# United States Patent [19]

**Loeb**

[11] **3,906,250**

[45] **Sept. 16, 1975**


[54] **METHOD AND APPARATUS FOR GENERATING POWER UTILIZING PRESSURE-RETARDED-OSMOSIS**


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[73] Assignee: **Ben-Gurion University of the Negev Research & Development Authority,** Beersheba, Israel



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[21] Appl. No.: **480,623**

[30] **Foreign Application Priority Data**

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May 10, 1974 Israel ..................................... 44799

[52] **U.S. Cl.** ........................ **290/1;** 60/649; 60/673; 310/2

[51] **Int. Cl.**[2] ........................................... **F03G 7/06**

[58] **Field of Search** ................ 290/1, 2; 310/2, 651; 60/641–673


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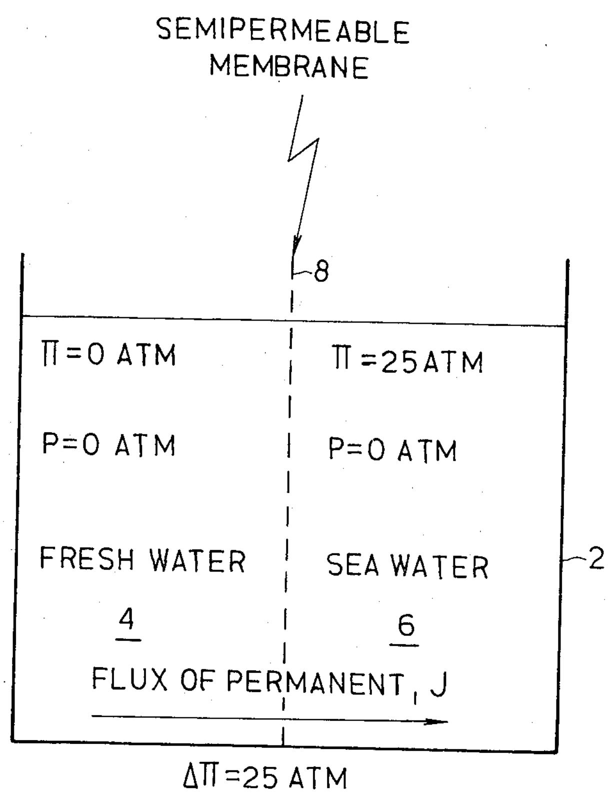
*Primary Examiner*—Robert K. Schaefer
*Assistant Examiner*—John W. Redman
*Attorney, Agent, or Firm*—Benjamin J. Barish


[57] **ABSTRACT**


A method and apparatus are described for generating power by utilizing pressure-retarded-osmosis. A first liquid having a relatively high osmotic pressure is introduced at a relatively high hydraulic pressure into a first pathway in which it contacts one face of a semi-permeable membrane, and a second liquid having a lower osmotic pressure is introduced at a lower hydraulic pressure into a second pathway in which it contacts the opposite face of the membrane. At every point in the two pathways, the hydraulic pressure difference between the liquids on the opposite faces of the membrane is maintained at a value which is less than the osmotic pressure difference between the liquids. Part of the second liquid passes by pressure-retarded-osmosis through the semi-permeable membrane, forming a pressurized mixed solution of greater volume than that of the first liquid introduced into the first pathway. The potential energy stored in the pressurized mixed solution is then converted into useful energy, such as electrical or mechanical power.

According to a further feature included in some of the described embodiments, after the potential energy stored in the pressurized mixed solution is converted into useful energy, the first and second liquids are recovered by separating from the mixed solution a quantity of the second liquid equal to the quantity which passed through the membrane, the original temperatures of the so-recovered first and second liquids are restored, the original hydraulic pressure difference is reapplied between the recovered first and second liquids, and the recovered first and second liquids are then recycled through the first and second pathways.


