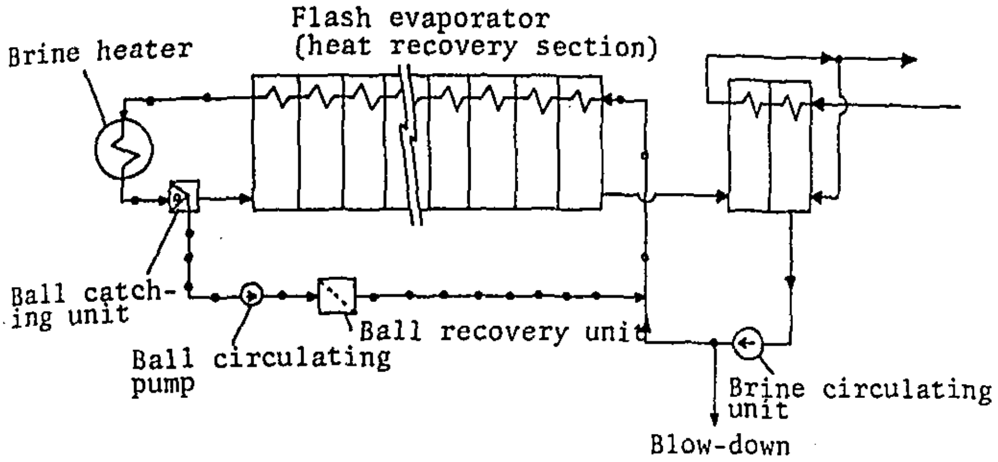


Fig. 5.6-27 Ball cleaning system

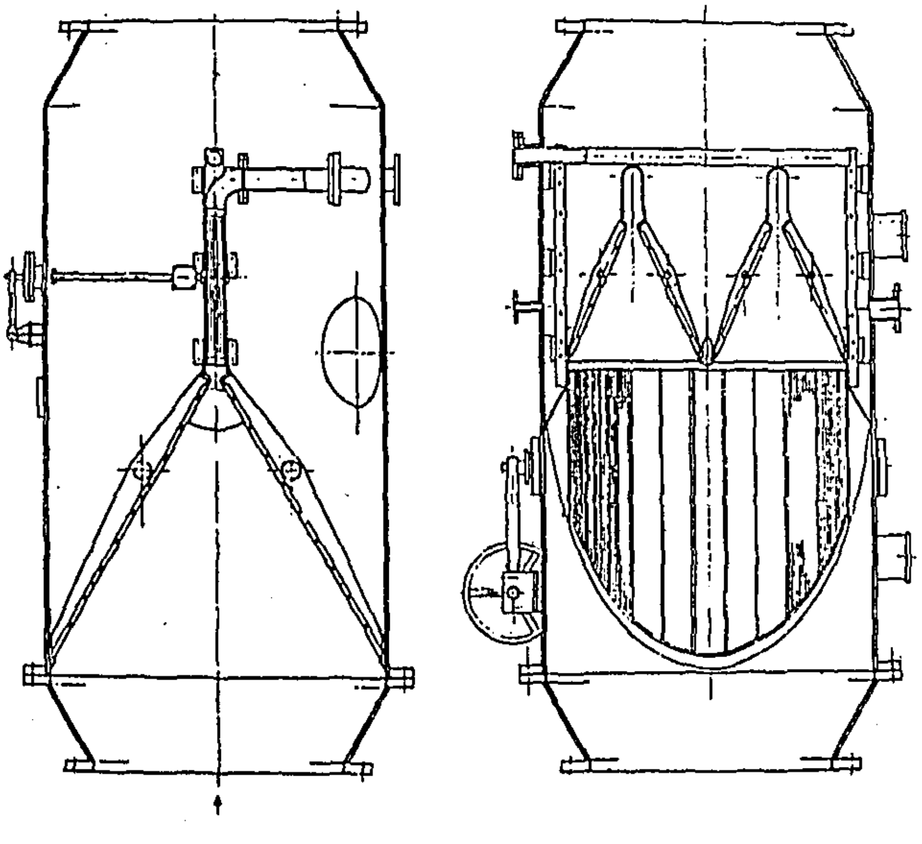


Fig. 5.6-28 Construction of ball catching unit

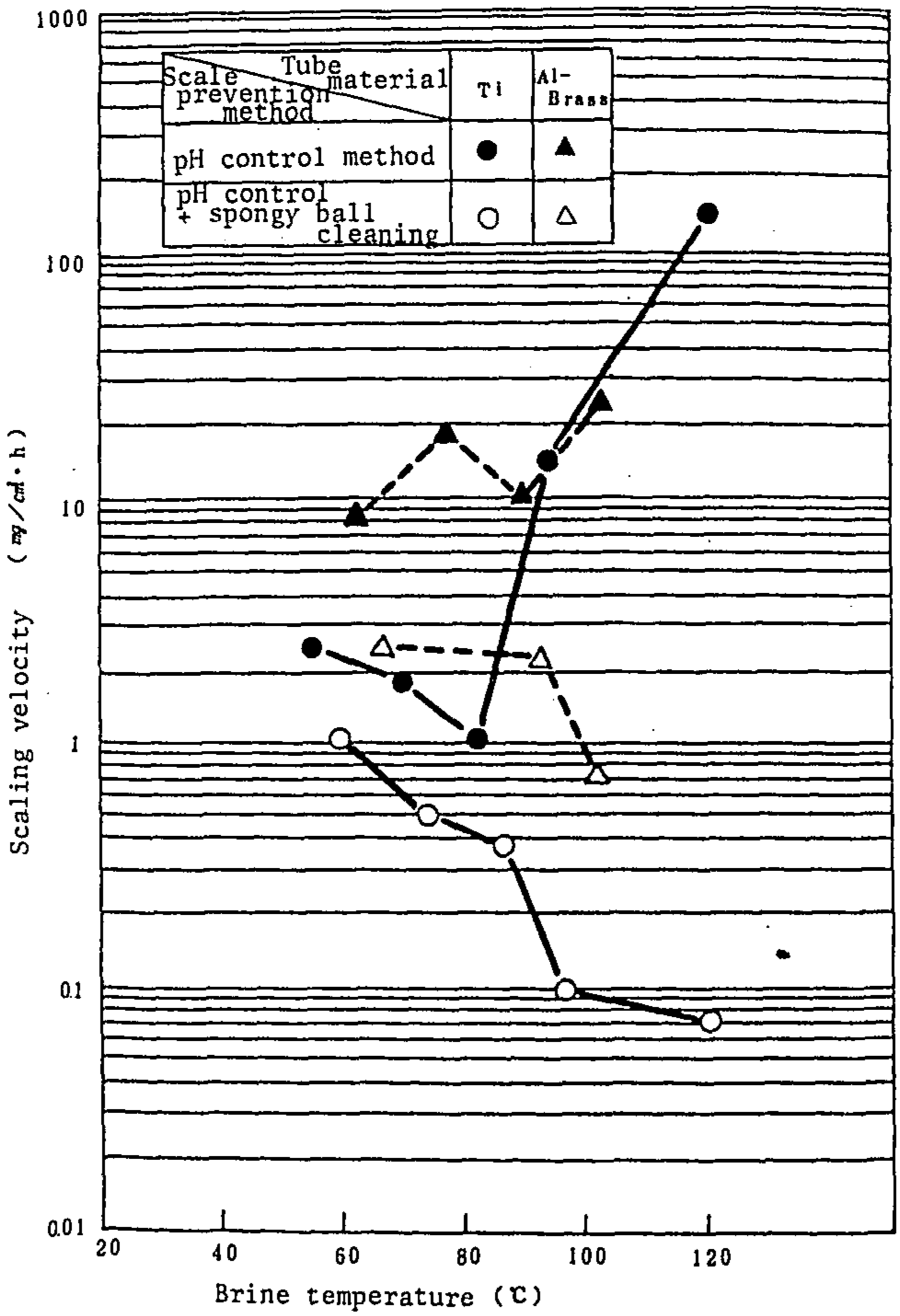


Fig. 5.6-30 Scaling velocity at long continuous operation (2,000 hr)

Fig. 5.6-29 Outside appearance of spongy ball

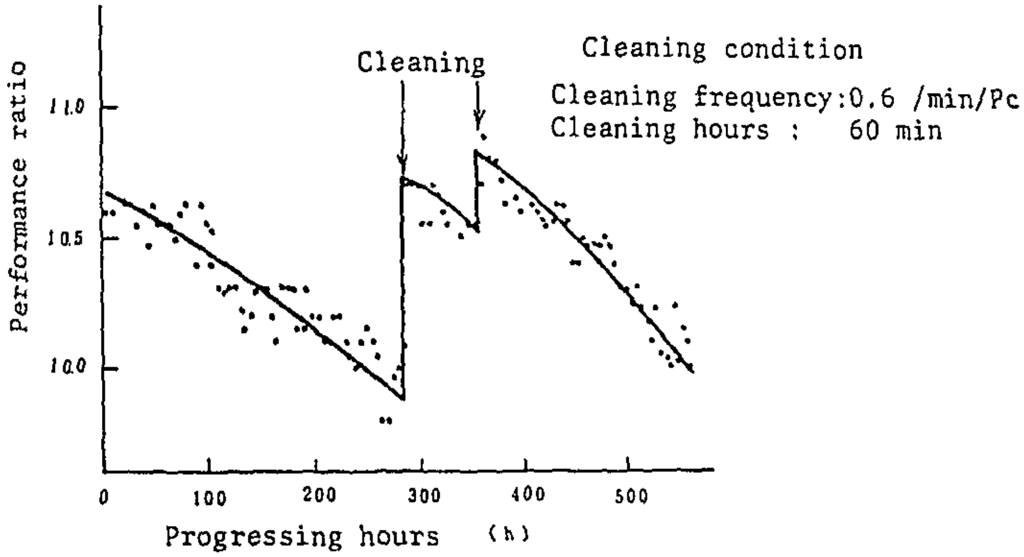


Fig. 5.6-31 Effect of spongy ball cleaning on 3,000 t/d plant at Chigasaki

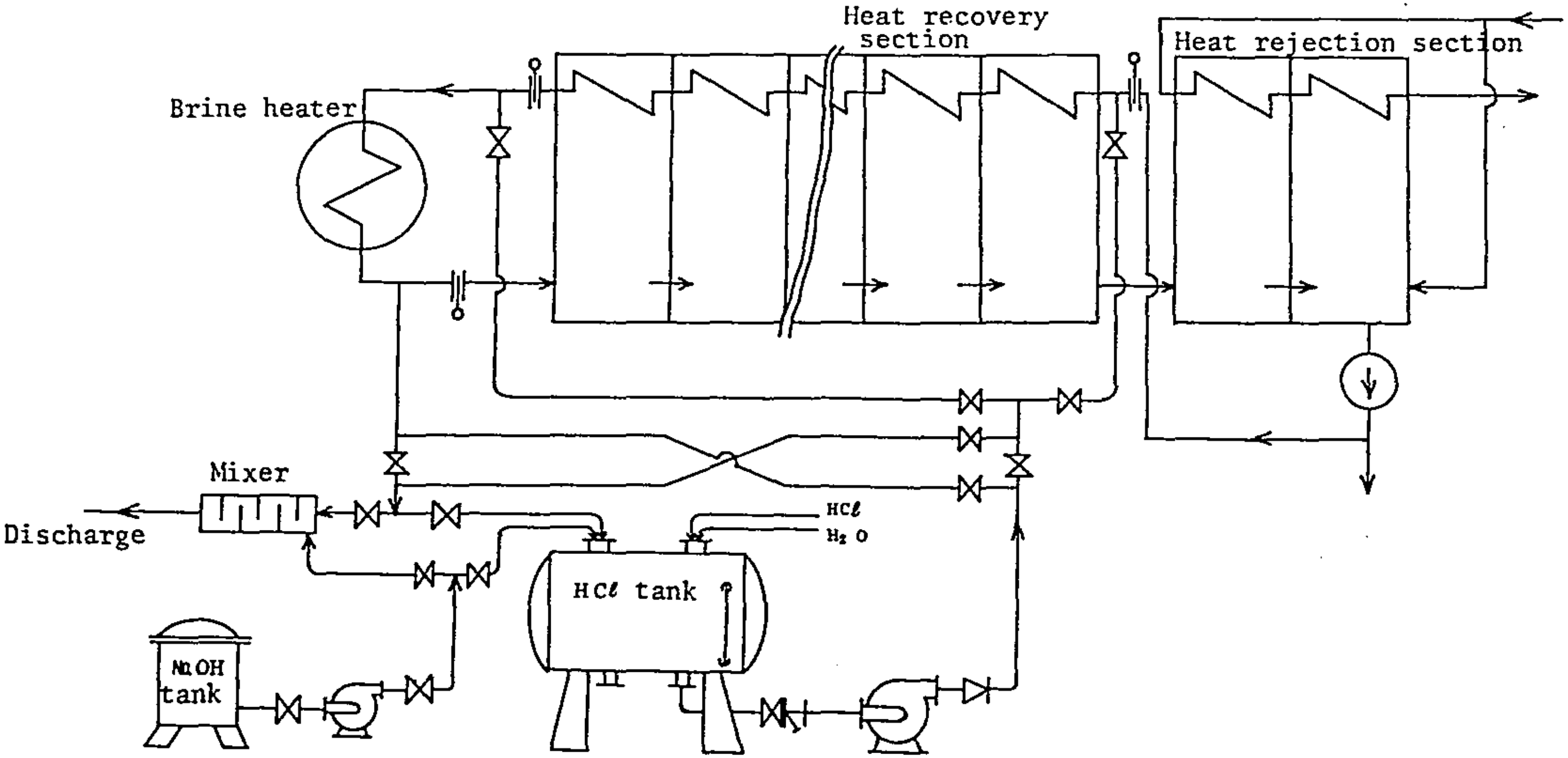


Fig. 5.6-32 Acid cleaning system

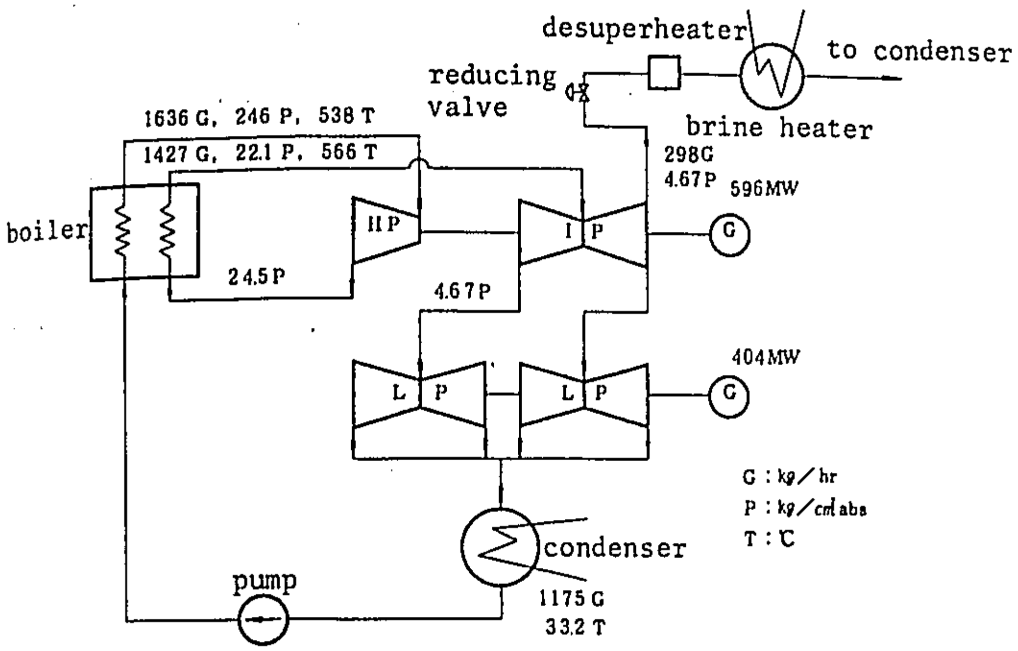


Fig. 5.7-1 Dual-purpose plant combining condensing type with extraction turbine

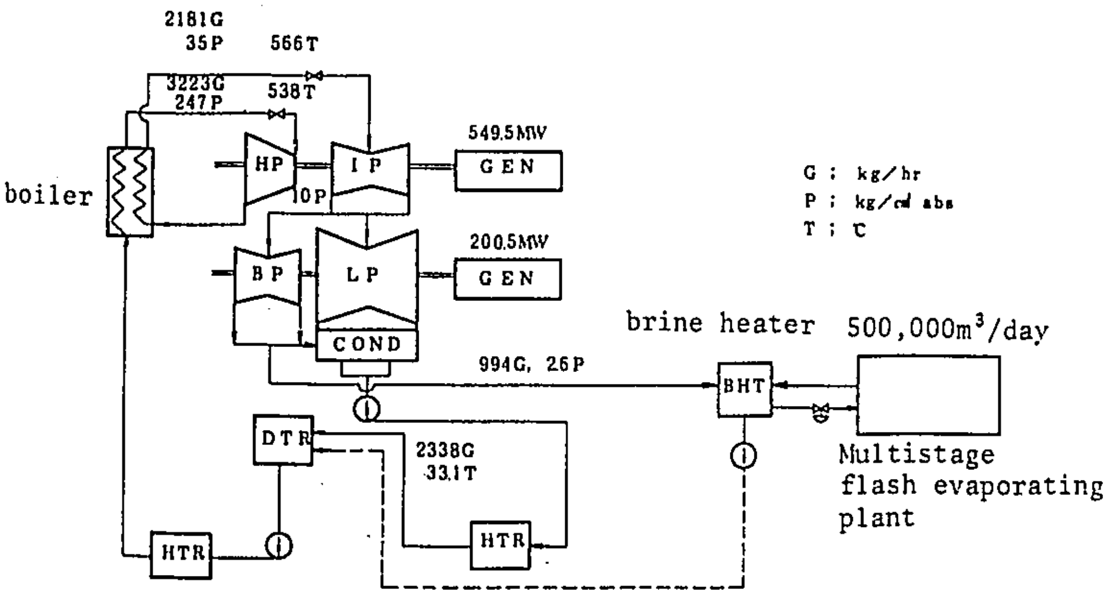


Fig. 5.7-2 Dual-purpose plant combining condensing type with back pressure turbine

