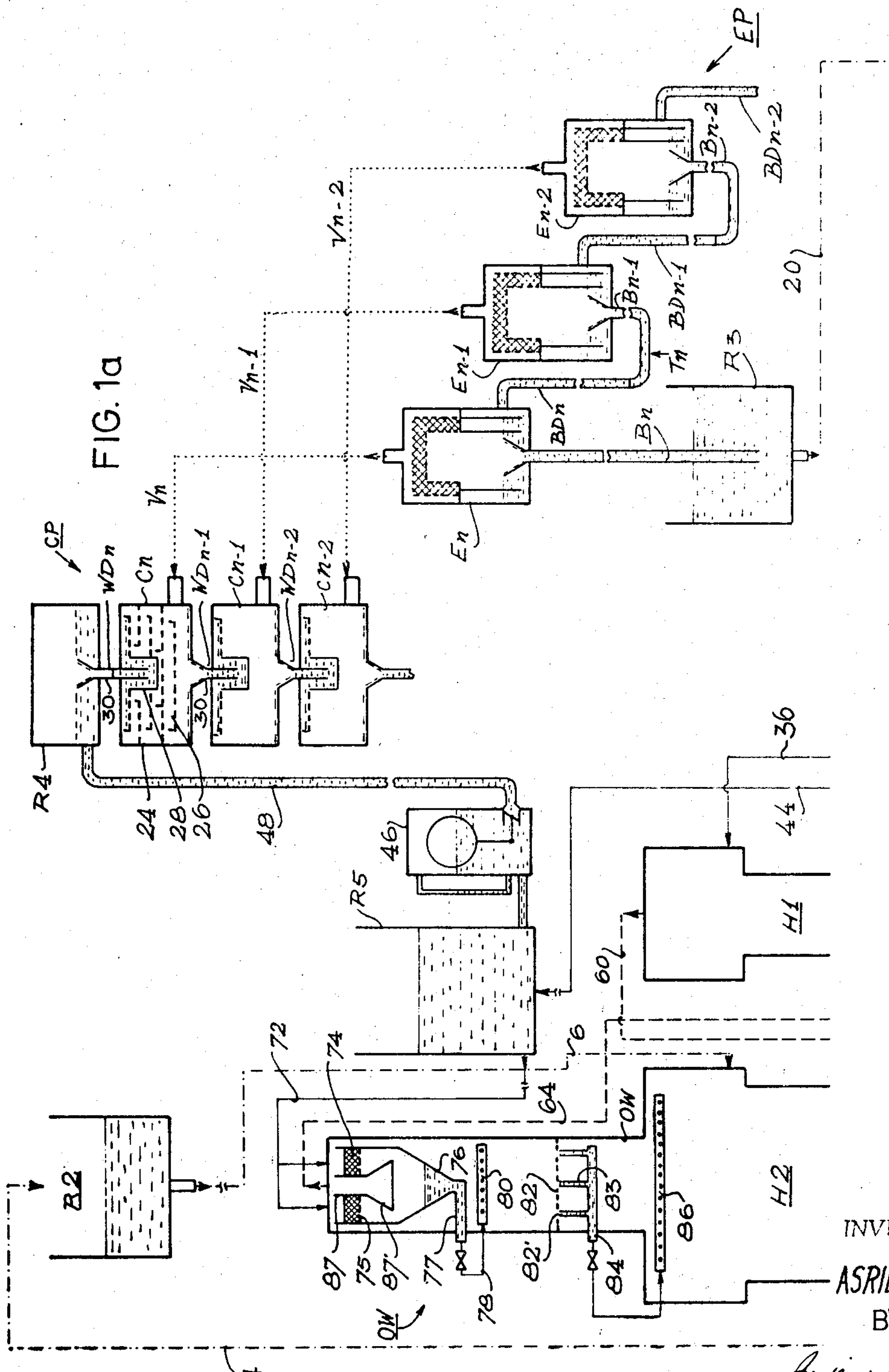

FIG. 1a

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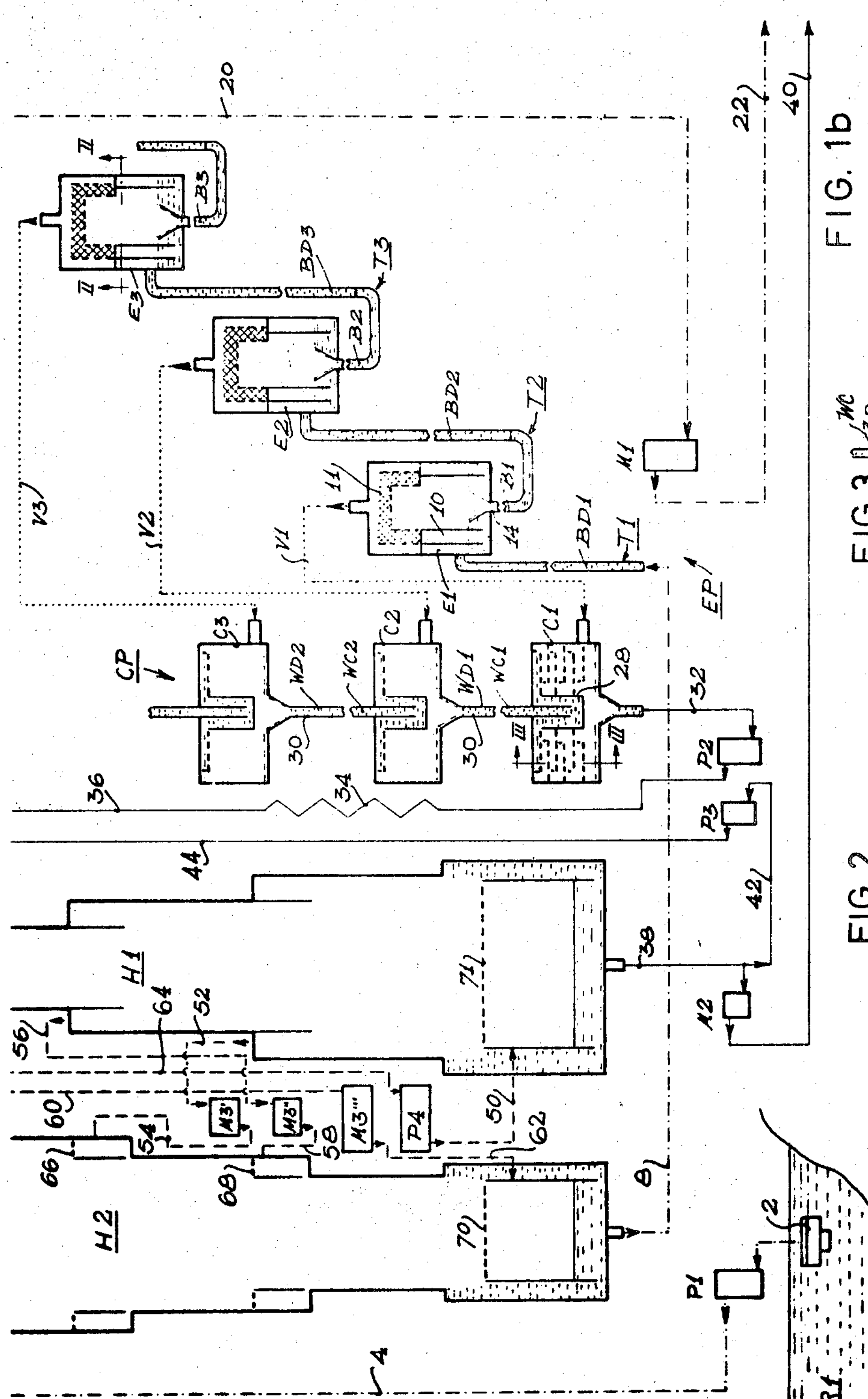