**31 Claims, 19 Drawing Figures**



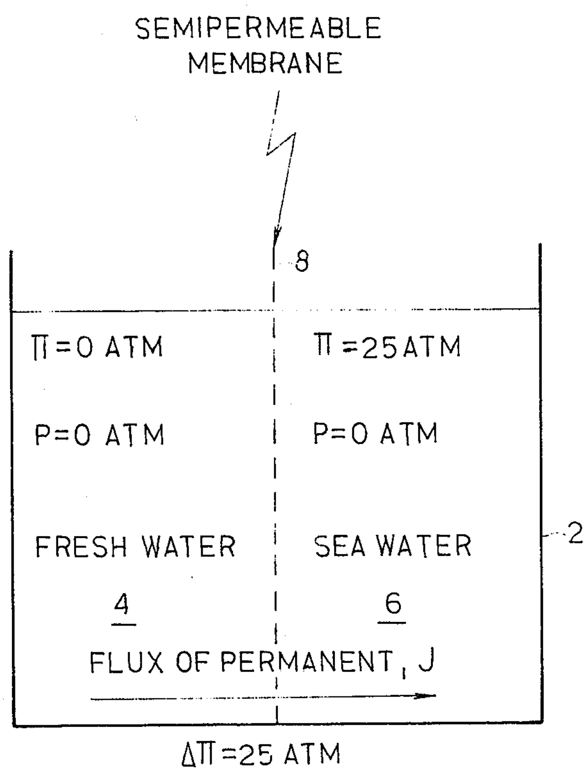



FIG.1

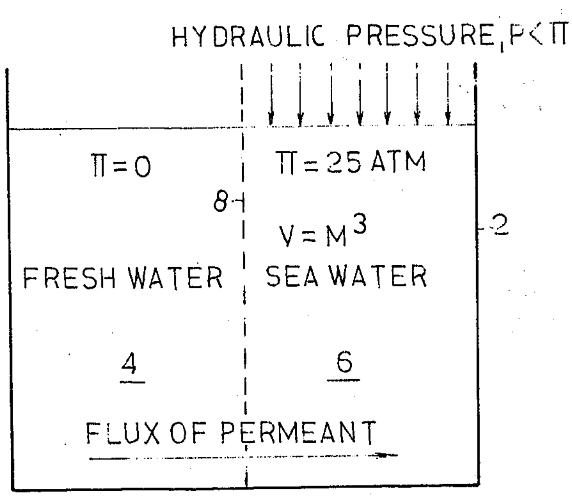



FIG. 2a