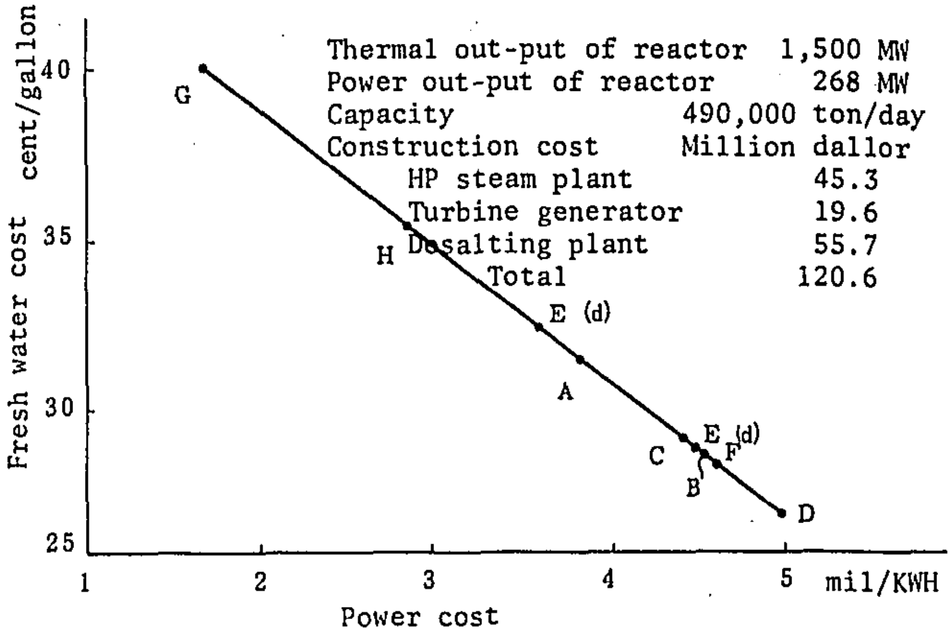


Fig. 5.7-3 Correlation power cost and water cost

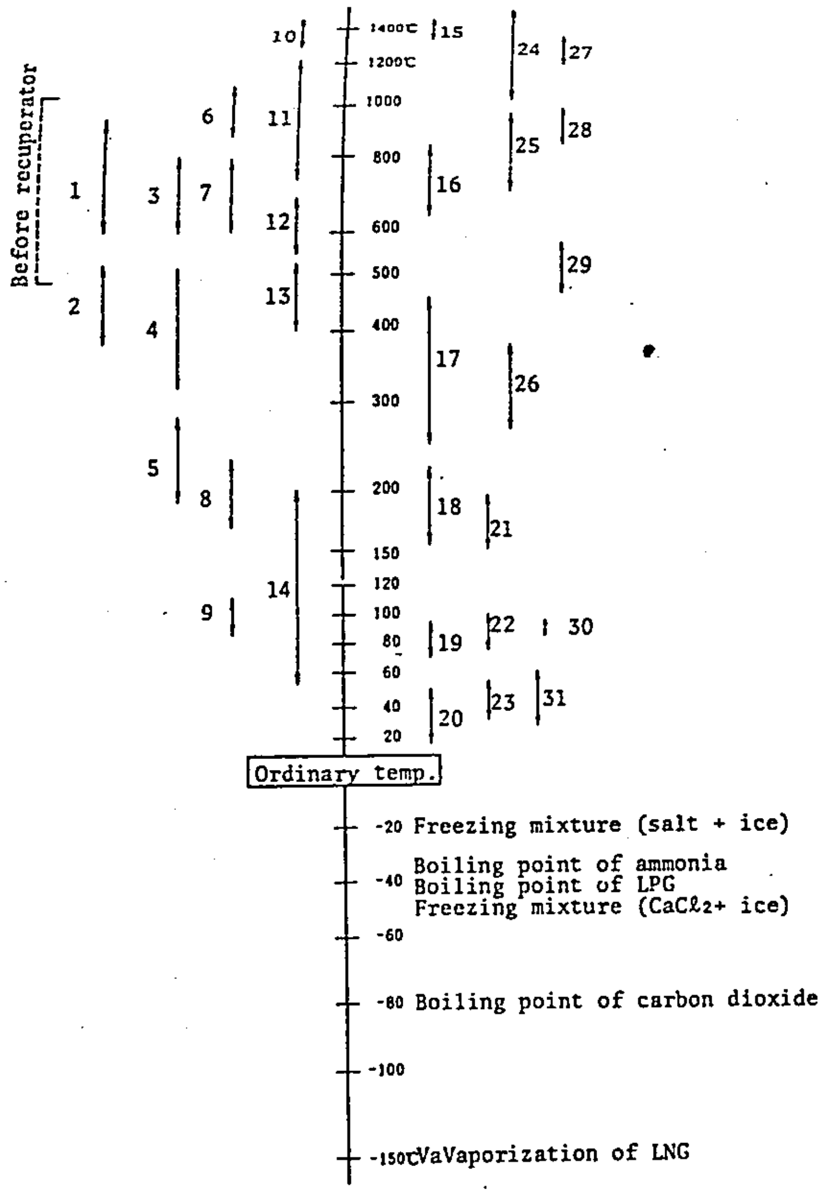
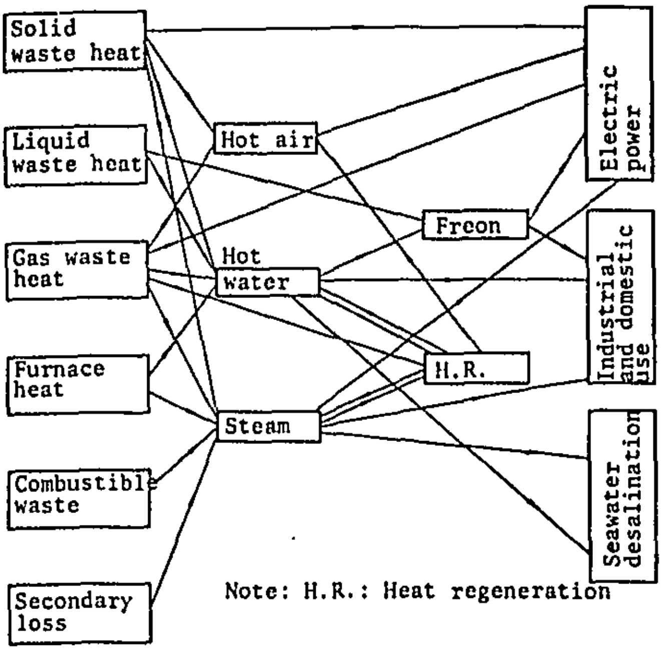


Fig. 5.7-4 Waste heat temperature range

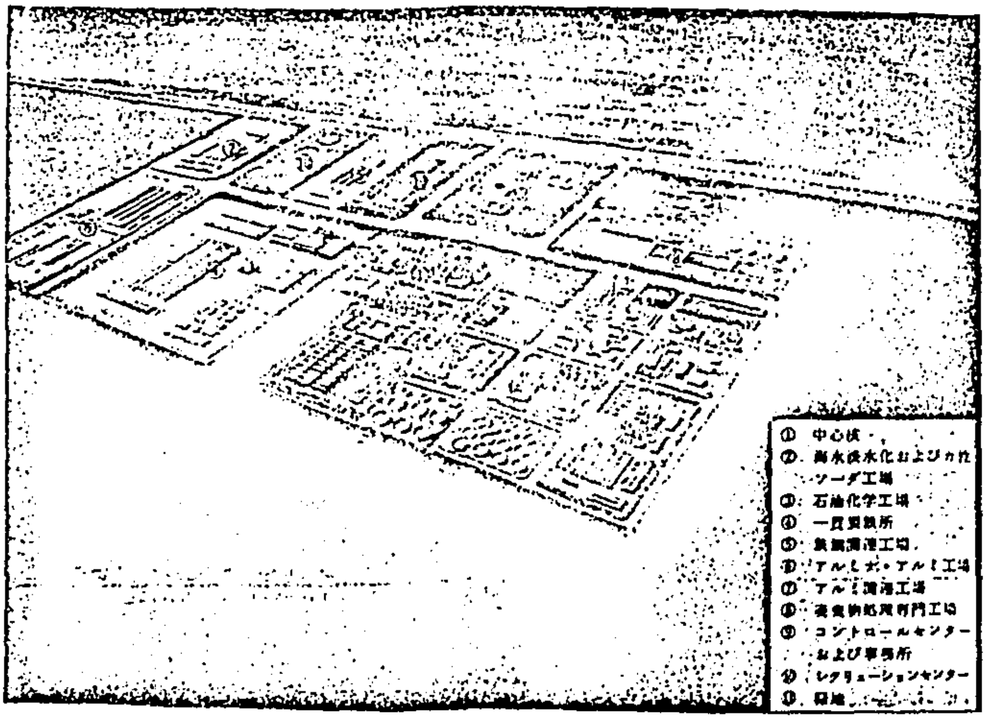
Note:

1. Glass crucible furnace waste gas
   Glass tank furnace waste gas
2. Zinc melting furnace waste gas
3. Rolled heating furnace waste gas
   Soaking pit waste gas
4. Molding furnace waste gas
5. Air heating furnace waste gas
6. Red coke
7. Coke oven gas
8. Coke oven waste gas
9. Brewing kiln steam
10. Converter gas
11. Copper smelting waste gas
12. Gas turbine waste gas
13. Internal combustion engine waste gas
14. Turbine back pressure steam
15. Blast furnace slag'
16. Sintered ore
17. Boiler waste gas
18. Sintering furnace waste gas
19. Evaporator condensate
20. Turbine cooling water
21. Blast furnace gas
22. Cooking waste liquid
23. Blast furnace cooling water
24. Converter steel ingot
25. Bloom
26. Brick calcined waste gas
27. Rolling mill heating furnace steel material
28. Slab
29. Rolled steel products
30. Deying waste liquid
31. Bloom spray cooling water
    Converter cooling water
    Heating furnace cooling water



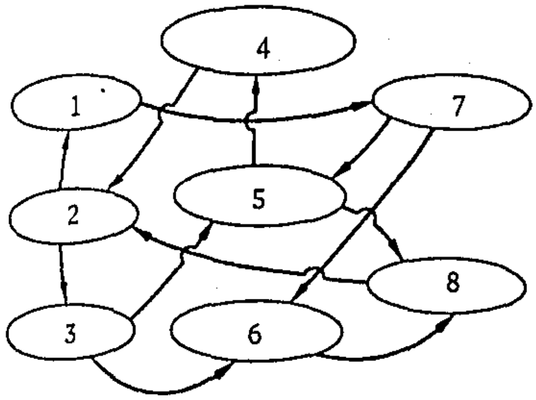
Note: H.R.: Heat regeneration

Fig. 5.7-5 Waste heat recovery and utilizing process



1. Utility center
2. Desalting plant and caustic
3. Petrochemical plant
4. Integrated steel plant
5. Steel-related industries
6. Alumina-aluminium plant
7. Aluminium-related industries
8. Industrial waste treatment plant
9. Control center and administration
10. Recreation center
11. Green area

Fig. 5.7-6 Nuclear power steel-chemical industry complex



Note:
1: Regional development sector
2: Economic growth sector
3: Correlative industries sector
4: Resources and energy sector
5: Sea water desalination sector
6: Water reusage technology sector
7: Water supply and demand sector
8: Environmental technology sector

Fig. 5.8-1 Correlation between total system and sectors in field of seawater desalination