FIG. 1b

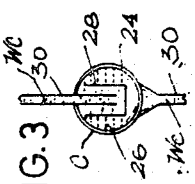

FIG. 3

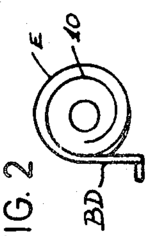

FIG. 2

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FIG. 4

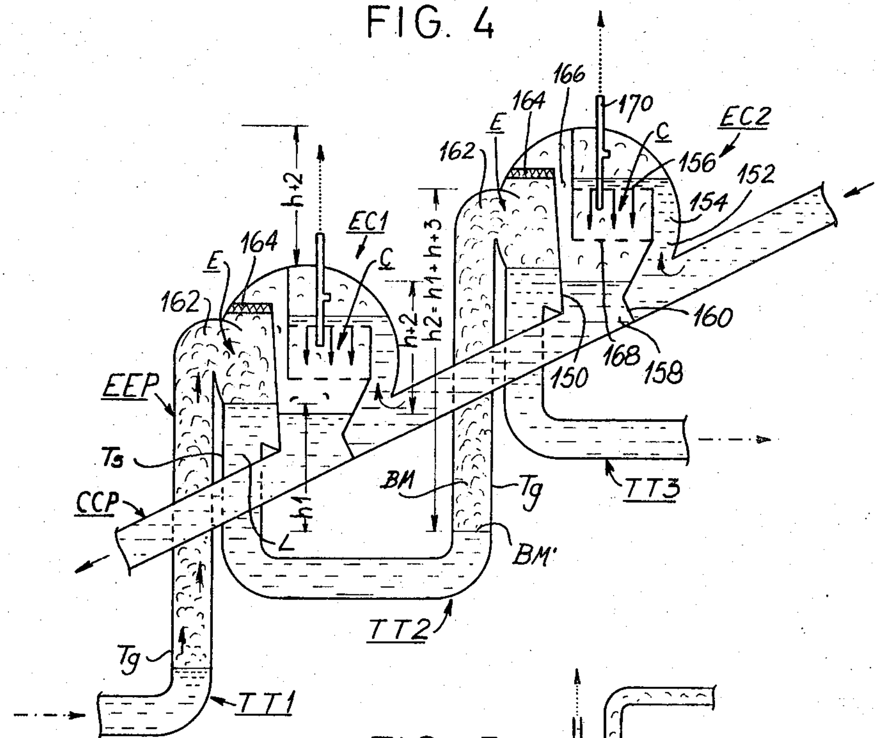
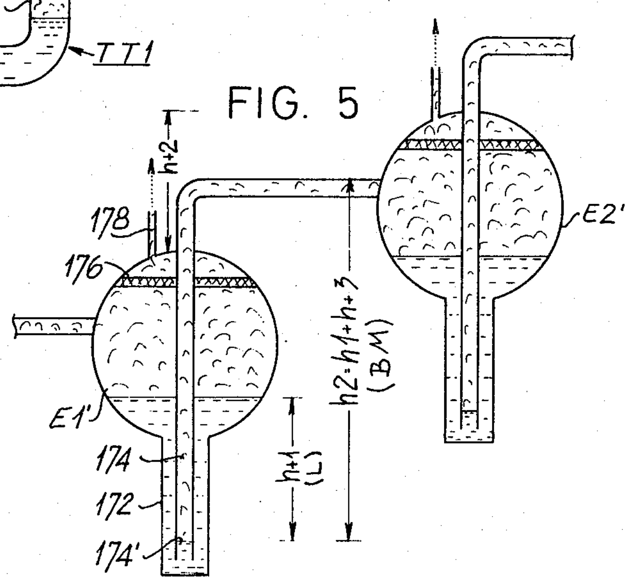


FIG. 5



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3,647,638

ASCENDING MULTI-STAGE DISTILLATION APPARATUS AND METHOD UTILIZING A FEED-LIQUID-LIFT SYSTEM

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Int. Cl. B01d *3/06, 1/14*
U.S. Cl. 203—11 7 Claims

ABSTRACT OF THE DISCLOSURE

A multi-stage distillation system, particularly useful for desalination of sea water, is described comprising a plurality of evaporators and condensers disposed at elevations increasing from the first stage of highest temperature and pressure to the last stage of lowest temperature and pressure, the feed liquid being moved upwardly against gravity through the evaporators by the energy of vaporization and difference in pressure between stages, and the cooling liquid moving downwardly by gravity through the condensers against increasing pressure. A feed-liquid-lift system is described including a down-flowing liquid column of the feed liquid continuously driving a longer up-flowing boiling-mixture column of the feed liquid mixed with vapor, while liquid at the lower end of the up-flowing boiling-mixture column continuously effect a hydraulic seal between adjacent stages.

BACKGROUND OF THE INVENTION

Field of the invention

The present invention relates to a multi-stage distillation system. It is particularly, but not exclusively, intended for the desalination of saline water in accordance with the known vapor-reheat flash evaporation process, and is therefore described below with respect to such process.

Description of the prior art

In the known vapor-reheat flash evaporation process, the vapors from the evaporation (or flash) chambers are brought into direct contact with distilled water circulating through the condensing chambers. The circulating distilled water is first introduced cold into the last-stage condensing chamber, and its temperature rises as it circulates from each stage to the next operating at higher temperature and pressure, its temperature and quantity reaching a maximum in the first stage condensing chamber. The distillate is then passed through a heat-exchanger where it heats by direct contact an intermediary fluid, e.g., an immiscible oil. The hot oil is circulated to a second heat-exchanger where it heats the incoming cold saline water also by direct contact. A part of the cooled distilled water is withdrawn as product, and the balance is recycled to act again as a condensing medium for the vapor from the evaporating chambers. Further details of the known vapor-reheat flash evaporation process are available in the literature.

One of the big drawbacks in this process is the need for a pump at each stage for pumping the distillate from one condensation chamber to the next, in which the temperature and the vapor pressure are higher. The requirement for such a pump at each stage increases the initial investment and the total power requirements, and as a practical matter limits the number of stages that may be included.

A brine self-lift system has been proposed as described on pages 36–1 to 36–12 of the report on the second European Symposium on Fresh Water from the Sea, Athens, May 9–12, 1967, therein referred to as the "Clementine" system. According to this system, a riser (a vertical pipe increasing in cross-section from bottom to top) connects two vertically-disposed evaporator stages. The open bottom of the riser is submerged a little below the brine surface in the lower evaporator stage, and its open top is projected a little above the brine surface in the upper stage. In this system, however, it is necessary that the riser increase in cross-section from bottom to top to avoid acceleration of the steam/brine mixture as it flashes (pages 36–3 of the above reference). Also, it is not possible to obtain substantial lift particularly at low temperature stages. These drawbacks seriously limit the applicability of the Clementine system, particularly with respect to large-volume, low cost desalination of sea water.

OBJECTS OF THE PRESENT INVENTION

An object of the present invention is to provide improvements to multi-stage distillation methods and systems of the foregoing type which, among other advantages, obviate the need for a pump at each stage.

A further object of the invention is to provide an improved flashing and self-lift method and system for the feed liquid, (e.g., sea water or brine) which utilizes the energy of evaporation of the flashing feed liquid in a wide temperature range (e.g., 20° C. to 180° C. or higher), enabling a lift of the feed liquid from stage to stage to a height considerably exceeding the obtainable by the Clementine system.

SUMMARY OF THE PRESENT INVENTION

According to one aspect of the invention, there is provided a multi-stage distillation apparatus for distilling feed liquid including a plurality of evaporators at different temperatures and vapor pressures both decreasing from one stage to the next, characterized in that said evaporators are disposed at increasing elevations with the highest temperature and vapor pressure evaporator at the lowest elevation, and that said evaporators are included in an evaporator conduit system utilizing the external energy of evaporation of the feed liquid to drive the feed liquid upwardly against gravity through the evaporators in succession, said evaporator conduit system including a communicating path connecting together each pair of adjacent evaporators in which communicating path there are formed and maintained a down-flowing liquid column of the feed liquid flowing out of one evaporator and a longer up-flowing boiling-mixture column of the feed liquid and vapor flowing into the adjacent higher elevation evaporator, there being liquid at the bottom of said up-flowing boiling-mixture column effecting a hydraulic seal between the adjacent evaporators.

From another point of view, the invention provides, in a multi-stage distillation apparatus for distilling feed liquid, an evaporator conduit system for effecting self-lift and flashing of the feed liquid, which evaporator conduit system is in the form of an ascending sinuous conduit having at the peak of each sinus a liquid-vapor separator, the pressure and temperature in the separators decreasing from bottom to top of the ascending conduit, each sinus of the ascending conduit including a down-flowing liquid column of the feed liquid and a longer up-flowing boiling-mixture column of the feed liquid mixed with vapor, the feed liquid in the valley of each sinus being in liquid form and effecting a hydraulic seal between the sinuses.