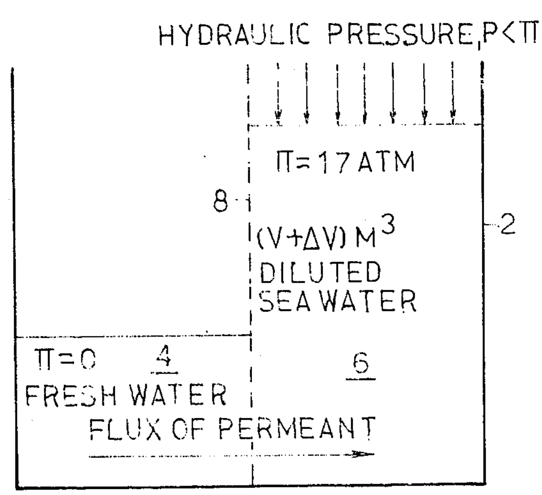



FIG. 2b

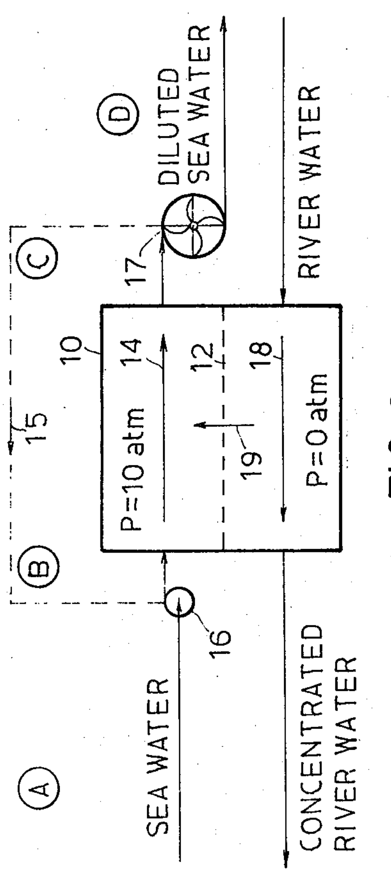



FIG. 3

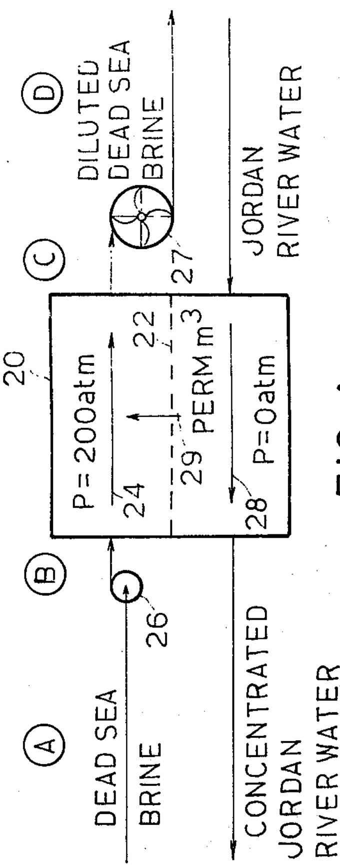



FIG. 4