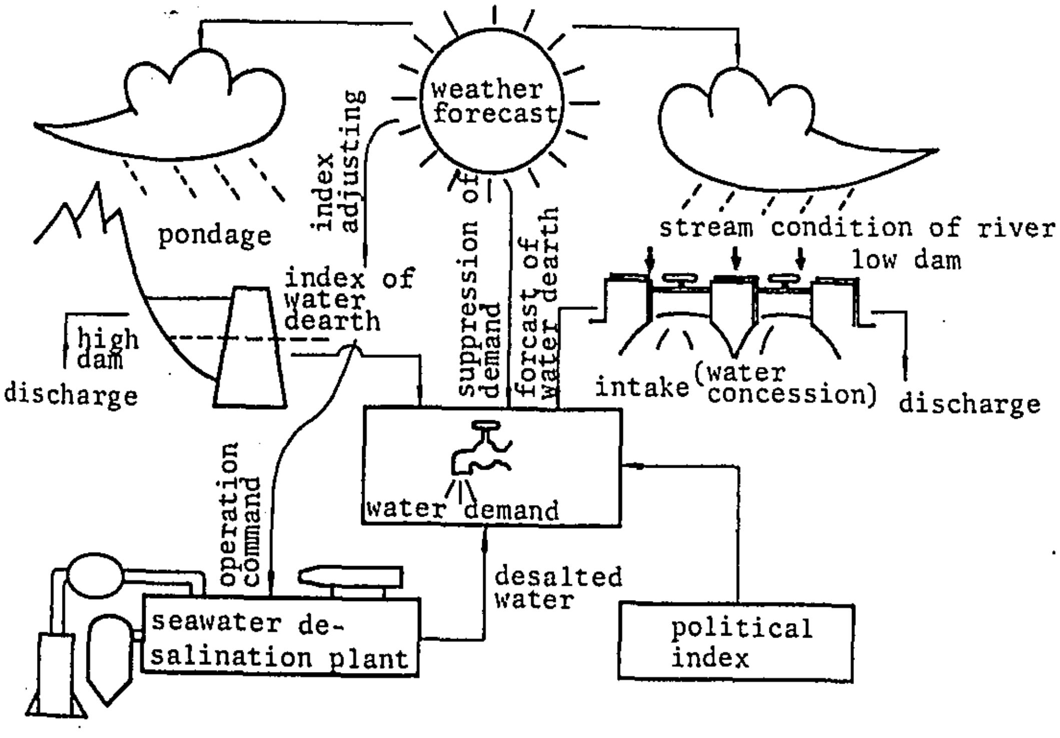


Fig. 5.8-2 Water supply system

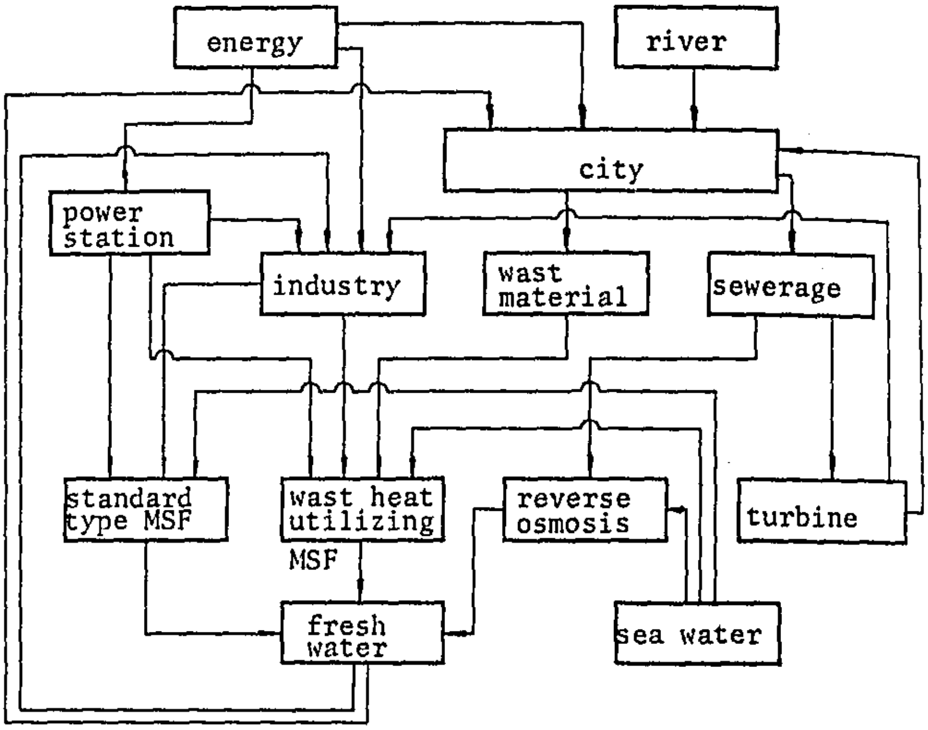


Fig. 5.8-3 Total system of desalination

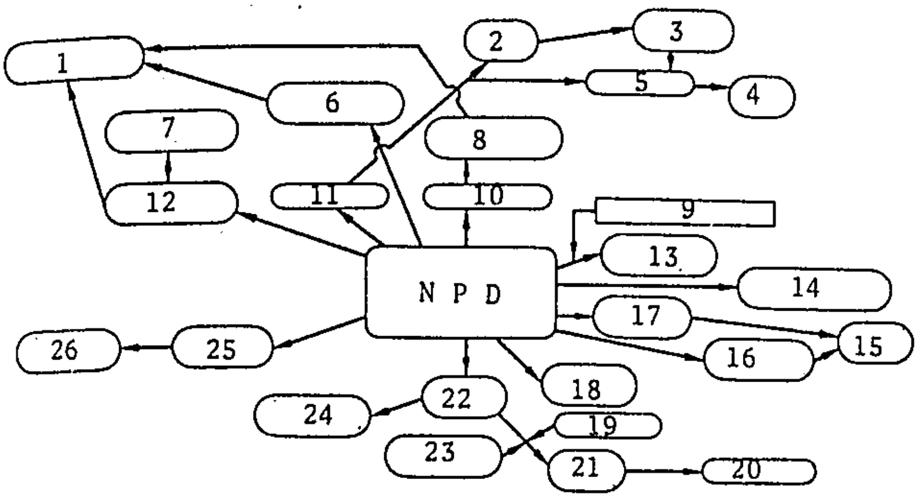


Fig. 5.8-4 Related group of nuclear power-desalination dual purpose plant

NPD Nuclear power-desalination dual purpose plant
1. Fossil energy saving
2. Settlement of water shortage
3. Decrease of intake dam construction
4. Arrangement of water distributing station
5. Dependence on desalination
6. Diversification of energy utilization
7. Advanses of equipment reliability
8. Efficient utilization of energy resources
9. Completion of monitoring
10. Efficient heat utilization
11. Production of desalted water
12. Development of multiple-purpose atomic energy utilization technology
13. Uncertainly of radiocative pollution
14. Reduction of air pollutants (SO_2, NO_2, etc.)
15. Stabilization of power supply
16. District service oriented plant location
17. Reduction of power cost
18. Change of seaside view
19. Promotion of thermal diffusion
20. Influence on fishery
21. Influence on marine organism
22. Discharge of concentrated hot wast water
23. Dilation by cooling seawater
24. Removal of mineral resources from seawater
25. Progress of desalination technology
26. Technical assistance for arid countries

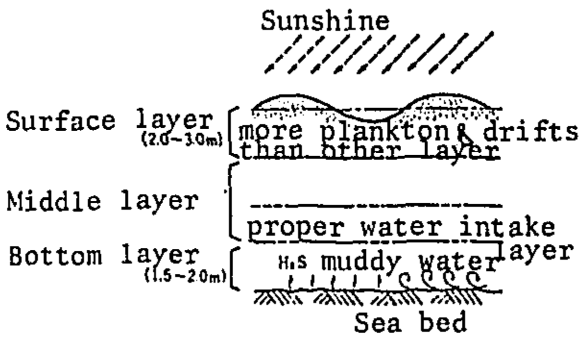
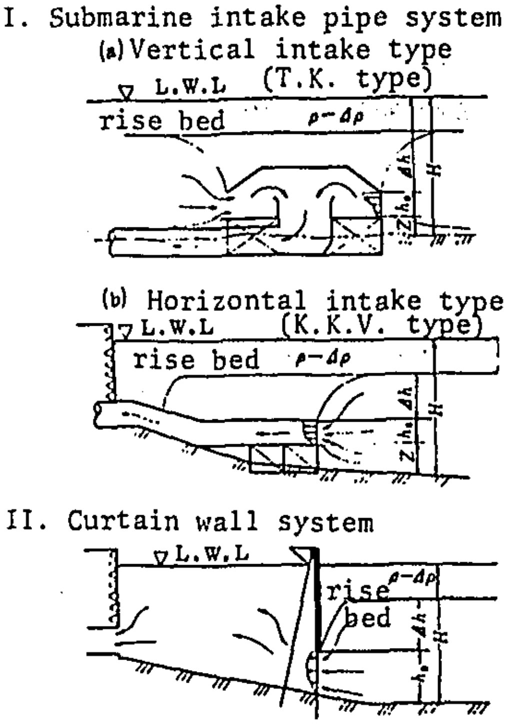


Fig. 5.9-1 Classification of water lot



·Fig. 5.9-2 Deep water intake system

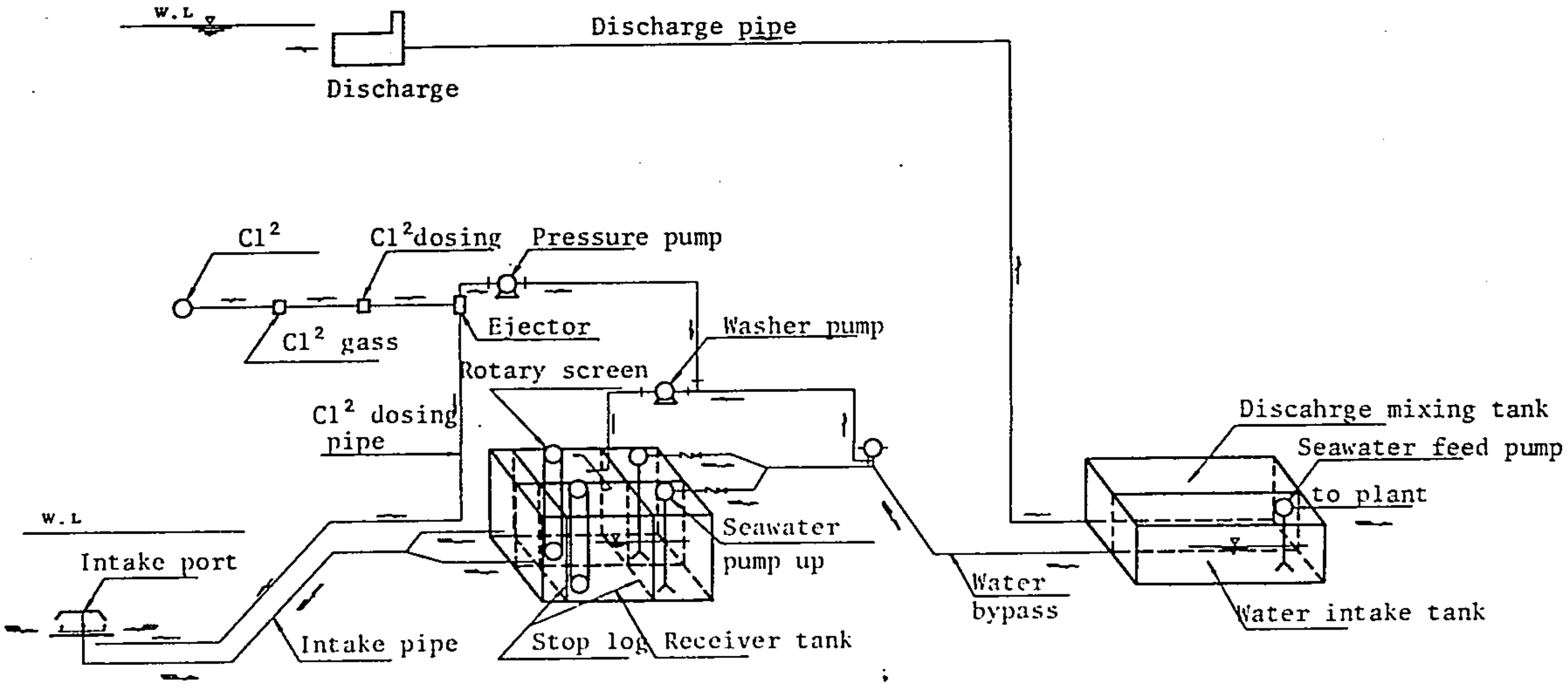
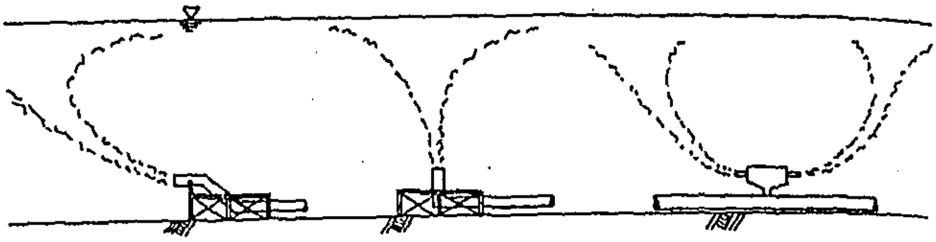


Fig. 5.9-3 Flow Sheet



Horizontal zetting type

Vertical zetting type

Multi-horizontal zetting type

Fig. 5.9-4 Submarine release type

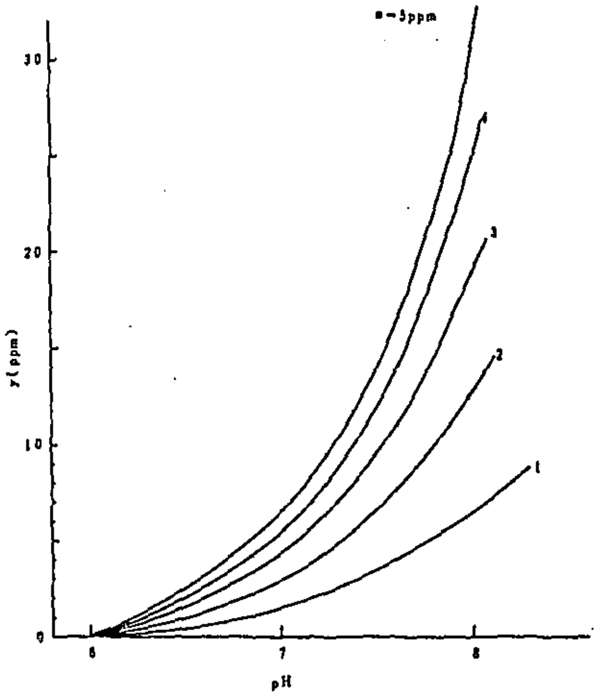


Fig. 5.9-5 Effect of pH on y value of ammonia - seawater under atmospheric pressure

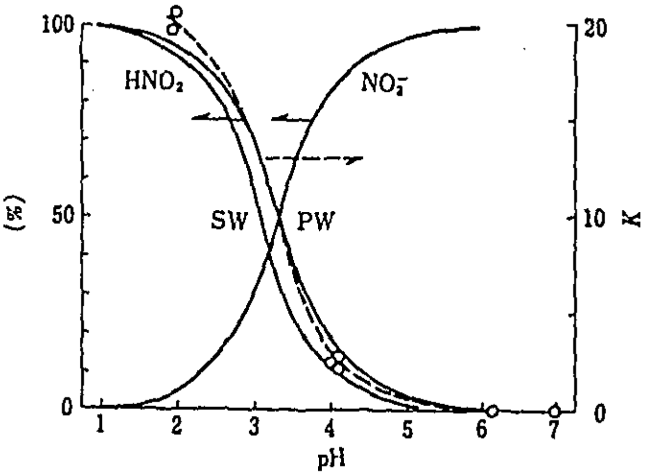


Fig. 5.9-6 Curve of K value and equivalent concentration in nitrous acid - water system

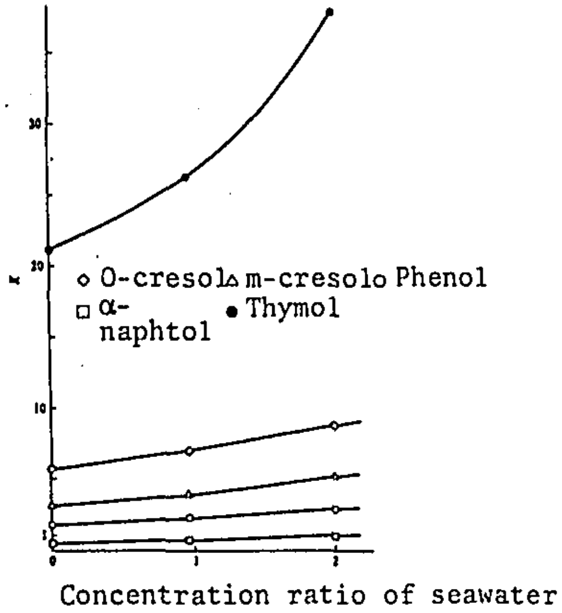


Fig. 5.9-7 Effect of salinity on K value at atmospheric pressure, pH=7, x=1 ppm

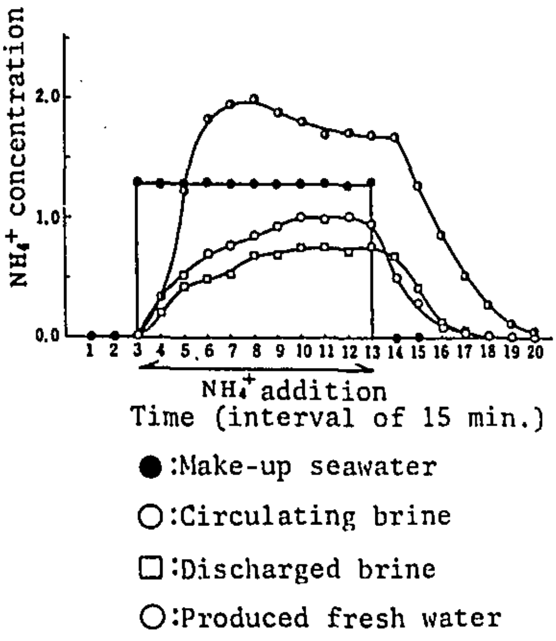
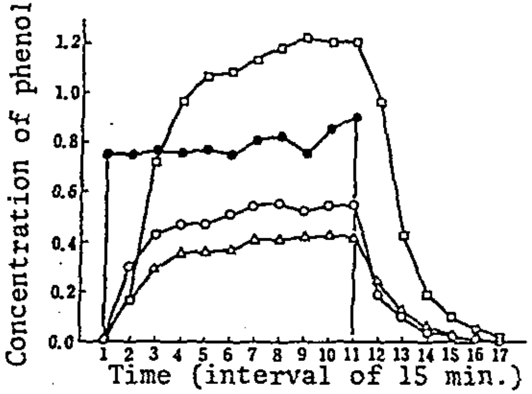