This feature of the invention uses the boiling phenomenon and the energy of vaporization to cause the feed liquid (e.g., saline water or brine) to rise and to flow from one stage to the next of lower pressure but of higher elevation. The difference in the vapor pressure within two consecutive evaporators forms and maintains a flowing boiling-mixture column of feed liquid and vapor within the connecting duct (sometimes called a "boiling duct"), the latter having a suitable form and cross-sectional area to produce the desired height of boiling column. The pressure differences between two consecutive stages are approximately equal for the evaporating and condensing chambers. However, the density of the water in the water duct is greater than the density of the boling mixture in the boiling duct. Consequently, the height of the boiling column is greater than the height of the water column for the same stage at equilibrium conditions. As an example, in a 103-stage system (as described below), the equilibrium heights for three stages, i.e. the 1st, the 52nd and the last, may be as follows:

	Stage No.		
	1	52	103
Boiling column, m	5.7	2.8	3.0
Water column, m	3.8	0.55	0.024
Height differences, m	1.9	2.25	2.976

In practice the difference in elevation of two consecutive evaporating chambers could be equal to the height of the water column of the same stage plus at least two meters.

Two forms of an evaporator conduit system for effecting self-lift and flashing of the feed liquid are herein described, one using U-tubes and the other based on the Well-type manometer system.

BRIEF DESCRIPTION OF THE DRAWINGS

The invention is herein described with reference to the accompanying drawings which illustrate, somewhat diagrammatically and by way of example only, several distillation systems constructed in accordance with the invention for use in the known vapor-reheat flash evaporation process. In the drawings:

FIGS. 1*a* and 1*b*, taken together diagrammatically illustrate one such system;

FIG. 2 is a section along lines II—II of FIG. 1*b*, illustrating the construction of one of the evaporating chambers; and

FIG. 3 is a section along lines III—III of FIG. 1*b*, illustrating the construction of one of the condensing chambers.

FIG. 4 illustrates a double-compartment unit usable for each stage in the apparatus of FIG. 1, one compartment serving as the evaporator and the other compartment serving as the condenser;

FIG. 5 illustrates a variation in the self-lift evaporator conduit system that may be used.

In the drawings, the paths of the several media flowing through the system are indicated as follows: saline water and brine, by a line of alternate dot and dashes; fresh water, by a continuous line; water vapor, by a line of dots; the intermediary immiscible fluid (e.g. a liquid hydrocarbon, hereinafter called "oil"), by a line of dashes.

The invention basically relates to distillation and could be used in a number of applications involving the distillation of liquids. The preferred application of the invention, however, relates to the desalination of sea water to produce pure water and/or salts, and the invention is therefore described below with respect to that application, but it will be appreciated that it is not limited to such application.

DESCRIPTION OF THE EMBODIMENT OF FIGS. 1–3

This system includes a plurality of flash chambers (hereinafter called "evaporating chambers" or "evaporators") E1, E2 . . . E*n*; a plurality of condensing chambers or condensers C1, C2 . . . C*n*; a first heat-exchanger H1 in which hot distilled water is used to heat the oil; and a second heat-exchanger H2 in which the hot oil is used to heat the incoming saline water. The condensers are disposed at increasing elevations, the first-stage condenser C1 of highest pressure and temperature being at the lowest elevation, and the last-stage condenser C*n* of lowest pressure and temperature being at the highest elevation. Similarly the evaporators are also disposed at increasing elevations, the first-stage evaporator E1 of highest temperature and pressure being disposed at the lowest elevation, and the last-stage evaporator E*n* of the lowest temperature and pressure being at the highest elevation. The two heat-exchangers H1 and H2 are of the direct-contact type. By way of example, there may be used the oil-droplet type in which up-flowing oil drops come into direct contact with down-flowing saline water, in one heat-exchanger, and with fresh water, in the other heat-exchanger, as disclosed on pages 58 and 59 of the above-cited Spiegler book.

All the evaporators are included in an evaporator conduit system, generally designated EP, in which the evaporators are connected in series by U-tubes T1, T2 . . . T*n*. Each U-tube includes a short leg in which there is formed and maintained a down-flowing liquid column of the saline water or brine flowing out of each evaporator (out of the heat-exchanger H1 in the first stage) driving a longer up-flowing boiling mixture column of the brine and vapor into the next higher elevation evaporator. As is clearly shown in FIGS. 1*a* and 1*b*, the cross-sectional areas of the communicating path formed by the down-flowing and up-flowing colums are substantially the same. It will also be noted from the drawings that the communicating path formed by these columns is continuously opened and unrestricted throughout its length. FIGS. 1*a* and 1*b* also shown (as do FIGS. 4 and 5) that while the down-flowing column is made up of liquid in the form of saline water, the longer, up-flowing column includes a lower portion also of liquid in the form of saline water and an upper portion of a boiling mixture of liquid and vapor. As previously mentioned, there is thus provided a hydraulic seal at the bottom of each column. With this arrangement, the high density down-flowing liquid column drives the lower density boiling mixture in the longer up-flowing column, thereby achieving a substantial lift to a higher elevation evaporator stage. All the condensers are included in a condenser conduit system, generally designated CP, including water ducts WD1, WD2, . . . WD*n*, connecting the condensers in series.

The saline water is drawn from a reservoir R1 and is pumped by pump P1 through filter **2** and pipe **4** to an open reservoir R2 at high elevation. The saline water then flows through pipe **6** into the upper end of heat-exchanger H2 where it is heated by up-flowing hot oil and exists at the bottom. It then passes through pipe **8** (constituting the short leg of the first U-tube T1), to the vertical conduit forming the boiling duct BD1, (constituting the long leg of U-tube T1) for the first-stage evaporator E1.

The first-stage evaporator E1 is at an elevation above the point of exit of the hot saline water from heat-exchanger H2. However, the temperature and the vapor pressure within E1 are lower so that the saline water is under boiling conditions in boiling duct BD1. Thus, the boling mixture in the boiling duct BD1 is of less density than the saline water in pipe **8**. This difference in density, together with the difference in pressure between heat-exchanger H2 and evaporator E1, effects a lift of the boiling mixture through boiling duct BD1 to the higher elevation of evaporator E1.

Each of the evaporators is provided with a cyclone construction (FIG. 2) in cluding a spiral baffle **10** for facilitating the separation of the distillate vapor from the brine droplets by gravity and centrifugal action. The vapor passes through demister **11** and exits through the top of the chamber and is directed by pipe VI to the first-stage condenser C1, the brine exiting from the bottom of the evaporator through brine outlet duct B1 which is connected to the vertical boiling duct BD2 of the second-stage evaporator E2. The vapor exiting from the second evaporator E2, at lower temperature and pressure, is directed through pipe V2 to the second-stage condenser C2, while the remaining brine exits from evaporator E2 through outlet B2 joined to the boiling duct BD3 of the next-stage evaporator at higher elevation and at lower pressure and temperature. The process continues through the remaining stages of the evaporators, the saline water being in a boiling condition in the boiling duct between each pair of stages and thus rising to the elevation of each succeeding stage. The residual brine in the last-stage evaporator E*n* exits through pipe B*n* into a tank R_3, and then flows downwardly through pipe **20** and power generator M1 to the brine outlet **22**.

It will be appreciated that the brine in the outlet conduit (e.g. B1 or B2) of one evaporator and at the bottom of the boiling duct (e.g. BD2 or BD3) leading to the next evaporator forms a hydraulic seal between the two evapotors so that the vapor at the higher pressure within the one evaporator will not flow to the next evaporator at lower pressure.

The above-described evaporator conduit system for effecting self-lift and flashing of the brine could be considered as being in theform of an ascending sinuous conduit having at the peak of each sinus a liquid-vapor separator (i.e., the evaporators), the pressure and temperature in which separators decrease from bottom to top of the ascending conduit, each sinus of the ascending conduit, including a down-flowing liquid column (i.e. the brine flowing through pipe **8** and brine outlet ducts B1, B2), and a longer upflowing boiling-mixture column (i.e., the brine and vapor flowing through the boiling ducts BD1, BD2, etc.), the liquid (brine) in the valley of each sinus effecting a hydraulic seal between the sinuses.

The evaporator in each stage is maintained at a slightly greater pressure than that in the condenser of its respective stage, to promote the flow of the vapor from the evaporator to its condenser. As indicated above, the temperature and pressure in the evaporator and condenser of each stage decrease from one stage to the next, and therefore the distillate entering the last stage condenser C*n* from its respective evaporator E*n* is at the lowest temperature and pressure. The cooling water distillate driven into the last stage condenser C*n* passes by gravity through all the condensers to the first-stage condenser C1, picking up heat and water from the vapor leaving the evaporators and entering the condensers. A tank R4 is provided above the last-stage condenser C*n* and receives a portion of the liquid distillate (taken from heat-exchanger H1 as will be described below), which portion is circulated through the condensers.