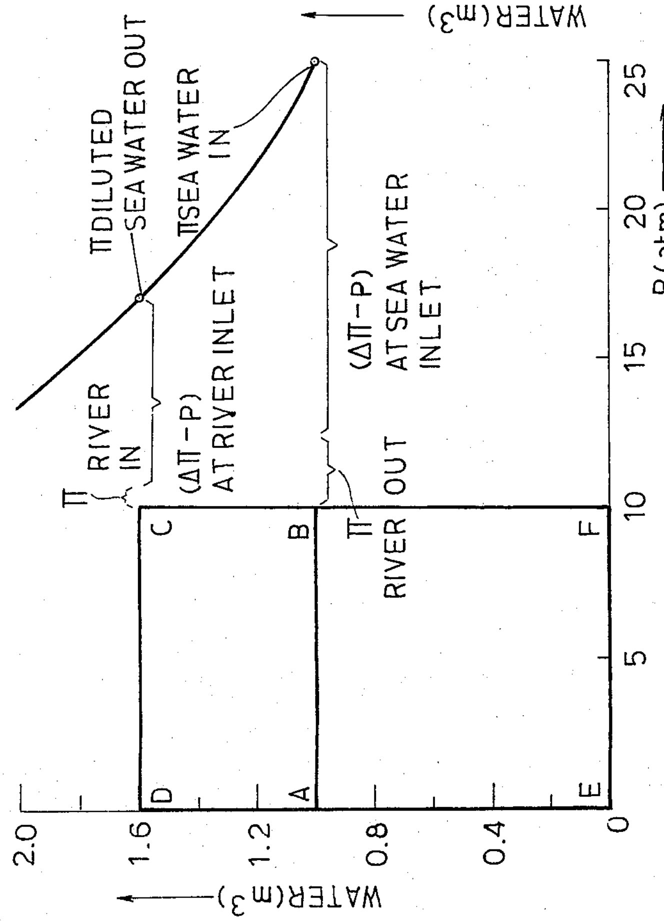



FIG. 3a

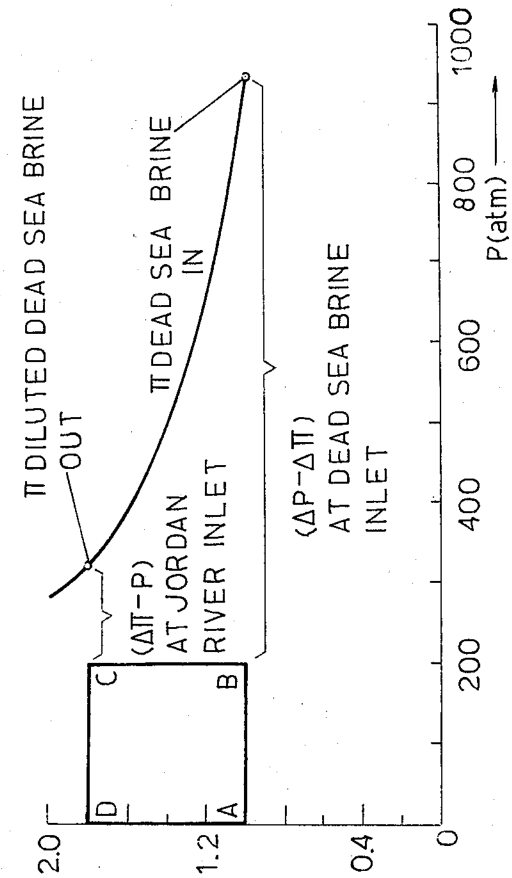



FIG. 4a

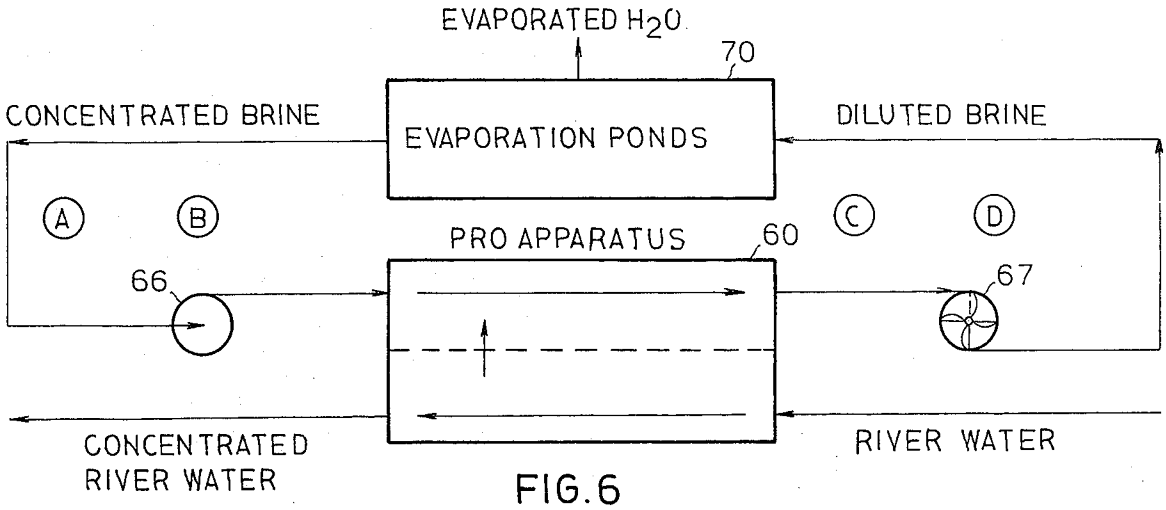



FIG. 6