Fig. 5.9-8 Change of NH_4^+ concentration after $NH_4C\ell$ addition to flash evporator



● : Supply seawater (after decarbonation)
○ : Circulating brine
△ : Discharged brine
□ : Produced fresh water

Fig. 5.9-9 Partition of phenol to brine and fresh water after addition of phenol to flash evaporator

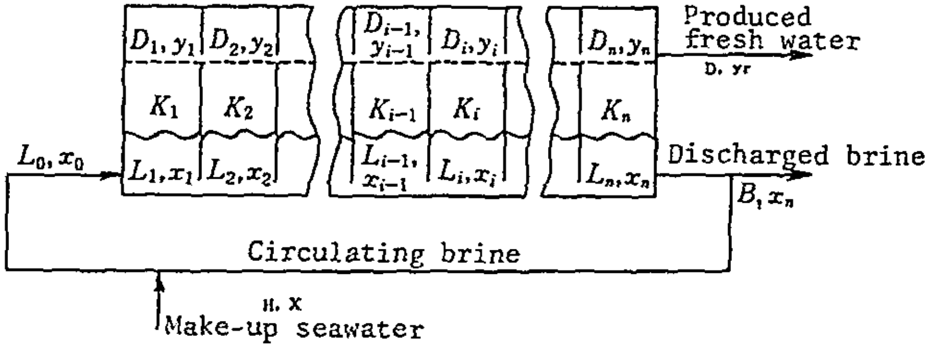


Fig. 5.9-10 Numerical value model in computer simulation

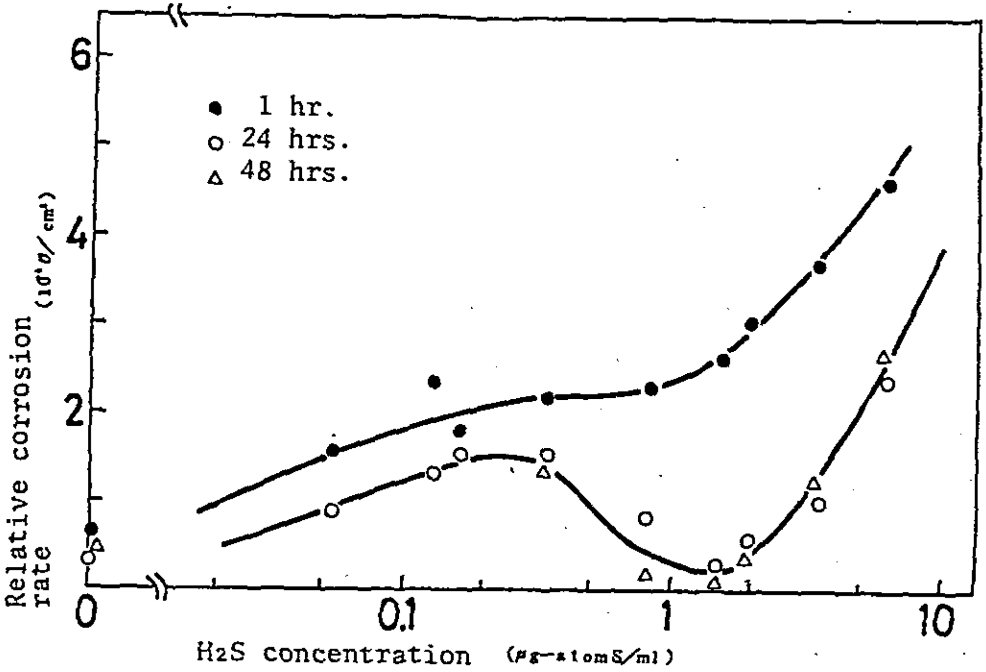


Fig. 5.9-11 Relation between H_2S concentration and corrosion rate of mild steel
(35°C, synthetic seawater, at constant concentration of H_2S)