Each condenser includes a chamber **24**, preferably cylindrical in shape (see FIG. 3), containing a plurality of perforated plates **26** disposed in different horizontal levels within the chamber, and provided with flanged ends, defining a sinuous path between the plates for the vapor. The vapor from the evaporator flows along this sinuous path between the perforated plates, while the liquid distillate flows by gravity from one plate to the next passing through the perforations in the plates and also spilling over the ends of the plates.

Each chamber **24** also includes a receptacle **28**, closed at the bottom and open at the top, and a duct **30** leading from the bottom of the chamber into the receptacle of the chamber next in line. The distillate from tank R4 flows through its duct **30** into the receptacle **28** of the last-stage condenser and overflows into the chamber of that condenser, picking up heat and condensing distillate from the vapor injected into the chamber. The liquid distillate then leaves condenser C*n* through its duct **30** and passes into the receptacle **28** of the next succeeding condenser (C*n*–1), overflows the receptacle in that condenser, and picks up heat and liquid distillate from the condensing vapor from the evaporator of that stage (E*n*–1), the process being continued throughout the remaining stages of the condenser.

Since the temperature and the vapor pressure increase from condenser C*n* to condenser C1, each of the ducts **30** between the stages will contain a liquid column. The height of each liquid column is determined by the difference in pressure between the two condensers, and thus will provide a hydraulic seal between the condensers to prevent the vapor from the chamber at the higher pressure to flow to the chamber at the lower pressure. The distillate thus circulates from one stage to the next by gravity, there being no need for a separate pump between each stage as required in the previously known systems.

The liquid distillate exits from the condenser stages through pipe **32** at the bottom of the first-stage condenser C1, and now is at a very high temperature and pressure. This distillate is pumped by pump P2 through a make-up heater **34**, where its temperature is raised further, and is then introduced through pipe **36** into the upper end of heat-exchanger H1. It flows down through this heat-exchanger, heating the up-flowing oil, and exits through pipe **38**. Part of the liquid distillate, now cold but at a high pressure, is directed through power generator M2 to the desalinated water outlet **40**. Another part is directed through pipe **42** to pump P3 from where it is pumped through pipe **44** up to a tank R5 placed at higher elevation. The latter tank is open and is therefore at atmospheric pressure. Tank R4 is maintained at a very low pressure, lower than the pressure of the last-stage condenser C*n*, and therefore the distillate will flow from tank R5 to tank R4 by the pressure difference between the two tanks, the flow being through a regulator **46** (of known construction) and a vertical duct **48**. From tank R4 the distillate is again circulated through the condenser stages, picking up heat and condensing distillate as described above.

In heat-exchanger H1, cold oil introduced at the lower end of the heat-exchanger, through pipe **50** and pump P4, is heated by the down-flowing liquid distillate. The hot oil exists from heat-exchanger H1 (as to be described below) and is introduced into heat-exchanger H2 where it is used for heating the incoming saline water.

Because the specific heat of the oil changes with a change in temperature, three sub-cycles of oil flow between heat-exchangers H1 and H2 are provided so as to equalize the heat capacities within each of three temperature ranges. For this purpose also, each of the heat-exchangers H1 and H2 (which as indicated earlier may be of the vertical, direct-contact, droplet-system type described in the above-cited Spiegler book) has a different horizontal cross-section area for each one of the said oil cycles. Thus, in one oil sub-cycle, a part of the oil exits through outlet pipe **52** from the lower, wider-diameter end of heat-exchanger H1, passes through power generator M3′, and then is introduced through pipe **54** into the upper, wider-diameter end of the heat-exchanger H2. In a second oil cycle, another part of the oil exits through outlet pipe **56** from a higher, smaller-diameter portion of heat-exchanger H1, passes through power generator M3″, and is introduced through pipe **58** into a lower, smaller-diameter portion of heat-exchanger H2. In the third oil cycle, the bulk of the oil exits from the upper, smallest-diameter end of heat-exchanger H1 through outlet pipe **60**, passes through power generator M3‴, and is then introduced through pipe **62** into the lowest, smallest-diameter portion of heat-exchanger H2. All the oil exits cold from heat-exchanger H2 through pipe **64** to pump P4 and is reintroduced into the lower end of heat-exchanger H1 through pipe **50**.

The preforated plates **66**, **68** and **70** are provided in heat-exchanger H2 above the oil inlet pipes **54**, **58** and **62**, respectively, in order to cause the oil to disperse, or to form drops, immediately after it is introduced through these inlets. A similar perforated plate **71** is provided above the oil inlet pipe **50** in heat-exchanger H1. The oil will thus flow upwardly in both heat-exchangers in a dispersed or "drop" form providing a large surface area for heat transfer.

An oil washing column OW is provided at the upper end of heat-exchanger H2. A small quantity of (e.g., about 5%) of the product distilled water from tank R5 is introduced through pipe **72** into the top of the oil washing column OW. It passes through a filter **74** (e.g., of fiber glass) supported on a screen **75**, and all or part of it is collected by a funnel **76** having a collector **77** directing it to a distributor **80**. The column further includes a perforated plate **82**, having large openings **82'** in which are fitted pipes **83**. The water passes through openings **82'** into pipes **83**, from where it is directed to a collector **84** and a second distributor **86**.

The oil introduced into heat-exchanger H2, flows upwardly in a dispersed or drop condition, as described above, and coalesces above distributor **86**, but as it rises further, perforated plate **82** disperses it again. It again coalesces above distributor **80** and then flows through filter **74** to an outlet pipe **87** formed with an inverted funnel portion **87'** at its lower end, the latter being spaced above the lower end of funnel **76** sufficiently so as to be above the level of the water therein.

Oil washing column OW at the upper end of the heat-exchanger H2 washes the oil of entrained saline water droplets and substantially reduces the salinity of the water carried within it, in the following manner: First, the water from distributor **86** forms an upper layer above the incoming saline water (introduced through inlet pipe **6**) so that the up-flowing oil coalesces in the washing water layer rather than in the incoming saline water. Thus, the water entrained in the oil during the coalescing of the latter above distributor **86** will be considerably less saline than would have been the case had the oil coalesced in the incoming saline water. These entrained water droplets are carried with the oil to heat-exchanger H1, and being less saline, contaminate the distilled water there to a less degree than would have been the case had the oil been permitted to carry entrained droplets of the raw saline water. Incidentally, the entrained quantity of water is the form of these "diluted" droplets may be greater than would have been the case without the repeated washing operations; however, as the oil is heated in heat-exchanger H1, it dissolves a great part of these droplets and, accordingly, a smaller amount of pure distilled water from heat-exchanger H1 than would otherwise have been the case. The salinity of the water entrained in the oil exiting from heat-exchanger H2 is further reduced as the oil rises (in dispersed from) in the space between perforated plate **82** and distributor **80**, since it is washed by the water descending by gravity from the distributor through the space. The salinity of the entrained water is still further reduced by filter **74** wherein it is washed by the fresh water introduced into the filter through inlet pipe **72**.

Following is a description (the values being approximate) of a proposed design illustrating one embodiment of the invention for purposes of example only.

In this example, there would be 103 stages constructed vertically or on an incline (preferably on a slope). The temperature and pressure of the saline water (sea water) at the inlet to the first evaporator stage E1 (i.e. at the bottom of boiling duct BD1) would be 180° C. and 100.3 meters water, respectively. The temperature drop of the saline water after each evaporating stage, and the temperature rise of the fresh water at each condensing stage, would be 1.5° C. The temperature of the vapor flowing into the first condenser would be 178.5° C., and the temperature of the cooling water flowing out from the first condenser would be 177.5° C. In the last stage (No. 103), the difference between the temperature of the vapor flowing into the condenser and the temperature of the cooling distillate flowing out from the condenser would be 2° C. This temperature difference gradually decreases from the last to the first stage, the difference being 1° C. in the

first stage. The temperature of the residual brine flowing out of pipe B*n* following the last evaporator would be 24° C., and the temperature of the cooling fresh water flowing into the last condenser would be 20.5° C.