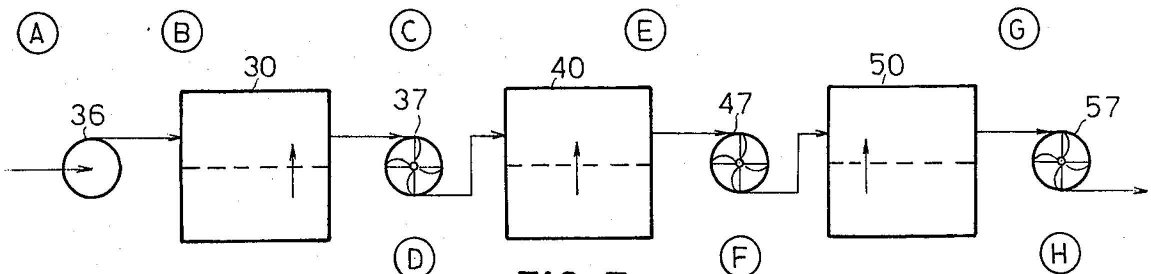



FIG. 5

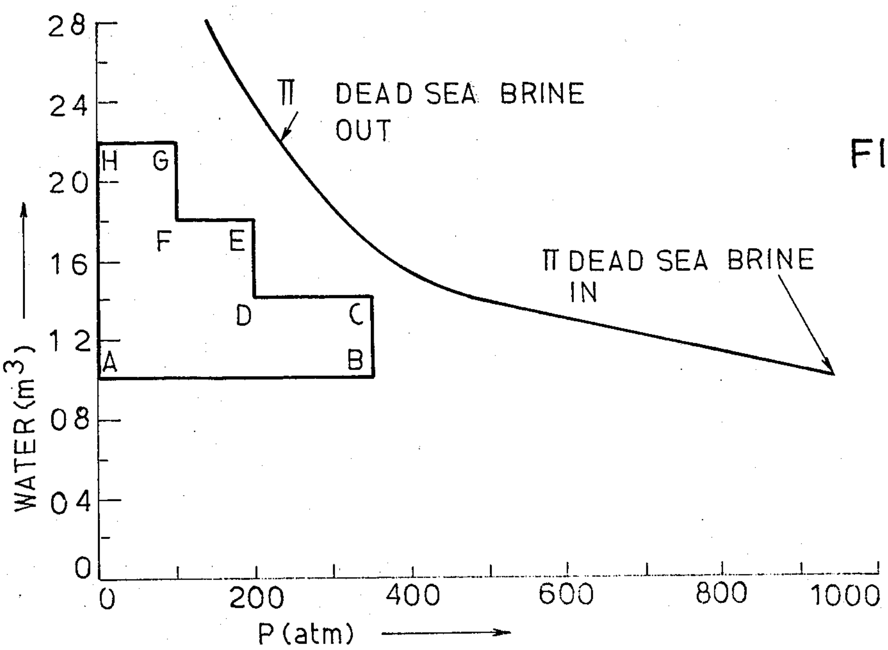



FIG. 5a

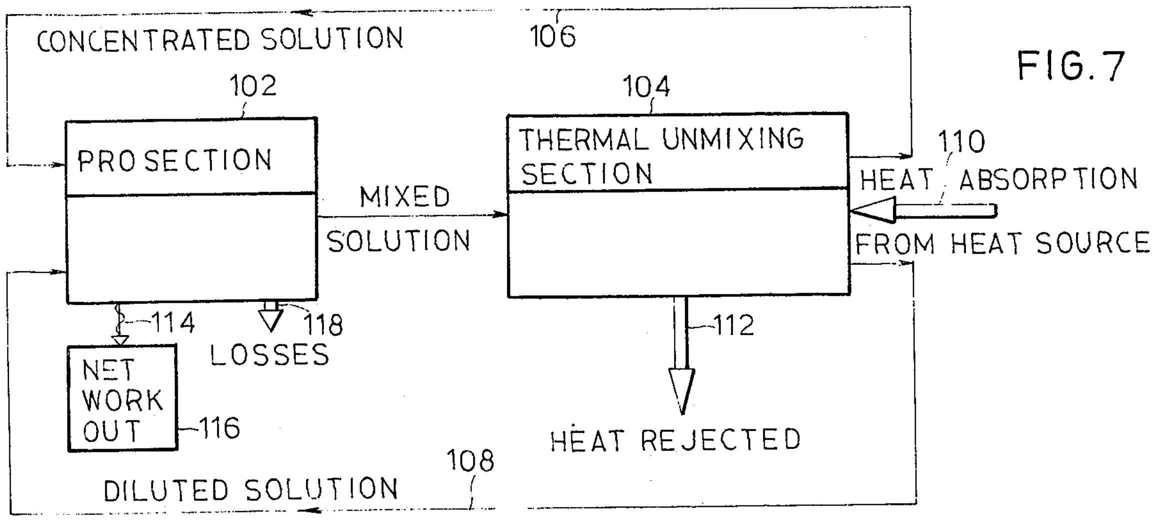



FIG. 7