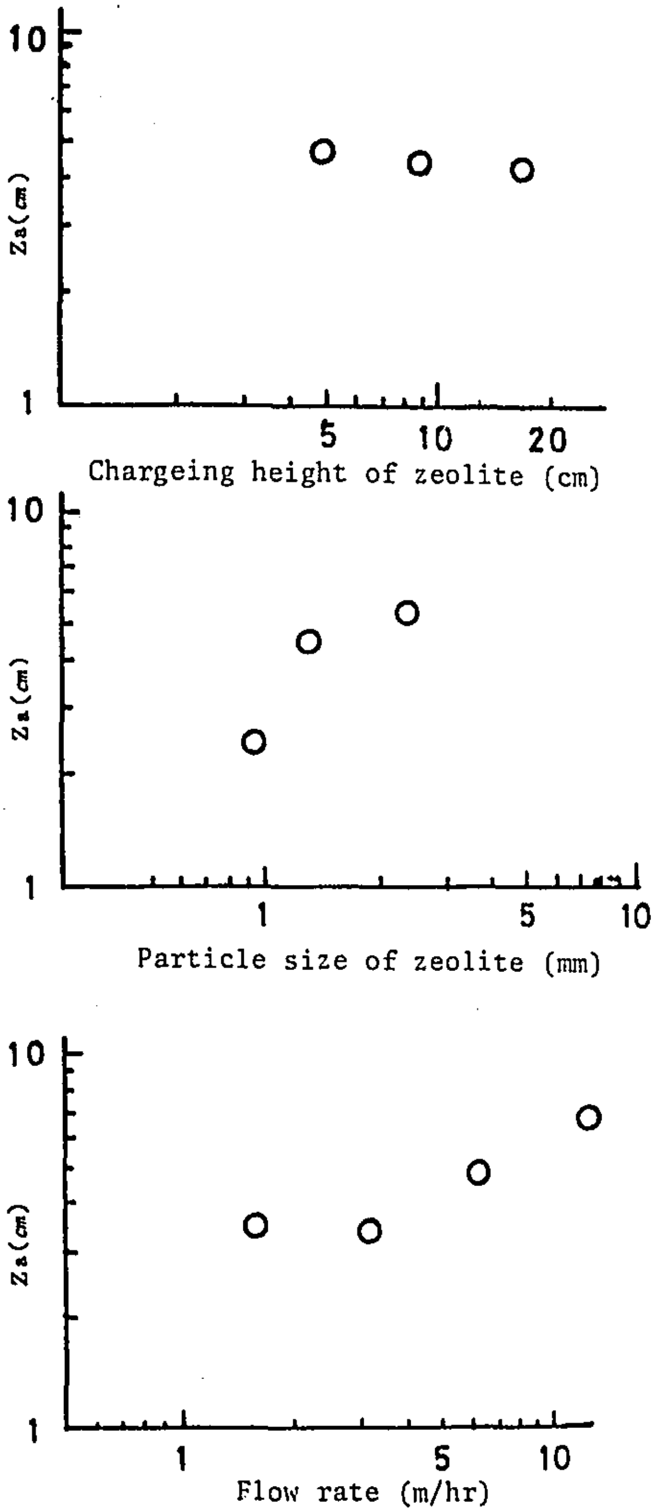


Fig. 5.9-12 Effect of various factors on height of adsorption zone of zeolite

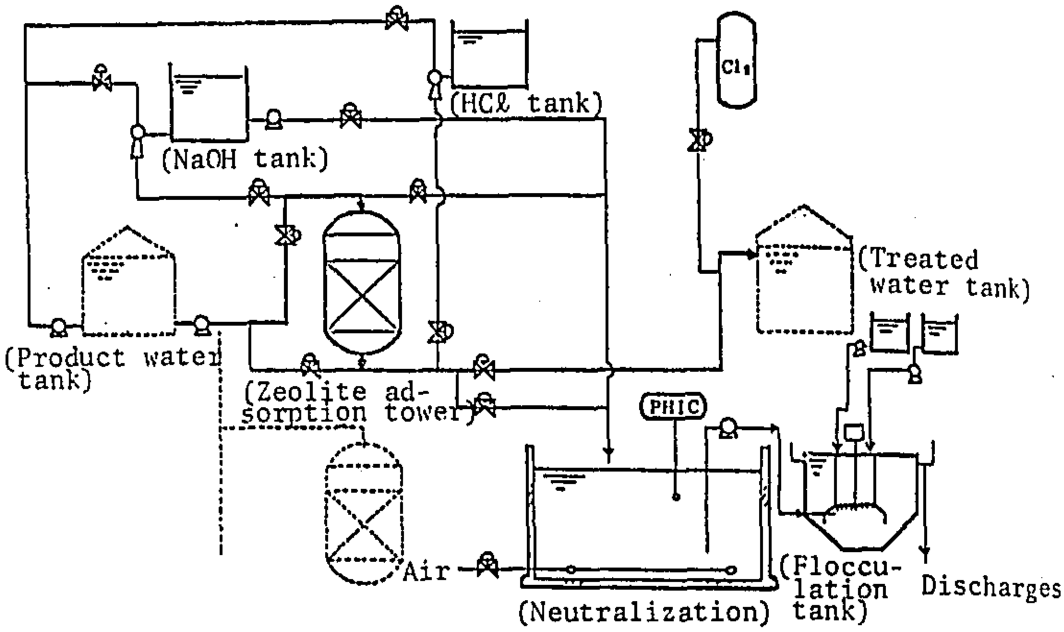


Fig. 5.9-13 Basic flow sheet of zeolite adsorption process

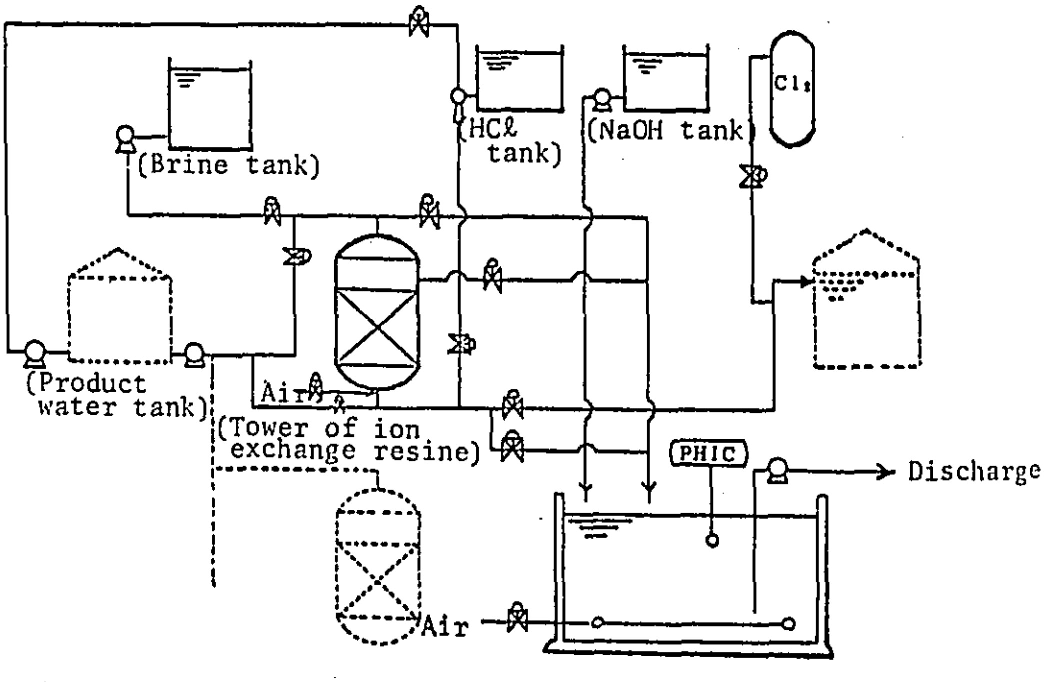


Fig. 5.9-14 Basic flow sheet of ion exchange resine process

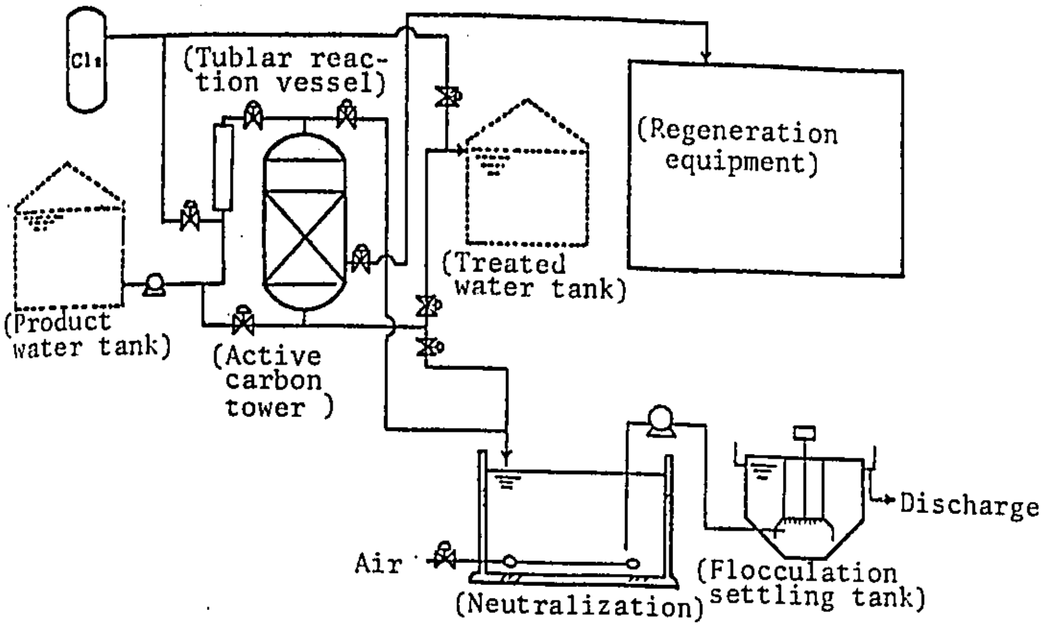


Fig. 5.9-15 Basic flow sheet of active carbon adsorption process

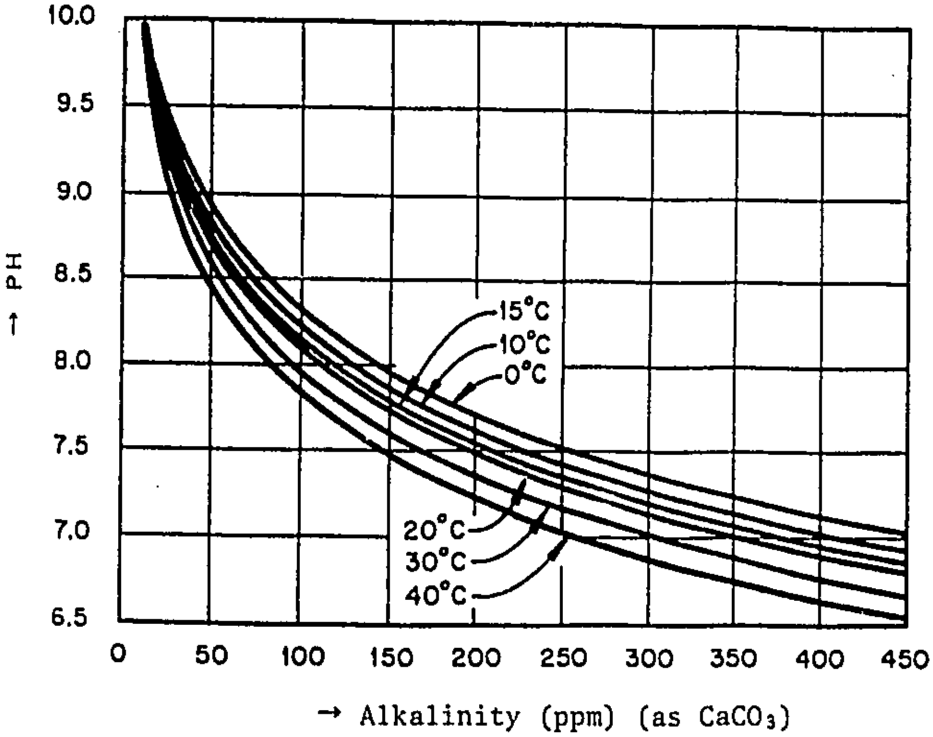


Fig. 5.9-16 Solubility of calcium carbonate at various pH values

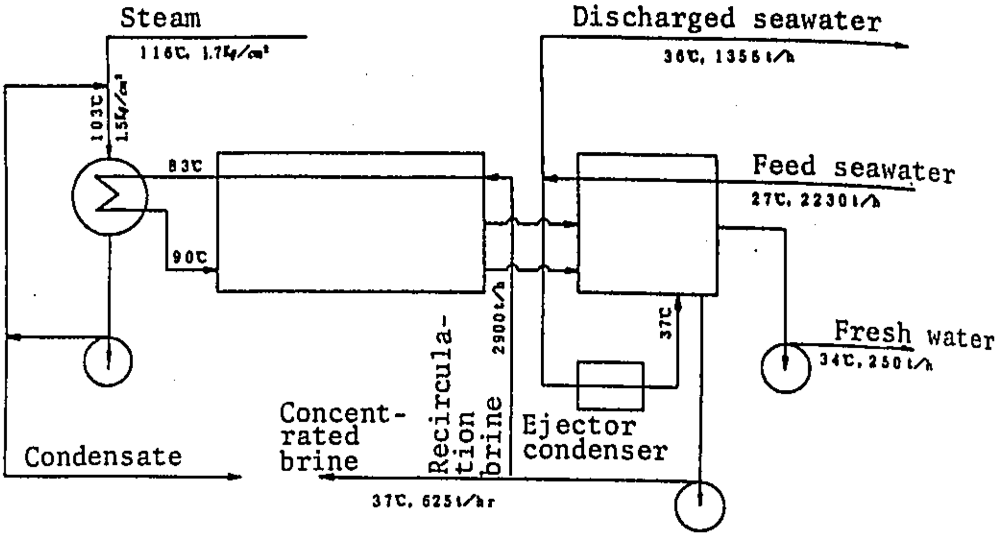


Fig. 5.9-17 Heatbalance of the multi-stage flash evaporation desalting plant at Benghazi city, Libya

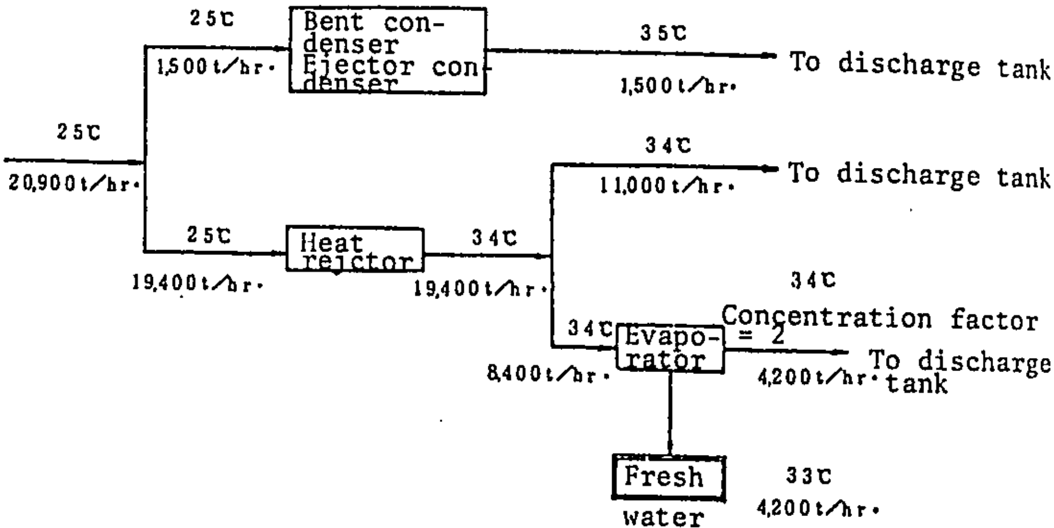
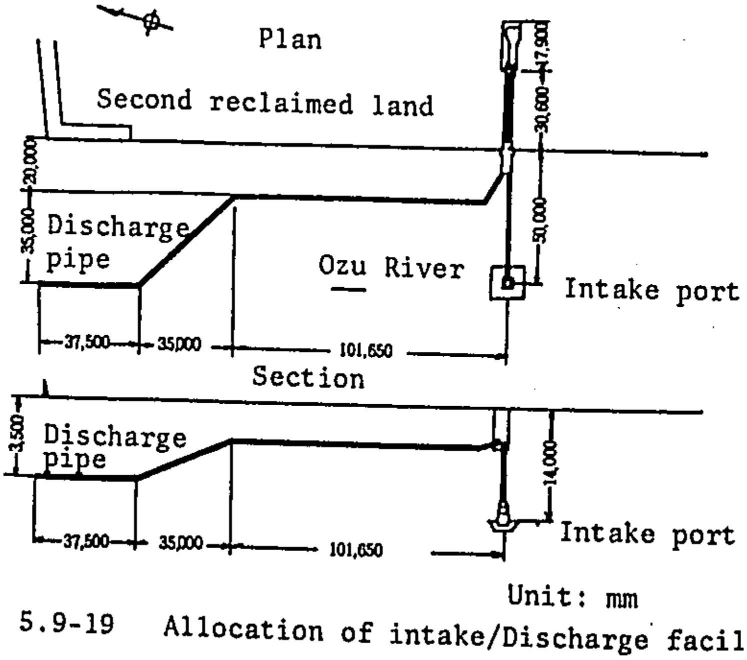


Fig. 5.9-18 Specifical flow rate of the discharged brine of a 100,000 m /day MSF desalting plant



Unit: mm

Fig. 5.9-19 Allocation of intake/Discharge facilities

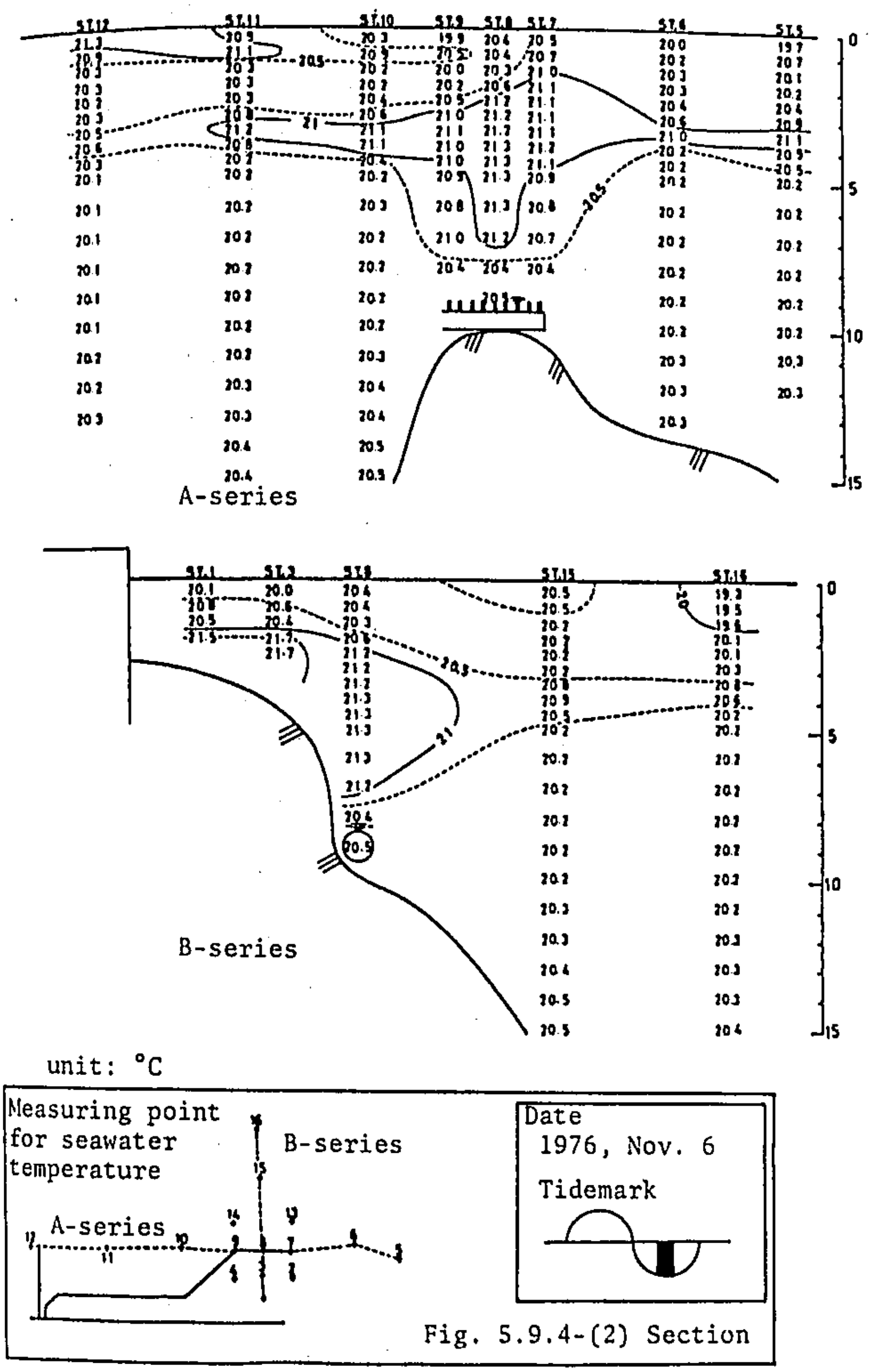


Fig. 5.9-20 Diffusing condition by multi-port nozzle

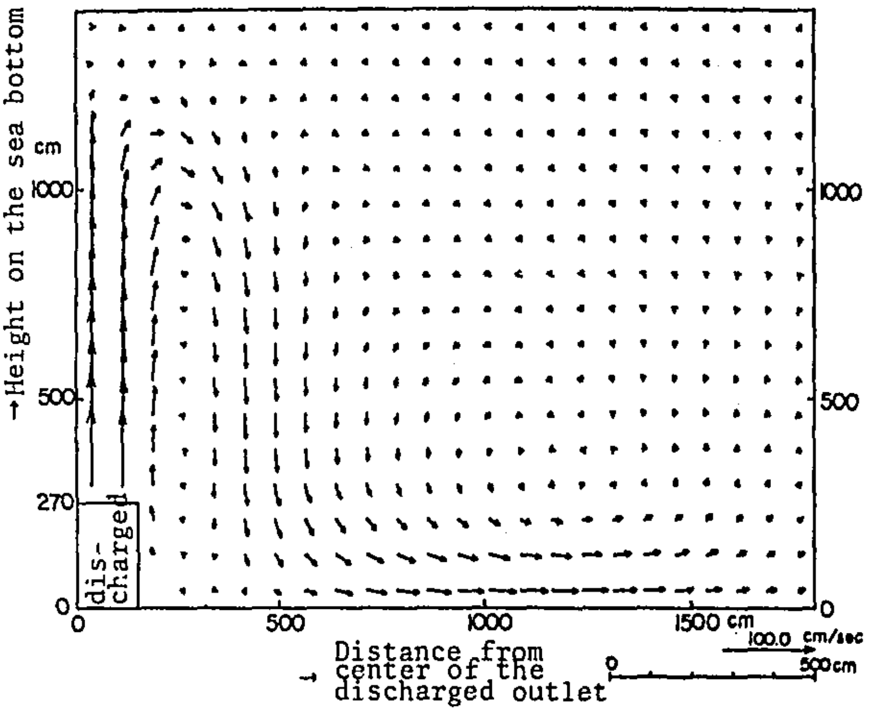


Fig. 5.9-21 Distribution of flow velocity across the section of central axis of the plume ($Q=5.6\ m^3/s$, $\Delta S=7.36‰$, $\Delta T=7℃$)

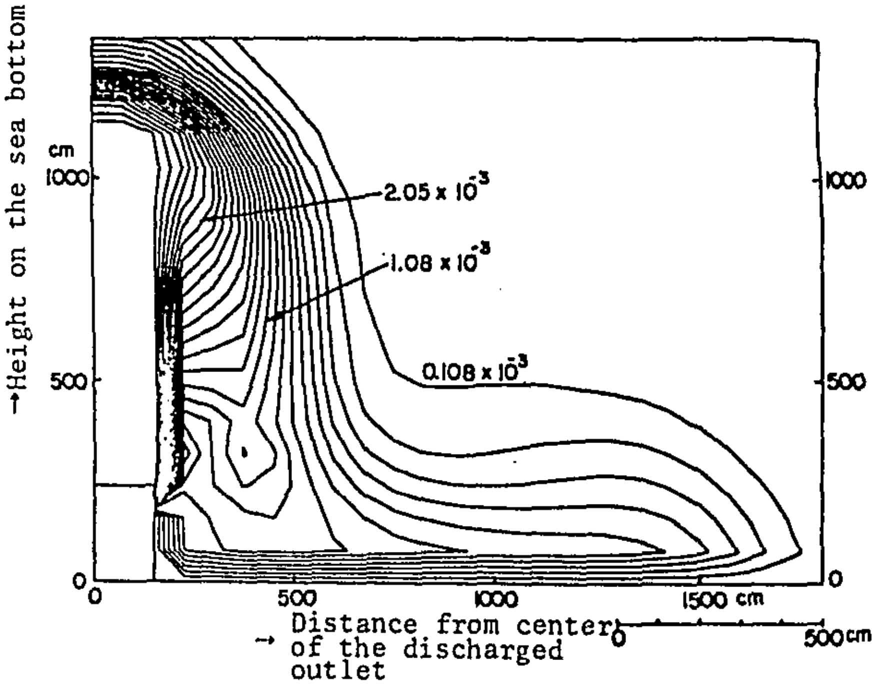


Fig. 5.9-22 Vertical distribution of the densities ($Q=5.6\ m^3/s$, $\Delta S=6.36‰$, $\Delta T=7℃$)