The vapor pressure of the fresh water flowing out of the first condenser C1 (at 177.5° C.) would be 9.65 kgs./cm.^{2} or 96.5 meters water. Assuming the useful height of each condensing chamber is 1.5 meters, and adding 50 meters for pressure drop in the water ducts and for flow regulation, the last condensing chamber C*n* (No. 103) would be at a height of approximately 300 meters above the first condensing chamber C1. The last evaporating chamber E*n* (No. 103) would be preferably also at a height of approximately 300 meters, assuming that the difference in elevation of two consecutive evaporators is equal to the height of the water column of the same stage plus two meters, as indicated earlier for the condensers. In this example, the saline water at the bottom of the first boiling column (BD1) would be 180° C. and 100.3 meters water pressure. The temperature and pressure would decrease to the last stage evaporator E*n* (No. 103) to about 25.5° C., 0.32 meters water pressure, respectively.

In the present example, about 1000 kgs. of desalinated water (at 20.5° C.) would be obtained through the distilled water outlet **40**, and about 3000 kgs. of residual brine (at 24° C.) would be obtained through the brine outlet **22** (the brine containing 4.7% salt) from 4000 kgs. of sea water (containing 3.5% salt) introduced into the plant at 17.5° C. About 3870 kgs. of distilled water (at 177.5° C.) would leave the first-stage condenser C1, and would be heated by fuel heat (conventional or nuclear) at heater **34** to 183° C. The enthalpies of pure water at 177.5° C. and 183° C. are 179.65 kcal./kg. and 185.40 kcal./kg., respectively. Consequently, the theoretical heat input (Q) per 1000 kgs. of fresh water product would be 3870·(185.40–179.65), or 22.25 kcal. The performance ratio (R) would be 45 kgs. of product water per 1000 kcal. The circulation ratio (saline water/product water) would be 4. The flash range would be 180° C. —24° C., or 156° C.

The 3870 kgs. of fresh water at 183° C. would be used to heat the oil heat-exchanger H1 from about 18.5° C. to 181° C., and the hot oil would be used in heat-exchanger H2 to heat 4000 kgs. of raw saline water from 17.5° C. to 180° C.

Assuming in this example that the saline water reservoir R1 is placed 5 meters below pump P1, and that reservoir R2 is placed 95 meters higher, and further assuming that the pump operates at 85% efficiency, the work W1 required to drive 4000 kgs. of sea water from reservoir R1 to reservoir R2 is (4000)·(5+95):(0.85) or 470,600 kg.m.

The saline water is heated in heat-exchanger H2 from 17.5° C. to 180° C., and enters the boiling duct BD1 of the first stage at 180° C., and at a pressure of 100.3 meters water. The boiling saline water within the boiling duct causes the mixture of brine and water vapor to flow into the first evaporator E1, where the temperature and pressure are 178.5° C. and 96.9 meters water, respectively.

As indicated earlier, from 4000 kgs. of sea water (at 180° C. and 100.3 meters water pressure) introduced into the first-stage evaporator E1, approximately 3000 kgs. of brine containing 4.7% salt flows out of the last-stage evaporator E*n* (at 24° C. and 0.292 meters water pressure). This brine flows from conduit B*n* into the open tank R3 placed at 15 meters below the last-stage evaporator E*n* (No. 103), and has a height of more than 5 meters. The brine height within the vertical or inclined duct B*n* above the brine level in tank R3 will be 10 meters and is approximately equal to the difference between the vapor pressure in the last-stage evaporator E*n* and the atmospheric pressure in meters of water at the brine level in tank R3. From tank R3, approximately 3000 kgs. of brine will flow through pipe **20** to power generator M1 placed at about 290 meters lower, and will leave through the brine outlet 22 at atmospheric pressure. The work generated (W'1) at 85% efficiency is (3000)·(290)·(0.85), or 739,500 kg.m.

The flashed vapor flowing from the last-stage evaporator E*n* at a temperature of 25.5° C., and a pressure of 0.320 meters of water, is condensed by approximately 2870 kgs. of fresh water (at 20.5° C.) flowing from the closed tank R4 through the last water duct WD*n* into the last-stage condenser C*n*. The temperature and quantity of the distilled water entering the last-stage condenser C*n* are both increased by the water condensing in the various condensers from their respective evaporators, so that about 3870 kgs. of distilled water leaves the first-stage condenser C1, at a temepratıre of 177.5° C., and a pressure of 96.5 meters water. As indicated earlier, the temperature of this distilled water is raised to about 183° C. in heater 34 before it is introduced into heat-exchanger H1.

The height of heat-exchanger H1 is a little more than 40 meters. The absolute pressure at the outlet of pump P2 is equal to the water vapor pressure at 183° C. (109.5 meters water) plus approximately 40 meters of water pressure from the water column in pipe 36. The specific volume of water at 177.5° C. is 1.124 liters per kg. The volume of the distilled water flowing into the inlet of pump P2 is approximately 4350 liters, and the work W2 required to drive the fresh water from the bottom of the last-stage condenser C1 to the top of heat-exchange H1 is approximately (4350)·(109.5+40—96.5):(0.85), or 271,200 kg.m.

In heat-exchanger H1 the distilled water flows from top to bottom and is cooled form 183° C. to 20.5° C., the upflowing oil being heated from 18.5° C. to 181° C. Of the cold distilled water (3870 kgs.) flowing out of heat-exchanger H1, 2870 kgs. flows through pipe 42, pump P3, and pipe 44 into the open tank R5 at a height of about 290 meters above pump P3. The pressure at the inlet of pump P3 is about 109.5+40, or 149.5 meters water, and the pressure at the outlet of pump P3 is approximately 300 meters water, so that the work W3 required to drive 2870 kgs. of the fresh water from the bottom of heat-exchanger H1 to tank R5 is (2870)·(300—149.5):(0.85), or 508,100 kg.m.

The remaining 1000 kg. of distilled water flows through power generator M2 where the pressure drops from 149.5 to 10.33 meters water (atmospheric pressure), the work generated (W'2) being (1000)·(149.5–10.33)·(0.85), or 118,200 kg.m.

The flow control unit 46 between tank R5 and closed tank R4 is of a known construction including a butterfly valve, a float ball, a sight glass, and a pressure control system. The vapor pressure within tank R4 is approximately 0.246 meter water, corresponding to the water temperature of 20.5° C., and the pressure within the last-stage condenser C*n* is approximately 0.270 meters water, corresponding to the temperature 22° C. of the outflowing water, so that the equilibrium pressure head within the water duct WD*n* is 27.0—24.6, or 3.4 cm. of water. The equilibrium pressure head within the first water duct WD1 is 9650—9317, or 333 cm. water. The specific volume of water at 176° C. is 1.122, so that this pressure head would actually be (1.122)·(9650—9317), or 375 cm. water at 176° C.

As an example of an intermediary oil, there may be used "Mobiltherm 600" whose properties are described in "Handbook of Heat Transfer Media" by Paul L. Geiringer, pages 164–171. In heat-exchanger H1, 3870 kgs. of distilled water at 183° C. are introduced per unit time into the upper end through inlet pipe 36, and 9680 kgs. of the oil at 18.5° C. are introduced into the lower end through pipe 50. The water exits from the bottom of the heat-exchanger at 20.5° C. through pipe 38, as described above, whereas the oil exits at the three different locations of the heat-exchanger, as follows: (a) 980 kgs. exit at 73° C. through pipe 52; (b) 800 kgs. exit at 127° C. through pipe 56; and (c) 7900 kgs. exit at 181° C. through pipe 60.

The quantities of oil remaining in heat-exchanger H1 after the deviation of 980 kgs. at 73° C. is 8700 kgs., and that remaining after the second deviation of 800 kgs. at 127° C. is 7900 kgs. The mean specific heat of "Mobiltherm 600" oil between 18.5° C.–73° C. and 73° C.–123° C., and between 123° C.–181° C., is 0.40, 0.445, and 0.49, respectively. By multiplying the above oil quantities by the corresponding mean specific heat we obtain the same heat capacity, which is equal to the heat capacity of 3870 kgs. of pure water. The mean specific heat of pure water is about 1.01 in the temperature range of the process (assumed to be one in the above calculations). The specific heat of sea water is about 0.97 in the temperature range of the process. It is thus seen that the heat capacities of the two liquid streams in heat-exchangers H1 and H2 are equalized for the three temperature ranges 18.5° C–73° C., 73° C.–127° C., and 127° C.–181° C. The oil at 181° C., 127° C. and 73° C., respectively, expands within their respective power generators M3''', M3'' and M3', and the difference between the inlet and outlet pressures of these three power generators is approximately equal to the difference between the pressure at the outlet of pump P2 and the pressure of the saline water at 180° C., this pressure difference being 149.5–100.3, or about 49 meters water.