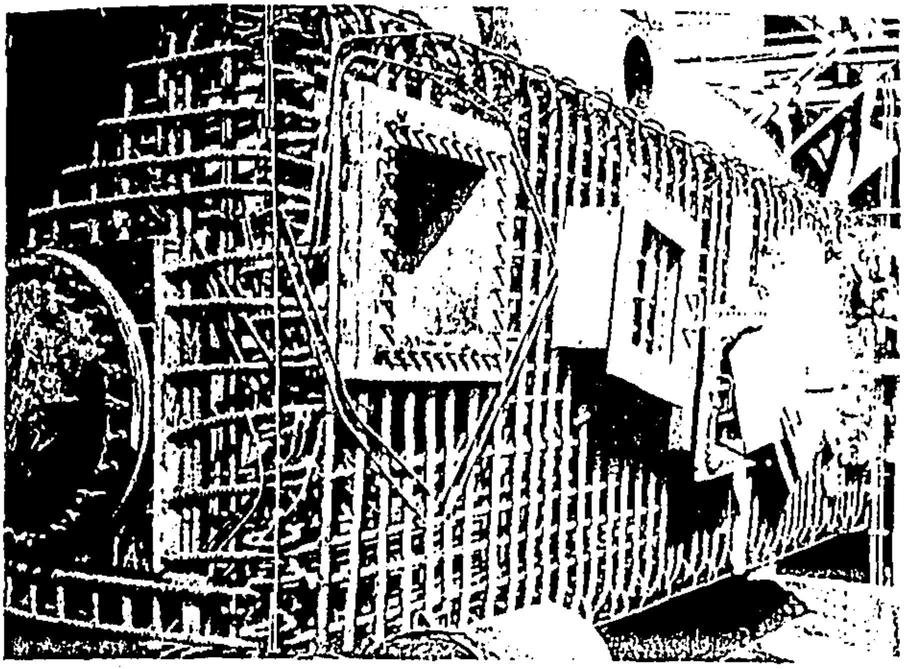


Photo. 5.5-1 RC-concrete vessel at San Diego Test Facilities

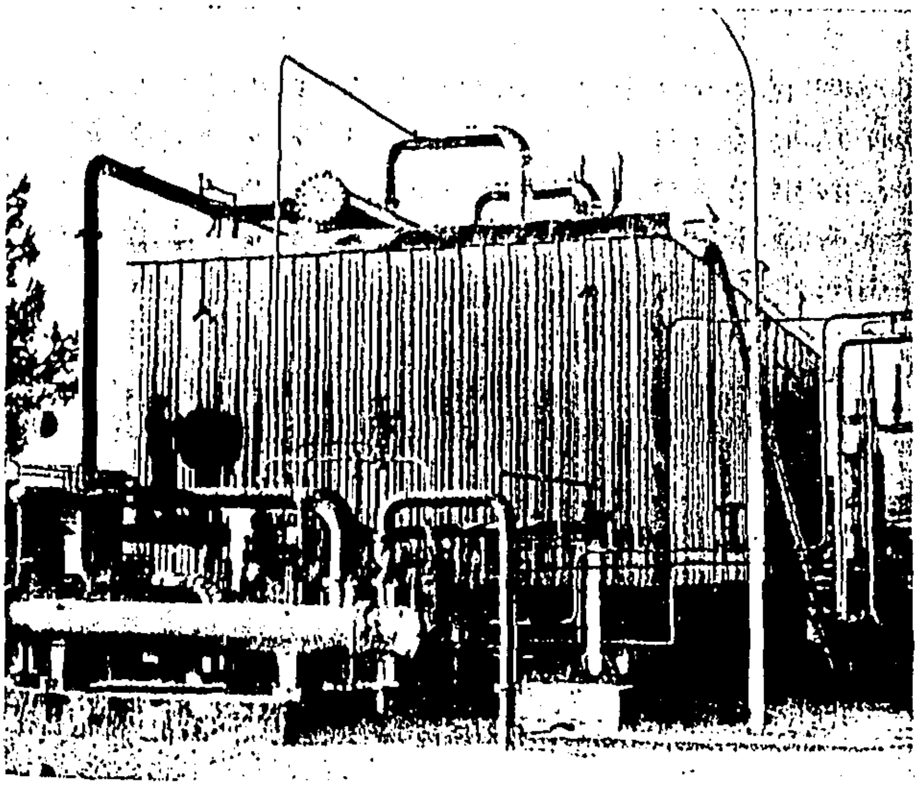


Photo. 5.5-2 Pilot plant built in France

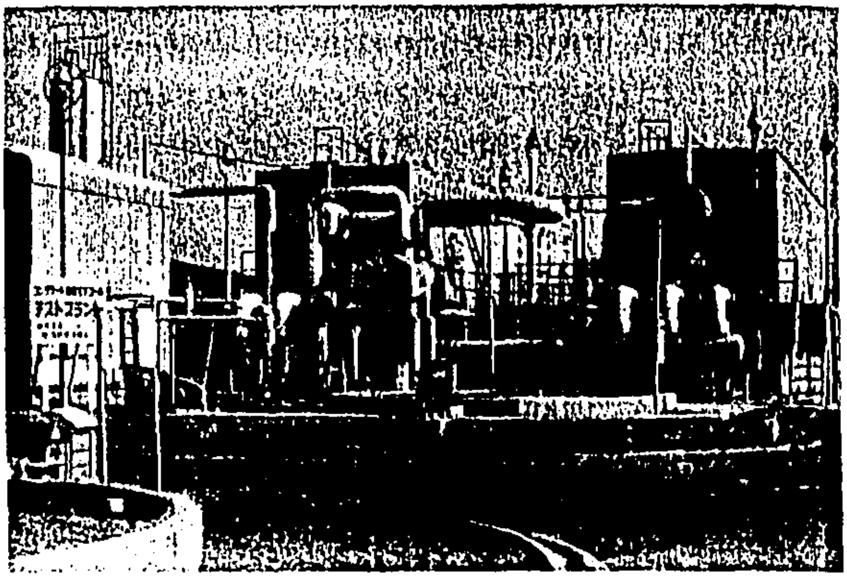


Photo. 5.5-3 Concrete vessels at Chigasaki Test Facilities

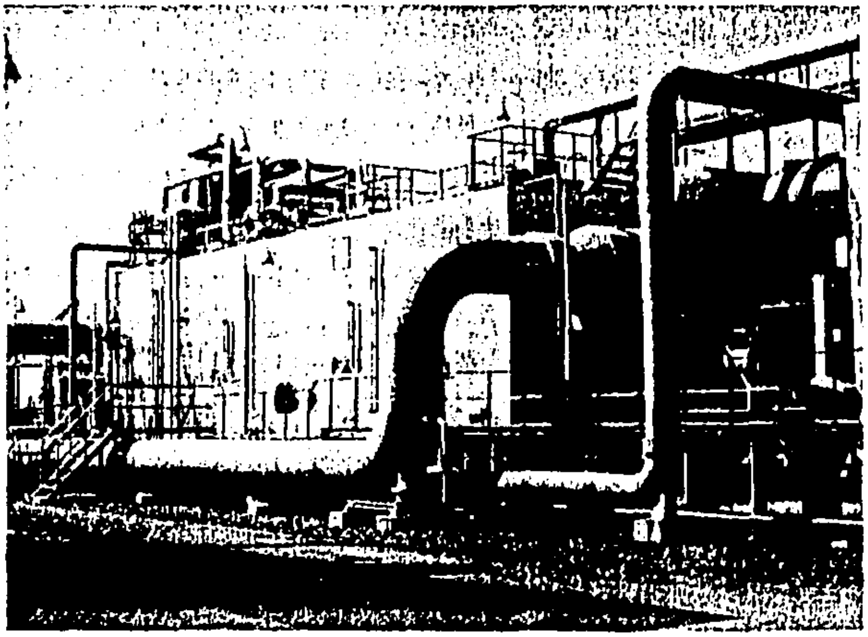


Photo. 5.5-4 Heat rejection section made of RC concrete at Ōita Test Facilities

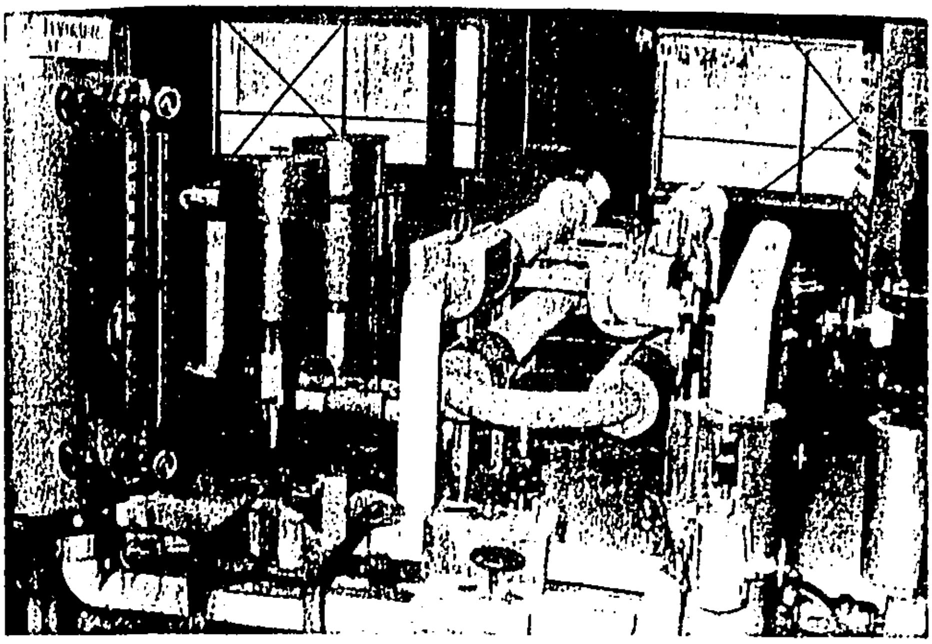


Photo. 5.5-5 Apparatus for testing corrosion resistance of concrete

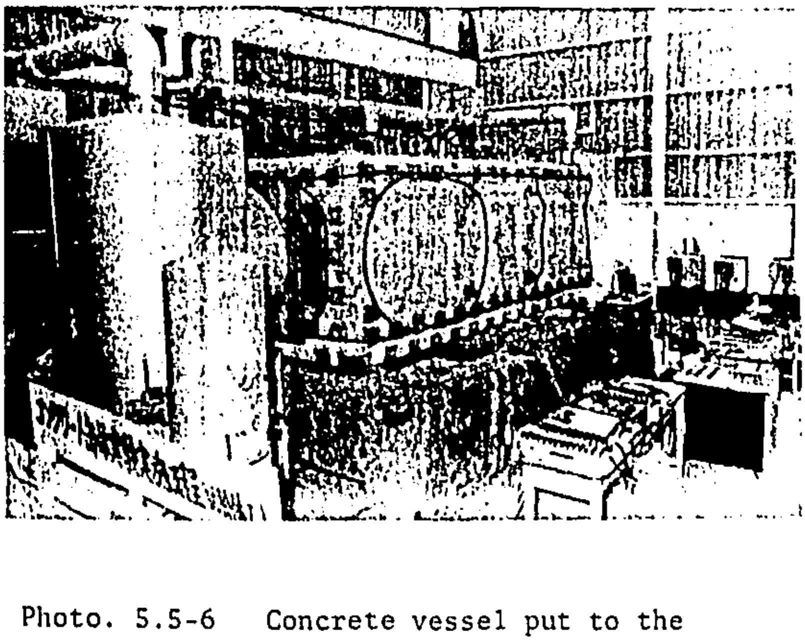


Photo. 5.5-6 Concrete vessel put to the thermal test

Photo. 5.5-7 Lining of steel vessel with cement mortar

Table 5.1-1 Main Components of Multistage Flash Distillation Plant

- Evaporator (heat rejection section and head recovery section)
  - Heating tube
  - Orifice
  - Condensate tray
  - Heating tube plate
  - Shell
  - Steam extraction valve
  - Stage bulkhead
  - Demister
- Brine heater
  - Heating tube
  - Shell
  - Steam extraction valve
  - Heating tube plate
  - Safety valve
  - Drain collector
- Pre-treatment facilities
  - Decarboxylation apparatus
  - Acid mixer
  - Reducing agent tank*
  - Sulfuric acid tand
  - Vacuum deaerator or last stage deaerator
  - Alkali tank
- Intake and discharge facilities
  - Seawater intake pipe
  - Seawater pit
  - Chlorination plant
  - Brine discharge pipe
  - Waste water pit
  - Screen
- Steam extraction apparatus
  - Ejector
  - Condenser
- Pumps and driving mechanism
  - Seawater feed pump and motor or turbine
  - Brine circulating pump and motor or turbine
  - Brine discharge pump and motor
  - Product water pump and motor
  - Condensate pump and motor
  - Make-up seawater feed pump and motor
  - Acid supply pump
- Instruments
  - Temperature recorder/regulator
  - Flow recorder/regulator

Liquid level regulator

pH regulator

Instrument power switchboard

Temperature recorder

Flow recorder

Thermometer

Flow-meter

Seawater monitor

pH meter

Concentration meter

Automatic control valve

Instrument air supply system

Ball cleaning facilities *

Sponge ball feeder

Sponge ball recovery device

Pickling and alkali cleaning system *

Steam generating system *

Boiler
Boiler water feed system
Fuel tank

Note: * Not always required.

Table 5.1-2 Performance ratios calculated from the simulation program in various partial load operations

T_{Bmax} (℃)	120.35	118.0	115.0	110.0	105.0	100.0
W_R (m^3/h)	993.0	960.0	900.0	800.0	750.0	650.0
Heating Steam (m^3/h)	13.73	13.64	13.59	13.48	13.18	13.00
Product Water (m^3/h)	144.3	136.0	123.3	103.2	90.7	73.4
Performance Ratio	10.51	9.97	9.07	7.65	6.88	5.64

Table 5.2-1 Composition of seawater

Component and symbol		Composition ratio (ppm)
Sodium	Na^+	10,561
Magnesium	Mg^{++}	1,272
Calcium	Ca^{++}	400
Potassium	K^+	380
Chloride	Cl^-	18,980
Sulfate	SO_4^{--}	2,649
Bicarbonate	HCO_3^-	142
Bromide	Br^-	65
Others	—	34
Total disolved solid		34,483

Table 5.2-2 Classification of heat tube fouling materials

Division	Designation	Composition
Deposits of slightly soluble salt	Sulphuric acid scale (hard scale)	Anhydrus calcium sulphate $CaSO_4$
		Calcium sulphate semi-hydrate $CaSO_4 \cdot \frac{1}{2}H_2O$
		Calcium sulphate dihydrate $CaSO_4 \cdot 2H_2O$
	Alkaline scale (soft scale)	Calcium carbonate
		Magnesium hydroxide $Mg(OH)_2$
Suspensions in brine	Sludge	Dirt, organic matter (inclusions in seawater
		Metallic oxide (corrosion products)

Table 5.2-3 Assortment of Fouling Prevention Methods

	Classification	Typical example	Method	Foul matter		
				Sulphuric acid scale	Alkali scale	Sludge
1	Operating conditions to be arranged outside the boundary of scale deposition.	Unsaturation maintenance method	Brine temperature and concentration to be fixed below saturated solubility.	◎	×	×
2	Scale compositions to be removed by pre-treatment	Lime - Magnesium carbonate method	Calcium carbonate to be settled and separated by adding up Magnesium carbonate and lime.	○	△	×
		Acid injection · decarbonation method	Carbonate in the sea water to be changed into carbon dioxide gas and removed by adding the acid.	×	◎	×
3	Threshold point of scale deposition to be altered by chemical process.	pH control method	To be decarbonated by adding the acid and to change pH in brine.	×	◎	×
4	Scale bonding prevention by physico- chemical process	Chemical dosing	Scale compositions to be maintained in supersaturation or to be changed into sludge by adding chemicals.	△	○	△
5	Scale bonding prevention by physical process	Seeding		○	○	○
		Contact stabilizing method	Scale deposition chamber to be provided so as to depose and remove scale compositions.	△	△	×
6	De-scaling by mechanical process	Sponge ball cleanning	Incipient stage of scale bonding to be removed by sponge ball.	△	△	◎

Table 5.2-4 Comparison of Decarbonation System

System	Packed tower	Cascade fall	Air bubbling
Structure	Tower, Spray nozzle, Punched plate	Multistage of water channels	Water channel, Air pipe line
Auxiliaries	Feed water pump Blower	Feed water pump	Blower
Performance	5 ppm	15 ppm	15 ppm
Capital cost	100%	80%	80%
Running cost	100%	70%	70%
Evaluation			

Table 5.2-5 Comparison of Deaeration Systems

Type	Packed power	Final stage	Heat recovery stages
Structure	Tower Spray nozzle Punched plate	(Vessel) Spray nozzle Punched plate	(Vessel) Spray nozzle Punched plate Spray tray Deaeration nozzle
Performance	Less than 10 ppb	20 ppb	15 ppb
Plant cost	100%	20%	40%

Table 5.3-1 Measurement scope by NETD

Chamber length (m)	Liquid depth (m)	Liquid flow (10^2ton/m·hr)	Evaporation temp. (°C)	F.T.D. 〔℃〕	Note
3.45	0.31~0.56	3.8~15.8	23.9~55.3	1.2 ~ 4.2	1) Empty stage
4.8	0.36~0.55	7.4~15.3	41.9~60.9	1.1 ~ 3.0	1) 〃
4.8	0.33~0.65	7.4~15.8	40.7~81.1	1.0 ~ 3.8	9) 〃
2.13	0.23~0.56	13.0~15.7	46.6~93.3	0.8 ~ 2.6	10) 〃
〃	0.41~0.56	10.7~13.3	43.2~51.9	1.2 ~ 2.8	10) accelerator
〃	0.45~0.51	9.4~11.0	39.5~48.2	1.8 ~ 3.4	11) 〃
3.45	0.46~0.56	12.3	37.2~45.1	2.1 ~ 2.6	12) MSF module
4.88	0.32~0.5	3.6 ~5.1	38℃	1.5	13) MEMS module
1.0	0.38~0.73	0.34~0.83	30 ~72.3	0.8 ~ 4.1	4,8) Empty stage, with baffle plates

Note: F.T.D.: Flash temperature differential

Table 5.3- 2 Configuration of heat tube and pitch (unit:mm)

Tube outer diameter	Pitch	Distance between partition groove center and tube center	Tube layout
19 (19.05)	25	19	Triangle or square layout
25.4	32	22	
31.8	40	26	
38.1	48	30	

Table 5.3-3 Thermal conductivity of heat tube materials

Material		Thermal conductivity [kcal/m·hr·℃] 20℃
Alminium brass	90 : 10	86.5
Cupro-Nickel	90 : 10	36.0
Cupro-Nickel	70 : 30	25.2
Copper		338.8
Stainless steel 316		11
Monel		17
Titanium		14.0
Steel		64.1

Table 5.3-4 Fouling factor

Water fouling factor	m^2·hr·℃/kcal			
Heating agent temp.	115℃ >		116~205℃	
Water temp.	52℃ >		53℃ <	
Flow velocity / Kind of water	1 m/s>	1 m/s<	1 m/s >	1 m/s <
Seawater	0.0001	0.0001	0.0002	0.0002
City water, underground water	0.0002	0.0002	0.0004	0.0004
River water* (minimum values)	0.0004	0.0002	0.0006	0.0004
Engine jacket	0.0002	0.0002	0.0002	0.0002
Distilled water	0.0001	0.0001	0.0001	0.0001
Boiler feed water	0.0002	0.0001	0.0002	0.0002
Fouling factor of others than water	m^2·hr·℃/kcal			
Steam (non-oiled)	0.0001			
Steam (oiled)	0.0002			
Compressed air	0.0004			

Table 5.3-5 Performance Data of Mist Eliminator

Eliminator type		Cyclone	Wire mesh	Collision plate type
Performance	Critical flow velocity	25 m/s	2 m/s	5.5 m/s
	Critical liquid drip dia	80 μ	10 μ	30 μ
	Pressure drop	Large	Small	Small
	Allowable liquid drop load	Large	Small	Large
Outstanding points	Merit	o Simple construction	o High arresting efficiency	o Few clogging
		o Few clogging	o Pressure small	
	Demerit	o Pressure drop large	o With liquid drop load	o Installation area, large
		o Installation area, large o Easy clogging	o Large chance of re-diffusion	o Large chance of re-diffusion

Table 5.3-6 Type of Steel Tube for Piping

Standard number	Description	Code	Operating classification	
			Pressure (kg/cm^2)	Temperature (°C)
JIS G3452	Carbon steel tube for piping	SGP	Less than 10	Less than 350
JIS G3454	Carbon steel tube for pressure piping	STPG	Less than 100	Less than 350
JIS G3456	Carbon steel tube for high temperature piping	STPT	Less than 300	Less than 450
JIS G3457	Arc welded carbon steel tube for piping	STPY	Less than 10	Less than 350
JIS G3459	Stainless steel tube for piping	SUSTP	Less than 65	Less than 800

Table 5.3-7 Standard Speed of Flow in Tube

Fluid	Use	Speed of flow m/s
Steam	Saturated steam	15 ∿ 20
	Superheated steam	20 ∿ 40
	Negative pressure steam	100
Liquid	Centrifugal pump suction system	2 ∿ 2.5
	Condensing pump suction system	Less than 1
	Pump discharge system	2 ∿ 3

Table 5.3-8 Characteristics of Various Vacuum Systems

Desctiption	Degree of vacuum		Characteristics
Nash pump	300 mmHg abs		Suitable for suction of moisture and air under low vacuum
Reciprocating vacuum pump	$1\sim10$	"	Capable of sucking substances containing a small amount of condensate gas. Requires periodical maintenance. Has a weakness in corrosion resistivity in some cases.
Water ejector	~50	"	Suitable for suction of substances with a large portion of vapor and a small portion of non-condensate gas
Steam ejector	$\sim10^{-2}$	"	Relatively free of trouble. Suitable for suction of condensate gas and non-condensate gas.
Oil ejector booster	$1\sim10^{-2}$		Requires an auxiliary pump. Relative free of trouble.
Mechanical booster	$10\sim10^{-2}$	"	Requires an auxiliary pump. Low at operating cost.
Oil dispension pump	$10^{-3}\sim$		Requires an auxiliary pump. Operatin liquid should be maintained in a clea state.

Table 5.4-1 Main Corrosive Environments in Multistage Flash Desalination Plant

each part of plant		corrosion environment
brine *heater*	inside of heat transfer tubes, tube plate, water box	deaerated brine* at 116°C ∿ 121°C
	Inside of vessel	heating steam at highest temperature of 127°C
heat recovery section	inside of heat transfer tubes, tube plate, water box	deaerated brine* at 34°C ∿ 116°C
	condenser chamber	flash steam at 36°C ∿ 118°C
	flash chamber, demister	deaerated brine at 39°C ∿ 121°C, (vapor and mist)
	inside of fresh water trough	deaerated fresh water at 37°C ∿ 119°C
heat rejection section	inside of heat transfer tubes, tube plante, water box	raw seawater at 25°C ∿ 34°C
	condenser chamber	*flash steam* at 31°C ∿ 36°C
	flash *chamber*, demister	deaerated brine at 34°C ∿ 39°C, (vapor and mist)
	inside of fresh water trough	deaerated fresh water at 31°C ∿ 36°C
vent condenser	inside of heat transfer tubes, tube plate, water box	raw seawater at 25°C
	inside of vessel	noncondensable gas and steam
decarbo-nator	(containing piping *before and after the* equipment)	seawater at pH 3 ∿ 4
piping for circulating brine		deaerated brine at 34°C ∿ 121°C
piping for seawater		seawater

* concentrative of brine inside tubes is about 1.7 times of seawater

Table 5.4-2 Heat Transfer Tubing Materials in 119 units of MSF Plant

(Total Surface Area of Surveyed Tubes: 985,800 m^3)

Material	Surface area of heat transfer tube (%)
Aluminum brass	53
90/10Cu-Ni	53
70/30Cu-Ni	1.8
70/30Cu-Ni+2%Fe,Mn	11.4
Titanium	0.8

Table 5.4-3 Testing result of copper alloy tubing in CDA-OSW desalting plant

Heat recovery section	43 ℃ ~ 99 ℃ DO 30~75 ppb	Instantaneous corrosion rate at 30 months	0.02 ~ 0.075 mm/y $715 > 613 > 706 > 687$
Outlet of brine heater	2.1 ~ 2.4 m/s 121 ℃ DO < 1 ppb	Average corrosion rate in 54 months	0.02 ~ 0.038 mm/y $715\text{mod.} > 687 > \substack{613\\706} > 715$
Vertical tube evaporator	104 ℃ DO < 1 ppb	Average corrosion rate in 29 months	0.0025 ~ 0.013 mm/y $613 > 715 > \substack{687\\706}$

* Aluminum bronze (Cu-7Al-3.5Sn-3.5Fe)

Table 5.5-1 Concrete shell development program

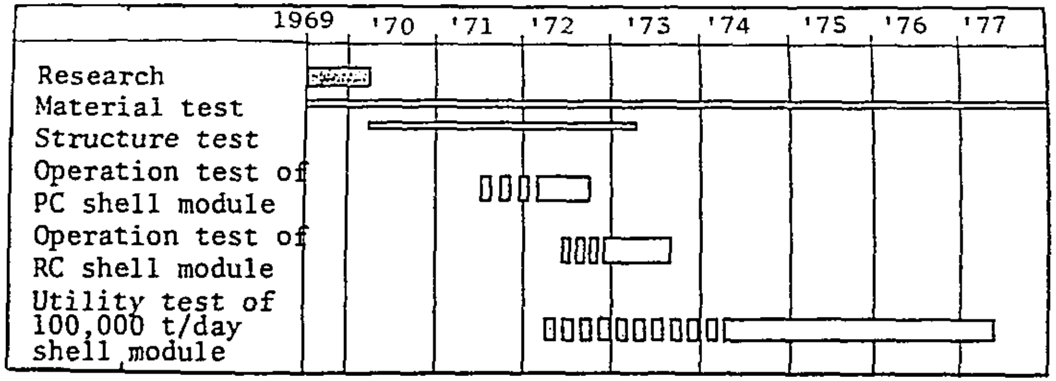


Table 5.5-2 Approximate design data of concrete shell modules

(1) Type:	Long tube MSF distillation type
(2) Structure:	Prestressed concrete shell (1971) Reinforced concrete shell (1972)
(3) Configuration	For heat recovery section Rectangular section, 4.3×4.5×15.0m (out side)
(4) Number of stages:	Heat recovery section (Prestressed or reinforced concrete): 5 stages *Heat rejection section (Steel): 2 stages*
(5) Evaporation rate:	300 t/d (Normal operation)
(6) Brine flow:	Approx. 500 t/h
(7) Temperature range of brine:	100 - 40 °C
(8) Concentration ratio of brine:	1.40 - 1.72

Table 5.5-3 Approximate design data of test module

(1) Type:	Long tube MSF distillation type
(2) Capacity:	9,500 m^3/day (High temperature operation)
(3) Number of stages:	Heat recovery section (Steel shell): 12 stages
	Heat rejection section (Reinforced concrete shell) : 3 stages
(4) Max. temp. of circulating brine:	121°C
(5) Concentration ratio of brine:	1.72 (60,000 ppm)

Table 5.5-4 Numerical values of constants in eq. 1

Temp. of condenced water (℃)	60		80		92	
Specimen	FC	PC	FC	PC	FC	PC
n	0.71	0.68	0.71	0.60	0.71	0.70
α	—	—	0.046	0.042	—	—
K_0	—	—	0.0087	0.021	—	—
K_0	0.0057	0.041			0.0092	0.0135

Items of analysis	Specimen Seawater, brine	 Scale	Analytical method	Reference
Qualitative analysis			Emission spectral analysis	
Ignition residue			Gravimetric analysis (1000°C, 1 hr) Differential heat gravimetric analysis	
Determination of structure			X-ray diffraction	
SO_4^{2-}			Gravimetric analysis, nephelometry ($BaCl_2$)	JIS K0101-31 (1966)
SiO_2			Colorimetric as molybdenum blue and yellow, gravimetric analysis	JIS K0101-33 (1966)
Magnesium			Atomic absorption spectrophotometry, titrimetric with EDTA (EBT)	JIS K0101-39 (1966)
Calcium			Atomic absorption spectrophotometry, titrimetric with EDTA (NN)	JIS K0101-38 (1966)
Fe			Atomic absorption spectrophotometry, colorimetric method (TPTZ)	JIS B8224-3243 (1969)
Cu			Atomic absorption spectrophotometry, colorimetric method (diethyl-dithiocarbamic acid soda)	JIS K0101-40 (1966)
Aluminium			Atomic absorption spectrophotometry (N_2O^-acetylene burner)	
Titanium			Colorimetric method (sulfosalicylic acid)	JIS H1662-2 (1966)
Cl^-			Argentometry (uranyl, potential difference)	JIS K0101-253 (1966)
COD			Alkaline $KMnO_4$ oxidation method	

Table 5.7-1 Generating power and steam cost in dual-purpose plant combining condensing type with extraction turbine

Utilization rate (%)	100	75	50	25	0
Utilization rate of power plant	310 7,440 hrs	70 % 6,132 hrs	70 % 6,132 hrs	70 % 6,132 hrs	70 % 6,132 hrs
Total electric power (10^6kWh/year)	7,112.6	5,874	5,893	5,911	5,930
Fuel comsumption (10^6kℓ/year)	1.7329	1.4211	1.3968	1.3726	1.3484
Annual fuel cost (100 million yen)	519.87	426.33	419.04	411.78	404.52
Construction costs 1 unit (100 million yen)	656 + 207 (housing, base, intake and discharge facilities) 863				
Annual fixed cost (100 million yen)	(863 × 0.1439) 124.2				
Annual expense (100 million yen)	644.07	550.53	543.24	535.98	528.72
Generating power cost (yen/kWh)	8.92	8.92	8.92	8.92	8.92
Steam cost (yen/ton)	422.8	1,556	1,545	1,234	—
Steam feed (10^6t/	2.277	1.707	1.138	0.569	—

Note: 1. Fuel heating value: 9,800 Kcal/ℓ, Unit cost: 30 yen/ℓ.
2. Annual interest 14.39% is adopted, based on 8% interest and 15-year-depreciation.
3. Steam cost is estimated by power credit method, based on generating power cost at 0% desalting.

Table 5.7-2 Generating power and steam cost in dual-purpose plant combining condensing type with back pressure turbine

Utilization rate (%)	100	75	50	25	0
Utilization rate of power plant	310 7,440 hrs	70 % 6,132 hrs	70 % 6,132 hrs	70 % 6,132 hrs	70 % 6,132 hrs
Total electric power (10^6kWh/year)	5,282.4	4,390.2	4,512.9	4,635.7	4,758.4
Fuel comsumption (10^6kℓ/year)	1.6932	1.3744	1.2965	1.2318	1.1605
Annual fuel cost (100 million yen)	507.96	412.32	388.95	369.54	348.5
Construction costs 1 unit (100 million yen)	667 + 179.5 (housing, base intake and discharge facilities) (846.5)				
Annual fixed cost (100 million yen)	(846.5 × 0.1439) 121.81				
Annual expense (100 million yen)	629.77	534.13	510.76	491.35	470.31
Generating power cost (yen/kWh)	9.884	9.884	9.884	9.884	9.884
Steam cost (yen/ton)	972.6	1,207.4	1,168.1	1,197.3	—
Steam feed (10^6t/year)	11.07	8.3	5.54	2.77	—

Note: 1. Fuel heating value: 9,800 Kcal/ℓ, Unit cost: 30 yen/ℓ.
2. Annual interest 14.39% is adopted, based on 8% interest and 15-year-depreciation.
3. Steam cost is estimated by power credit method, based on generating power cost at 0% desalting.