In heat-exchanger H2, the cold sea water (about 4000 kgs. per unit time), introduced at the top through pipe 6, passes in counter-current heat-exchange with the up-flowing hot oil and exits through pipe 8 (at 180° C.). The upflowing oil passes through the oil washing column OW at the upper end of heat-exchanger H2, exits through pipe 64 (at 18.5° C.) and is then driven by pump P4 back into heat-exchanger H1, as described earlier.

The volume of the expanding oil can be computed as follows: in power generator M3''', 7900 kgs. having a specific volume of 1.172, equalling 9300 liters of oil at 181° C.; in power generator M3'', 800 kgs., having a specific volume of 1.134, equalling 910 liters of oil at 127° C.; in power generator M3', 980 kgs. having a specific volume of 1.094, equalling 1070 liters at 73° C. The sum total is 9680 kgs. of oil equalling 11,280 liters. The work of expansion W3', assuming 85% efficiency, is: (11280).(49).(0.85), or 469,800 kg. m., i.e., 1.28 kwh.

The volume of the oil flowing out from the top of heat-exchanger H2 at 18.5° C. and driven by pump P4 into the bottom of heat-exchanger H1 is: (9680).(1.054), or 10200 liters. The work of compression W4 at 85% efficiency is (10200).(49):(0.85), or 588,000 kg. m., i.e., 1.6 kwh. The work input in the oil cycle at 80% efficiency is 1.6–1.28, or 0.32 kwh. per 1000 kgs. of distilled water produced.

The total work input in the water and oil cycles per 1000 kgs. of distilled water produced at 85% efficiency of the pumps and power generators is: W1+W2+W3+W4—W1'—W2'—W3'; or 470,600+271,200+508,100+588,000—739,500—118,200—469,800, equalling 764,200 kg. m., or 1.4 kwh.

It will be appreprıated that power generator M2 could be omitted, in which case the distilled water will exit through pipe 40 at a pressure of about 140 meters water, whereupon it could be directed to a high level reservoir.

For comparison purposes, there is set forth below a table (Table 1) providing a comparison of the operating conditions and the values of Q (theoretical heat input) and R (performance ratio) according to the example set forth above (referred to as Plant No. 1) and to three examples based on known flash evaporation systems (referred to as Plant Nos. 2, 3 and 4.) A second table (Table 2) is also set forth providing the liquid-liquid heat-exchange operating conditions with respect to Plant Nos. 1 and 4. Plant No. 2 is a multi-stage flash plant with brine heater at 290° F. and 250° F. (see page 152, Fig. 164 of "1965 Saline Water Conversion Report," published by the U.S. Office of Saline Water). Plant No. 3 is a multi-stage flash plant with brine recirculation that has been constructed in Eilat, Israel (see the article by L. Steinfeld, and A. Kikinis, published in "Journal of the Association of Engineers and Architects in Israel," May–June 1965). Plant No. 4 is the vapor reheat and liquid-liquid heat exchange process developed by FMC Corporation, Santa Clara, Calif. (see U.S. Office of Saline Water Report No. 78 of September 1963).

In Table 1, the liquid heated by the condensation of the flashed water vapor is designated "SW" for the incoming sea water and circulating brine in Plant Nos. 2 and 3, and by "W" for the circulating and out-flowing fresh water in Plant Nos. 1 and 4. The flashing brine is designated "B."

stages. These are referred to as chambers **1** and **2**, the saturated pressure and temperature being P_1, P_2 and T_1, T_2 respectively. The open bottom of the riser is submerged a little below the brine surface within the lower chamber **1**, and its open top is projected a little above the brine surface within the upper chamber **2**. In an example given in page 36–9 of the above reference:

P_1=14.7 lb./in.^{2}abs. (or 1.033 kg./cm.^{2})
P_2=13.3 lb./in.^{2}abs. (or 0.935 kg./cm.^{2})
T_1=212° F. (or 100° C.)
T_2=207° F. (or 97.2° C.)

In these conditions 5 kgs. of water vapor will flash from 1000 kgs. of saline water containing 3.5% of salts.

TABLE 1

Comparison of Operating Conditions and Values of Q and R

	Plant 1		Plant 2	Plant 3	Plant 4
No. of stages	103		72	30	20
LAST STAGE					
	Temp., °C.	Pres.(m. water)	Temp., °C.	Temp., °C.	Temp., °C.
B(out)	24	0.292	21.1	34.6	32.2
B(in)	25.5	0.320	22.93	36.56	37.2
B(in)-B(out)	1.5	0.028	1.83	1.96	5
W(in)	20.5	0.246			22.2
W(out)	22	0.270			27.2
W(out)-W(in)	1.5	0.024			5
SW(in)			15.6	27.8	
SW(out)			17.43	29.76	
SW(out)-SW(in)			1.83	1.96	
B(out)-W(out)	2	0.022			5
B(out)-SW(out)			3.67	4.84	
FIRST STAGE					
SW(out)-SW(in)			1.7	1.96	
SW(in)			136.3	84.64	
W(out)-W(in)	1.5	3.33			5
W(in)	176	93.17			
B(out)-SW(out)			3.6	4.96	
SW(out)			138	86.6	
B(out)-W(out)	1	0.70			5
W(out)	177.5	96.50			122.2
B(in)-B(out)	1.5	3.40	1.7	1.96	5
B(out)	178.5	96.90	141.6	91.34	127.2
B(in)	180	100.30	143.3	93.3	132.2
Q	$\frac{\text{22.3 Kcal.}}{\text{1 kg. product}}$		30.6	70	73
R	$\frac{\text{45 kg. product}}{\text{1,000 Kcal.}}$		32.7	14.3	13.3
Circulating ratio	4		5.8	10.5	5.5
Flash range, °C	156		122	59	100
Temp. rise by make-up heater, °C	5.5		5.3	6.7	13.3

TABLE 2

Comparison of Liquid-Liquid Heat-Exchange Operating Conditions

	Plant 1			Plant 4		
	Top		Bottom	Top		Bottom
Oil heater (H1):						
Water cooling, ° C	183	⟶	20.5	122.2	⟶	22.2
Oil heating, ° C	181	⟵	18.5	121.1	⟵	21.1
Δt, ° C	2		2	1.1		1.1
SW heater (H2):						
Oil cooling, ° C	18.5	⟶	181	21.1	⟶	133.9
SW heating, ° C	17.5	⟵	180	19.4	⟵	132.2
Δt, ° C	1		1	1.7		1.7

In these examples the intermediary fluid used is lighter than water, as is indicated by the arrows.

A further comparison may be made between the brine-lift system of the present invention and that of the above-mentioned "Clementine" system as described on pages 36–1 to 36–12 of the report on the Second European Symposium on Fresh Water from the Sea, Athens, May 9–12, 1967.

In the "Clementine" method of lift, as briefly described above, a riser (a vertical pipe increasing in cross section from bottom to top) connects two vertically mounted

The volume of the flashing or boiling mixture of brine and vapor at P_2=0.935 kg./cm.^{2} and T'_2=97.7° C. T'_2 is the saturation temperature of the saline water at P_2, while T_2 is the saturation temperature of pure water at P_2) is 10.000 l. approx., and the external energy of vaporization of 5 kgs. water is approx. 84,500 kilogram-meters (approx. 40 kcal. per kg. of vaporized water). Theoretically this amount of energy may lift the flashing or boiling mixture to a height of 84.5 meters. However a lift of only 0.91 to 0.95 meters was obtained experimentally by the "Clementine" system.

In the system of the present invention as described above, the outlet or brine ducts (B) and the boiling ducts (BD) are connected together at their bottoms and form U-tubes. These are used to connect the bottom of each evaporator chamber (called chamber **1**, below) with the succeeding upper chamber (called chamber **2**, below) at a point above the brine level of the latter. The brine flowing through the bottom of chamber **1** into the shorter branch (hereinafter called "brine branch") of the U-tube forms in this branch a brine column of a height h_1. The gauge pressure $Pg=h_1d$, where d is the density of the brine, assumed to be 1. This pressure together with the difference of pressure in the two chambers (P_1-P_2) will counterbalance the gauge pressure $Pg'=h_2d'$ of the boiling mixture column within the longer branch (hereinafter called the 'boiling branch") of the U-tube leading to the inside of chamber **2**, and the pressure drop Δp due to the flow of the brine and of the boiling mixture through the U-tube.

The conditions for the brine lift from stage 1 to stage 2 are defined by the inequality:

$$h_1d+(P_1-P_2)>h_2d'+\Delta P \quad (1)$$

The lift is $(P_1-P_2)+a$ (both in meters water). The symbol a is the height of the vapor-filled space of the condenser in which direct condensation is effected between the vapor and the cooling liquid, and the pressure drop in meters water due to fluid flow. For large plants, a is equal to or greater than 2 meters, preferably 3 meters. The lift for large plants is $h+3$ meters, where $h=P_1-P_2$ in meters water; and the height of the boiling mixture column h_2 is h_1+h+3 meters. In the above example $h=0.98$ meter water. Assuming that 50% of the vaporization takes place within the boiling branch and 50% within chamber **2**, then mean density d' of the boiling column for the conditions of the above example is approx.:

$$d'=1000/(2.5\times1780+997.5)=0.18 \text{ kg./l.}$$

and the above inequality gives:

$$h_1+0.98>(h_1+0.98+3)0.18+\Delta P \quad (2)$$

or

$$0.82h_1+0.26>\Delta P \quad (3)$$

if $h_1=2$ then $\Delta P<1.90$ meters water for a U-tube length of 12 meters approx.

It is easy to see that for a large diameter U-tube, and especially for a large diameter boiling branch, very great velocity of the boiling mixture could be reached by giving the brine branch of the U-tube the required height.

The U-tube system or any system (e.g., Well-type manometer system, as described below with respect to FIG. 5) using a brine head as counterbalance for brine lift against gravity, could be applied also at low temperature stages.

In the example given earlier, the conditions of the last two stages (n-1 and n, or 102 and 103) are:

(1) Saturation pressure and temperatures:

Stage 102

$P_1=0.032$ kg./cm.^{2}
$T'_1=25.5°$ C.

Stage 103

$P_2=0.0292$ kg./cm.^{2}
$T'_2=24°$ C.

(2) $h=P_1-P_2=0.320-0.292=0.028$ meter water.

(3) 1.8 kgs. water are vaporized from 750 kgs. of brine containing 4.6% of salts.

Assuming that 50% of the flashing will take place within the flashing or boiling branch and 50% within chamber 2, then the mean density d' of the boiling column is approx.:

$d'=750/(0.9\times43,000+749.1)=0.019$ kg./1 and the inequality (1) gives:

$$h_1+0.028>(h_1+0.028+3)0.019+\Delta P \quad (4)$$

or

$$0.981\,h_1-0.03>\Delta P \quad (5)$$

if $h_1=2$ then $\Delta P<1.93$ meter water for a total U-tube length of 10 meters approx.

The external energy of vaporization of 0.9 kgs. water flashed within the boiling branch is approx. 12,500 kg.-meters. (32.7 kcal. per kg. approx.).

This means a theoretical lift of 17 meters approx. for 50% of flashing within the boiling column, while only 3.028 meters of lift are required for large size desalination plants, between the low temperature stages

$$(T'_1=25.5° \text{ C. and } T'_2=24° \text{ C.})$$

In the "Clementine" system h_1 is the small part of the riser submerged in the brine within chamber **1**. If we take $h_1=0.1$ meters, then inequality (5) gives:

$$0.981\times0.1-0.03=0.068>\Delta P$$

or

$$\Delta P<0.068 \text{ meter water}$$

This explains why it is not possible to obtain valuable lift at low temperature stages with the "Clementine" system. It also explains why the "Clementine" system study indicates that the riser should increase in cross section from bottom to top to avoid acceleration of steam/brine mixture as it flashes (p. 36-3 of the above reference).

DESCRIPTION OF THE EMBODIMENT OF FIG. 4

FIG. 4 illustrates another embodiment of the invention, in which each stage is constituted of a single double-compartment unit EC1, EC2 . . . ECn, one compartment serving as the evaporator and the other as the condenser.

Each unit is in the form of a cylindrical tank having a partition **150** dividing the tank into an evaporator compartment E and a condenser compartment C. U-tubes **TT1**, **TT2** . . . TTn connect all the evaporator compartments in series, and inclined tube CCP connects all the condenser compartments in series. The brine, from which the vapor is flashed, moves upwardly against gravity through the U-tubes and evaporators in succession. The pure-water upon which the flashed vapor is condensed moves downwardly by gravity through inclined tube CCP and through the condensers in succession.

With reference to unit EC2, FIG. 4, the pure-water flowing downwardly through inclined tube CCP enters the condensing chamber through inlet **152** and flows upwardly through a passageway **154** forming a pool at the top of a perforated plate **156**. It then passes downwardly by gravity, in the form of threads or drops, through the latter plate and exits through outlet **158** back into the inclined tube CCP where it flows to the next lower-stage unit, EC1 in this case. Inclined tube CCP is provided with a baffle **160** directing the pure-water to move in this path.

The brine in the vaporator conduit system EEP moves upwardly against gravity through the U-tubes **TT1**, **TT2**, etc., from the evaporater E of one stage to that of the next stage. This is lifted from one stage to the next by the energy of vaporization of the water. For this purpose, each of the U-tubes (e.g., U-tube **TT2** connecting the evaporators of EC1 and EC2) includes a short leg Ts (comparable to the brine outlet duct B1, B2, etc., in FIG. 1) connected to the outlet of one evaporator, and a long leg Tg (comparable to the boiling ducts BD1, BD2, etc., in FIG. 1) connected to the inlet **162** of the next, higher elevation evaporator. The vapor pressure within the evaporator of the lower stage unit (EC1) is such that the brine within short leg Ts is in liquid form, thus providing a down-flowing liquid column L in short leg Ts. The vapor pressure within the evaporator of the higher stage unit (EC2) is slightly less, so that the brine moving upwardly through the long leg Tg is in boiling condition, thereby forming an up-flowing boiling mixture column BM of brine and water vapor. The bottom of this boiling-mixture column is indicated at BM', which is at a level below the top of the down-flowing liquid column L in the short leg Ts. The portion of the down-flowing liquid column L in the short leg Ts above the level BM' is of sufficient height that its gauge pressure, together with the pressure difference between the two stages, counterbalances the gauge pressure of the up-flowing boiling-mixture column BM in the long leg Tg, and the pressure losses due to fluid flow. Thus, the boiling mixture is lifted from the evaporator of one stage to that of the next while the liquid in liquid column L, as well as in the juncture thereof with the up-flowing boiling mixture column BM (i.e., the liquid to the level

BM'), effects a hydraulic seal between the adjacent evaporators.

In FIG. 4, h represents the difference in vapor pressure, in meters of water, between the units EC1 and EC2; $h1$ represents the height of the liquid column L in short leg Ts above the level BM'; and $h2$ represents the height of the boiling mixture column BM in the long leg Tg. For large size plants, an efficient condenser should include condensing and flow spaces of a height of approximately 1.5 meters, and an additional 0.5 meter of height should be reserved for pressure drop due to the down-flow of the recirculating pure water from stage to stage. The U-tube should therefore be designed so that the height ($h2$ of the boiling mixture column BM is greater than $h1+h+2$ meters, preferably being equal to $h1+h+3$ meters, to effect a lift of the brine from one evaporator to the next evaporator stage.

The boiling mixture flows through inlet **162** into the evaporator wherein the flashing is completed. The liquid phase (i.e., brine) drops to the bottom and flows out through the short leg Ts of the U-tube to the next evaporator, as described earlier. The vapor rises, passes through a deminster **164**, down through a passageway **166**, and up through a perforated plate **168** of large size diameter perforation into the condenser chamber C where it condenses on the down-flowing pure-water threads and/or drops flowing through the condenser by gravity. The water quantity and temperature are thus increased as the pure water flows downwardly through the condenser conduit system from one condenser to the next. The pressure difference between the two stages is maintained by a vacuum control device **170** in each stage, the device **170** being shown as including a tube having openings communicating with the condenser chamber both above and below the pool formed on perforated plate **156**. The liquid between the baffles **160** of two consecutive stages forms a hydraulic seal between the stages.

VARIATION OF FIG 5

It will be appreciated that the same type of brine lift system can be provided by conduit means other than the U-tubes **TT1**, **TT2**, etc., and also that the evaporator of each stage may be in a separate unit from the condenser.