Table 5.7-3 Water and power cost for dual-purpose plant

Method	Water cost (cent/gallon)	Power cost (mil/KWH)
Power Credit (d)	25.9	4.90
Power Credit (f)	27.2	4.60
Prorating (b)	27.4	4.58
Power Credit (e)	27.6	4.50
Prorating (c)	27.9	4.43
Prorating (a)	30.7	3.76
LP Steam-Cost (e)	31.5	3.58
Water Credit (h)	35.0	2.71
Water Credit (g)	40.0	1.52

Table 5.7-4 Calculated result of steam cost

Distribution method	P.D.M.	Power credit method	
Performance ratio	14	12	14
Steam cost yen/ton	2031	1093	1108
	2354	1252	1266

P.D.M.: Proportional distribution method

Note: Upper terms in steam cost: Fuel cost 30 yen/ℓ
Lower terms in steam cost: 35 yen/ℓ

Table 5.7-5 Amount and Level of Waste Heat

	Exhaust heat (Kcal/Y)	Level	
Steel manufacturing	$6x10^{14}$	Above 200°C;	Loss-45%, high level; dispersion
Thermal power plant	$3.48x10^{14}$	Temperature rise of 8°C;	condenser
Nuclear power plant	$1.5x10^{13}$	Temperature rise of 8°C;	condenser
Cement manufacturing	$1.65x10^{13}$	160 ∿ 300°C;	waste gas
Glass manufacturing	$2.63x10^{10}$	200 ∿ 250°C;	glass tank kiln, waste gas
Refractory manufacturing	$1.3x10^{12}$	200 ∿ 300°C	
Ceramic manufacturing	$2.5x10^{12}$	Around 300°C;	tunnel kiln, square kiln
Aluminium smelting	$9.1x10^{12}$	200 ∿ 800°C;	combustion exhaus gas and air
Copper smelting	$7.0x10^{9}$	Average 500°C;	sensible heat of exhaust gas
Zinc smelting	$5.8x10^{10}$	Average 300°C	
Incinarator	$1.85x10^{14}$		Industrial waste
cf	Required colory		
Desalination, 100,000 t/day	$1.62x10^{12}$	(PR = $10x10^{4}$(t/D) x 10^{3}(kg/t) x 540(Kcal/kg) x 300(d/y)	

Table 5.7-6 Trial Design of a Dasalting Plant Using Waste Heat

	Case 1	Case 2	Case 3
1) Outline of specification	Hot-roll will waste thermal water	Coke oven gas cooler	Low temperature waste gas
Amount of water production (t/D)	1,413	14,064	6,816
Amount of waste thermal water (t/h)	2,000	2,860	1,386
Temperature of waste thermal water (°C)	43	67	67
Amount of waste cooling water (t/day)	4,093	17,559	8,059.6
Number of stages	6	18	18
2) Construction cost			
Total consturction cost	968,550	3,854,000	278,700
Unit construction cost (10,000 yen/TpD)	(68.5)	(27.4)	(40.5)
3) Water cost (yen/m^3)	458.5	157.1	209.5
4) Standard physical unit of energy (ℓ heavy oil/m^3 product water)			
Ejector steam	2.48	1.059	1.07
(ejector power)	0.38	0.188	0.189
Power consumption	1.189	0.287	0.291
Total	3.667	1.346	1.361
(Ejector power)	(1.569)	(0.475)	(0.480)

Table 5.7-7 Studies on Multi-purpose Use of Power Plant Energy

Studies related to multi-purpose use of heat energy produced by thermal power plants	Society for the Research on Multi-purpose Use of Nuclear Power Plants	Society for the Research on Nuclear Power Industrial Comples	Nuclear Power Multi-purpose Use Deliberation Committee
National Institute of Resources, Science and Technology Agency	Public Utilities Bureau, Ministry of International Trade and Ministry	Minister's Secretariat, Ministry of International Trade and Industry	Japan Atomic Industrial Forum
As a step to provide a measure for effective utilization of unused heat energy of thermal power plants, a system of transferring heat energy of thermal power plants for regional air-conditioning, industrial complex, desalination of seawater and agriculture will be studied from technical, economical and institutional points of view.	Multi-purpose use of energy of nuclear power plants for desalination and industrial process will be considered. Models will be used for this purpose and technical problems arising from model plants will be analyzed mainly from a viewpoint of power generation. Feasibility of multi-purpose use of energy will be studied from technical, economical and institutional points of view with economic calculations.	Two nuclear power plants scheduled to be completed in the last half of the 1980s will be used as energy sources for the utility center which supplies such utilities as electric power, process steam and fresh water from the sea to a large industrial complex to be established in the last half of the 1980s. Models of nuclear powered steel mills, chemical plants and smelters will be used.	Multi-purpose use of energy by combining a high temperature gas furnace and a fast breeder ATR with various utility consuming industries will be considered and models of actual site will be used by designating large industrial complex and Okinawa, for example. A combination of a reactor and MHD power generation is also considered.
• The last half of the 1980s is considered. • Models will be used for quantitative analysis of the savings of energy resources and economic benefits derived from the proposed power generating system as compared with the existing system, and studies will also be made as to the pollution control effect of the new system.	• Target is set for 1980. • A large light water reactor will be used. • Studies will be made mainly on technical aspects of multi-purpose use of power plant energy, such as equipment, operation and maintenance requirements and control of the plant.	• Last half of the 1970s and first half of the 1980s are considered. • Large light water reactor and high temperature gas furnace (in case of steel manufacturing). • Models will be used without designating specific sites.	• Year 2,000 is considered • FBR, HTGR, ATR • Models of specific locations will be used.

Table 5.8-1 Samples of Water Cost Calculated for the Five Desalting Processes

Cost elements		ED	RO	MSF	VC-VTE-MSF	VFVC
Capital costs	(10^3\$)	55,622	3,369	7,300	5,950	6,830
Plant and equipment	(")	1,700	2,275	6,300	5,500	6,350
Feedwater pretreatment	(")	400	-	n.r.	n.r.	n.r.
Feedwater supply	(")	250	280	1,000	450	480
Water treatmission	(")	212	259	-	-	
Brine disposal	(")	3,000	580	-	-	
Total capital costs	(c/kgal)	21.7	13.1	28.4	23.2	26.6
Operation and maintenance costs	(c/kgal)	9.4	12.9	19.2	11.9	10.1
Payroll extras, general and *administrative overhead*	(c/kgal)	2.0	2.0	6.7	4.6	
Maintenance materials	(c/kgal)	1.0	1.4	1.9	3.3	3.6
Membrane assembly or replacement turbing	(c/kgal)	5.3	10.0	3.1	1.7	1.9
Chemicals	(c/kgal)	3.0	3.0	3.0	3.0	n.r.
Fuel or steam	(c/kgal)	n.r.	n.r.	21.3	11.3	n.r.
Electric power	(c/kgal)	14.7	13.2	9.0	0.9	41.7
Total water cost	(c/kgal)	57.1	55.6	92.3	59.9	83.9
Plant capacity	(MGD)	5.0	5.0	5.0	5.0	5.0
Total dissolved solids	(TDS ppm)	2,477	3,610	35,000	35,000	35,000
Interest rate	(%)	4.7/8	4.7/8	4.7/8	4.7/8	4.7/8

ED;	electrodialysis
RO:	reverse osmosis
MSF;	multistage flash distillation
VC-VTE-MSF;	vapor compression-vertical tube evaporator-multistage flash distillation
VFVC;	vacuum-freezing vapor-compression

Table 5.8-2 Water Cost Calculation Standard for MSF Plants

(a) Capital cost calculation standard

Item	Calculation standard	Remarks
1. Plant and construction cost	Cost of main facilities and construction cost. This cost is given as C_1 (yen). The value for C_1 changes according to design specification and is not fixed.	
2. Plant intake and pump station capital cost	This cost C_2 is given as follows; $C_2 = 8{,}100 \times Fw$ (yen)	Fw represents the plant capacity (m^3/day). Length of intake tube is assumed to be 500 m. The factor indicates construction cost per unit of capacity (m^3/day) and is expressed by [yen/(m^3/day)].
3. Indirect capital cost	This cost C_3 is given as follows; $C_3 = 0.14 \times C_1$ (yen)	
4. Total capital cost	$C_T = C_1 + C_2 + C_3$ (yen)	

(b) Standard annual operation cost calculation

Item	Calculation standard	Remarks
1. Steam cost	Steam cost V_1 per ton for a dual-purpose plant of desalination and power generation is given as follows; $V_1 = 32 \times Fc + 143$ (yen/steam ton) in the above, $25 \leq Fc \leq 35$	 Fc; Unit price of heavy oil (yen/ℓ)
2. Electric power cost	Electric power P required for operation of the entire system of a desalting plant is given as follows; $P = 2.5$ (kWh/product water m^3) Unit price of power V_2 is given as follows; $V_2 = 12$ (yen/kWh)	 Contract demand; 5,200 kW Supply; 14,000 V Unit price of energy; 1,190 yen/kW 10.14 yen/kWh
3. Maintenance material cost	Maintenance material cost V_3 is given as follows; $V_3 = 0.005 \times C_T$ (yen/year)	
4. Operating personnel cost	Operating personnel cost V_4 is given as follows; $V_4 = 174$ million (yen/year)	Personnel cost; 4,000,000 yen/year·pers. No. of operators; 43.5
5. Maintenance personnel cost	Maintenance personnel cost V_5 is given as follows; $V_5 = V_4$ (yen/year)	
6. Overhead	Overhead V_6 is given as follows; $V_6 = 0.3\ (V_4 + V_5)$ (yen/year)	

7. Chemical cost	Chemical cost V_7 is given as follows; $V_7 = 2$ (yen/product water m^3)	Unit price of concentrated sulphric acid; 9 yen/kg (coast-wise vessels) Unit price of deforming agent; 2,500 yen/kg
8. Rate of interest and amortization	The coefficient is given as 0.1. The annual capital expenditure V_8 is given as follows; Note: Annual capital expenditure V_8 is calculated with the following equation. ($V_8 = 0.1 \times C_T$/No. of working days x Fw) (yen/product water m^3)	50% by government bonds (7.5% interest) and 50% by public subscribed bonds (8.5% interest). Useful life; 20 years Residual value; 10%
9. Insurance cost	Insurance cost V_9 is given as follows; $V_9 = 0.02 \times C_T$ (yen/year)	

Table 5.8-3 Water cost for MSF plant (100,000 m^3/day)

Unit: yen/m^3

Cost elements		Fresh water cost	
Plant cost		41.93	41.93
Maintenance cost	Materials	2.04	
	Labor	2.68	4.72
Operation cost	Steam	101.29	
	Electric power	12.10	
	Chemicals	3.61	
	Administrative	5.48	
	Total water cost	1.61	124.09
Total			170.74

Table 5.9-1 Result of ammonium chloride additive test (partition of ammonium)

Exp. No	NH_4^+ concentration				pH		Concentration rate of ammonium
	Make-up seawater (A)	Circulating brine	Discharged brine	Product water (B)	Circulating brine	Discharged brine	
1	0.134	0.074	0.058	0.235	7.10	7.18	1.75
2	0.460	0.270	0.198	0.714	7.19	7.27	1.55
3	0.876	0.604	0.485	1.39	7.12	7.20	1.69
4	1.29	1.01	0.763	1.97	7.01	7.01	1.53
5	2.52	2.74	2.88	2.04	6.78	6.81	0.81

Table 5.9-2 Comparison between calculated value and measured value of phenol in MSF plant

	Make-up seawater	Product water		Circulating brine		Discharged brine	
		Measured	Calculated	Measured	Calculated	Measured	Calculated
Phenol	0.400ppm	0.602	0.585	0.280	0.298	0.210	0.253
O-cresol	0.310	0.600	0.618	0.143	0.146	0.070	0.072
m-cresol	0.550	0.950	0.973	0.320	0.323	0.220	0.221

Table 5.9-3 Water quality of evaporation plant in Key west aqueduct commission

Item	Value	Item	Value
pH	6.6	Cl^-	71 ppm
temperature	30 ℃	NO_3^-	0 〃
turbidity	0	Fe	0.08 〃
MO alkalinity	8	Mn	0.008 〃
total hardness	42	SiO_2	0.1 〃
electric conductivity	175 µ℧	F^-	0 〃
Ca	2.0ppm	B	0.01 〃
Mg	3.7 〃	Cu	0.04 〃
Na+K	36.2 〃	Cr(VI)	
CO_3^{2-}	9.7 〃	CN	
SO_4^{2-}	6 〃	PO_4^{3-}	<0.05 〃

Table 5.9-4 Water quality of 3,000 m^3/day test plant at Chigasaki test facility for national R & D program, MITI

Item	make-up seawater	product water
pH	8.3	7.0 ppm
temperature	20.9 ℃	31.2 ℃
turbidity	2.12 ppm	0
COD	2.71 〃	trace
Ca (as $CaCO_3$)	885 〃	〃
Mg (〃 〃)	5479 〃	〃
Fe	0.19 〃	0.1 ppm
SiO_2	0.41 〃	0.02 〃
Cl^-	18.78 ‰	trace
NH_4^+-N	0.05 ppm	0.01 ppm
NO_2^--N	3.7 γ/ℓ	0.1 γ/ℓ

Table 5.9-5 Water quality of evaporation plant at Shevchenko (Caspian coast), U. S. S. R.

Item	Value	Item	Value
temp.	50 ~ 60 ℃	Ca	1.0 ~ 2.5 ppm
alkalinity	0.10 meq/ℓ	Mg	1.5 ~ 3.0 〃
hardness	0.20 meq/ℓ	Na+K	2.0 ~ 4.0 〃
pH	6.5 ~ 7.5	Fe	0.2 ~ 0.3 〃
TDS	5 ~ 10 ppm	CO_2	2.8 〃
Cl^-	0.2 ~ 0.5 〃	dissolved oxygen	0.3 ~ 0.6 〃
SO_4^{2-}	1.0 ~ 2.5 〃		

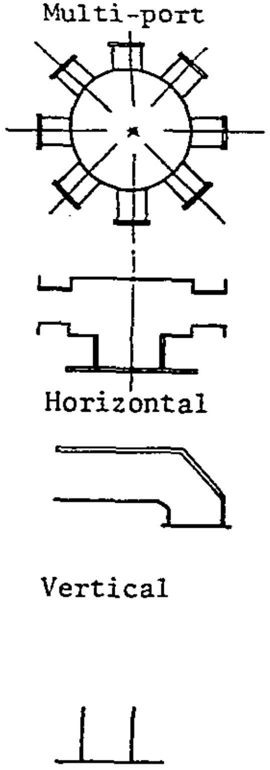
Investigation item		1st	2nd	3rd	4th	5th	Type of nozzle
Period		Aug.6, 1975 ∿Aug.8,1975	Dec. 8, 1975 ∿Dec.18,1975	Feb.22, 1976 ∿Feb.26,1976	Jun. 1, 1976 ∿Jun.10,1976	Nov. 5, 1976 ∿Nov.13,1976	
Area		Whole area of Ozu anchorage	Neighborhood of release-port	Neighborhood of release-port	Neighborhood of release-port	Neighborhood of release-port	
Item	Temperature (vertical)	6 points	17 points	12 points	18 points	16 points	
	Temperature (horizontal)		4 serieses	4 serieses	6 serieses	5 serieses	
	Chlorinity	6 points	17 points	12 points	18 points	16 points	
	Flow rate	6 points	2 points	1 point	7 points	5 points	
Diffuser nozzle			Multi-port nozzle	Multi-port nozzle	Multi-port nozzle Vertical nozzle Horizontal nozzle	Multi-port nozzle Vertical nozzle Horizontal nozzle	
Condition of intake/release: Intake	Flow rate Q (m^3/hr)	No intake/release	3620 ∿ 4830	4470	4800 ∿ 5810	3940 ∿ 4030	
	Temp. (°C)		16.0 ∿ 17.9	10.2	15.5 ∿ 17.0	19.9 ∿ 20.2	
	Chlorinity Cl (‰)		18.1	18.25	17.75	17.75	
Release	Flow rate Q (m^3/hr)		3660 ∿ 4870	4310 ∿ 4330	4630 ∿ 5650	4010 ∿ 4040	
	Temp. (°C)		26.1 ∿ 27.9	20.8	24.4	24.8 ∿ 32.5	
	Chlorinity Cl (‰)		17.8 ∿ 18.9	18.4	18.2	17.7	

Table 5.9-7 Conditions of numerical simulation

Flow rate Q (m^3/s)	Salinity difference ΔS ($^0/_{00}$)	Temp. difference ΔT (°C)	Density difference Δρ	Inner froude number at the discharged outlet Fo	H(m) Depth of the discharged outlet	D(m) Diameter of the discharge pipe	H/D
5.96	6.36	7.0	-2.7×10^3	3.0	11.7	3.0	3.9