Such a variation is shown in FIG. 5, wherein each evaporator **E1'**, **E2'** is in a separate unit; also instead of U-tubes, there is used a "Well-type manometer" arrangement. Each evaporator is in the form of a horizontal cylinder and is connected to a vertical tube or leg **172** closed at its bottom. The lower end of a tube **174** is disposed near the bottom of leg **172**, its upper end being connected to the inlet of the adjacent higher evaporator **E2'** above the liquid level therein. The vapor flows upwardly through a demister **176** to an outlet conduit **178** connected to the condenser of that stage.

It will be appreciated that the height ($h1$) of the liquid within the evaporater and its leg **172**, from the upper level thereof to the level of the bottom **174'** of the boiling mixture column formed therein, constitutes the equivalent of the liquid column L in the short leg of Ts of the system described in FIG. 4, and that the height ($h2$) of the boiling-mixture column within tube **174** corresponds to that of boiling-mixture column BM within the long leg Tg of FIG. 4. The liquid in the evaporator and the bottom of tube **172**, together with the liquid in the lower part of tube **174**, up to the level **174'**, effect the hydraulic seal between the evaporator stages.

If desired, a plurality of cylinders or legs **172**, with a tube **174** for each, may be provided between the evaporators to accommodate the flow of a larger quantitiy of the boiling mixture. A similar plurality of U-tubes (e.g. **TT1** . . . **TT***n*) may be provided in the embodiments of FIGS. 1–4.

Also, it may be required to agitate the mixture passing from one evaporator to the next. This is preferably done by injecting water vapor or by mechanically agitating the liquid mixture at the bottom of the longer leg of the U-tube in FIGS. 1–4, or at the bottom of the tube **174** in FIG. 5.

It will be seen from FIGS. 4 and 5, that, as above-described in connection with FIGS. 1*a* and 1*b*, the cross-sectional areas of the down-flowing and up-flowing column are of substantially the same order of magnitude. Further, as can be readily seen in FIGS. 4 and 5, the communicating path between adjacent evaporator stages formed by these columns is continuously opened and unrestricted throughout its length. The down-flowing column in each case is entirely of liquid in the form of saline water, while the longer up-flowing column includes a lower portion also of liquid in the form of saline water and an upper portion of a boiling mixture of liquid and vapor.

The herein described system is to be distinguished from that described by C. R. Wilke, C. T. Cheng, A. L. Lendasma and J. S. Porter, as published in Chemical Engineering Progress, vol. 59, No. 12, December 1963, proposing heating the incoming sea water by direct condensation of water vapor thereon, and using a cycle of an oil (Aroclor) as cooling liquid in the condensers and as heating liquid in "tube mixers where heat transfer takes place between Aroclor and sea water feed." That system uses pumps to drive liquid Aroclor and water condensate from one condenser to the next of higher temperature and vapor pressure. In addition, the incoming sea water is heated by direct condensation in a horizonal multi-stage system, and the feed is pumped from one condensing heating stage to the next of higher vapor pressure.

However, as shown in FIG. 14 of the above publication, the fuel cost decreases while the power cost increases with the increased number of stages. In the system described "A 20-stage plant was selected as approximately optimum" by the authors of the article. A heat investment of approximately 80 kcal. per kg. of pure water produced is required by that system.

The present invention, however, enables the construction of apparatus having a much larger number of stages, there being 103-stages in the embodiments described. As shown by the foregoing data, the expected heat investment per kg. of pure water is substantially lower than that described in the system of the above publication apart from the substantial reduction in the power and equipment costs obtained by the elimination of the pumps at each stage.

To reduce the overall height required for a system having a large number of stages, the evaporators and condensers could be arranged in a plurality of banks, the first stage of each bank starting out at the same lowest elevation, and the last stage of each bank ending at the same highest elevation.

It is to be understood that the values set forth above are based on preliminary calculations given merely to illustrate the principles of the invention, and are not necessarily exact nor required to utilize successfully the various features of the invention described. It is to be further understood that the embodiments described are purely for purposes of example, and that many other variations, modifications and applications of the invention, or the several features theerof disclosed, may be made.

What is claimed is:

1. A method of distilling feed liquid by passing same through a plurality of evaporator stages of successively decreasing temperature and vapor pressure and of successively increasing elevation, characterized in utilizing the energy of vaporization of the feed liquid, in addition to the pressure difference between stages, to drive the feed liquid through the evaporator stages in succession by: passing the feed liquid through a continuously-open unrestricted communicating path having two legs connecting each pair of adjacent stages; forming in one leg of said communicating path a continuously-flowing down column of feed liquid flowing out of one evaporator stage at the bottom thereof; forming in the other leg of said communicating path a continuously-flowing up column containing feed liquid in the liquid state at the lower end and above it a boiling-mixture of feed liquid and vapor flowing into the adjacent evaporator stage always at a point above the liquid level therein and above the pressure head height corresponding to the difference in pressure between the two stages; driving, with the flowing down column of feed liquid through the communicating path, the boiling mixture of the up column into said adjacent evaporator stage by causing the liquid from the down-flowing column to continue its flow back up, in a continuous flowing liquid path, into the up-column; the cross-sectional area of the down-flowing column being of the same order of magnitude as that of the up-flowing column; and continuously effecting a hydraulic seal between the adjacent evaporator stages by the liquid at the lower end of said up-flowing column.

2. Feed liquid distillation apparatus having a plurality of evaporator stages of successively decreasing temperature and vapor pressure and of successively increasing elevation, characterized in that there is included a continuously-open unrestricted communicating path connecting each pair of adjacent stages and utilizing the energy of vaporization of the feed liquid, in addition to the pressure difference between stages, to drive the feed liquid through the evaporator stages in succession, said communicating path including one leg connected to the bottom of one evaporator stage in which leg there is formed and maintained a continuously-flowing down column of the feed liquid flowing out of said stage, and a second leg connected to the adjacent evaporator stage at a point above the liquid level therein and above the pressure head height corresponding to the difference in pressure between the two stages, in which second leg there is formed and maintained a continuously-flowing up column containing the feed liquid in the liquid state at the lower end and above it a boiling-mixture of the feed liquid and vapor flowing into said adjacent evaporator stage, the liquid constituting the down-flowing column extending continuously therefrom and up into the up column as a continuous flowing liquid path; the cross-sectional area of said one leg being of the same order of magnitude as that of said other leg, the liquid at the lower end of said up-flowing column in said other leg continuously forming and maintaining a hydraulic seal between the adjacent evaporator stages.

3. Distillation apparatus according to claim 2 further including a condenser conduit system having a plurality of series-connected condensers, one for each stage, at different temperatures and vapor pressures decreasing from one stage to the next, said plurality of condensers also being disposed at increasing elevation with the highest temperature and vapor pressure condenser at the lowest elevation, there being a pump for pumping a cooling liquid to the condenser of highest elevation and lowest pressure, said cooling liquid passing by gravity from one condenser to the next while it is heated by the condensation thereon of the vapor from the evaporator of its respective stage,

said condenser conduit system including a communicating path connecting together each pair of adjacent condensers in which communicating path there are formed and maintained a down-flowing cooling liquid column and a shorter up-flowing cooling liquid column.

4. Distillation apparatus according to claim 3, wherein said condenser communicating path includes a duct leading from the lower end of the condenser of each of said stages into the interior of a receptacle disposed in the condenser of the next lower stage.

5. Distillation apparatus according to claim 3, further including a first heat-exchanger for heating an intermediary heat-exchange medium by direct contact with the hot liquid distillate leaving said condenser conduit system, and a second heat-exchanger for heating feed liquid by direct contact with the hot intermediary heat-exchange medium.

6. Distillation apparatus according to claim 5, wherein said intermediary heat-exchange medium is a liquid immiscible with said feed liquid and liquid distillate, the apparatus further including means for diverting a small fraction of the cold liquid distillate to the second heat-exchanger at the end thereof through which the intermediary immiscible liquid exits, producing a separating layer of liquid distillate between the feed liquid and the exiting intermediary immiscible liquid in which separating layer the latter liquid coalesces, thereby diluting the feed liquid entrained in the intermediary immiscible liquid before the latter is directed to said first heat-exchanger.

7. Distillation apparatus according to claim 5, wherein both said heat exchangers are of the vertical, direct-contact type, there being means for providing a plurality of cycles of flow of the intermediary heat-exchange medium between said heat-exchangers and each operating within a temperature range different from that of the other cycles, both of said heat-exchangers including for each cycle separate heat-exchange portions having a different horizontal cross-sectional area than that of the other portions.

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