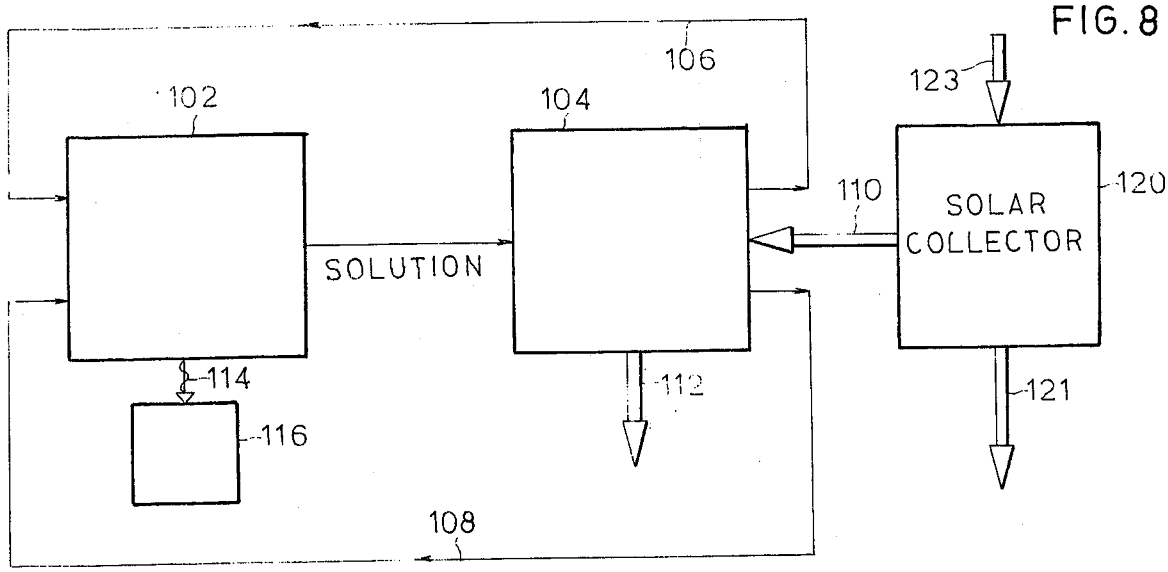



FIG. 8

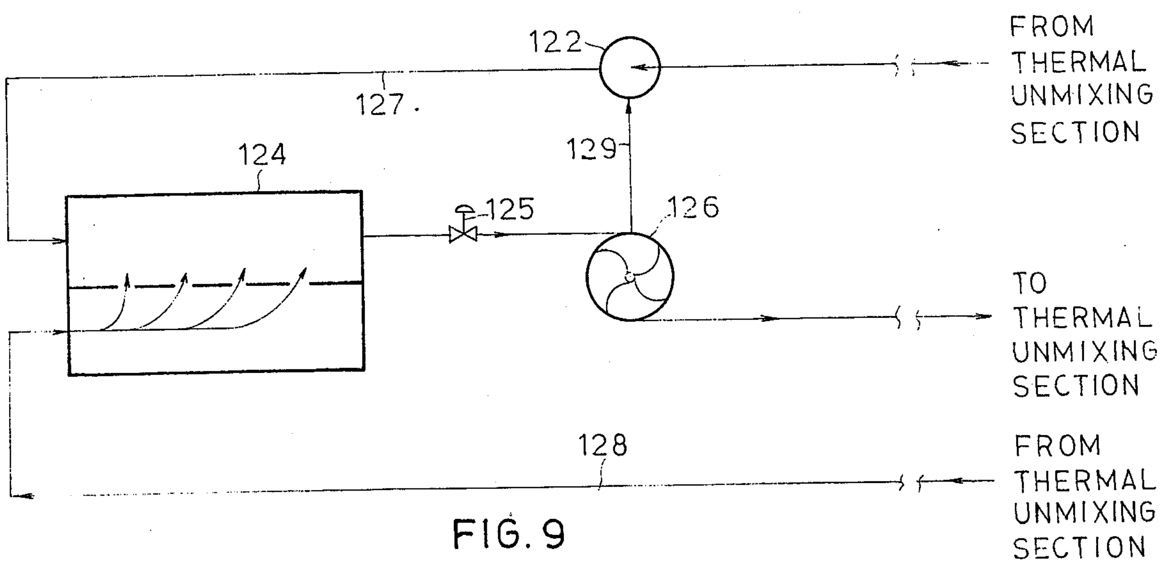



FIG. 9

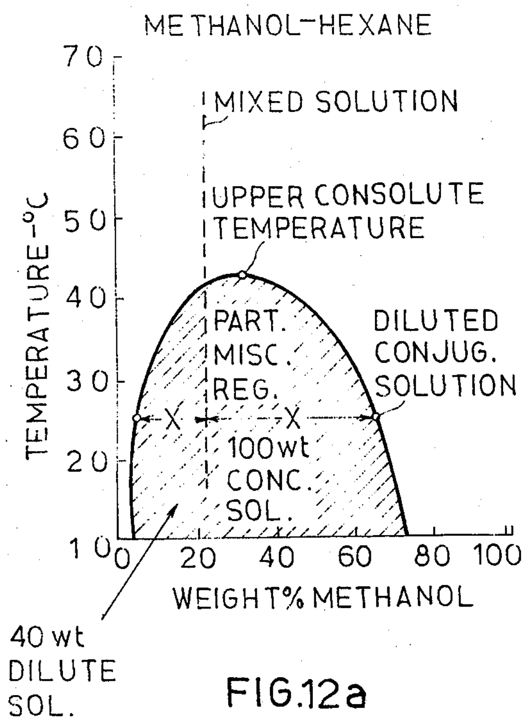



FIG.12a

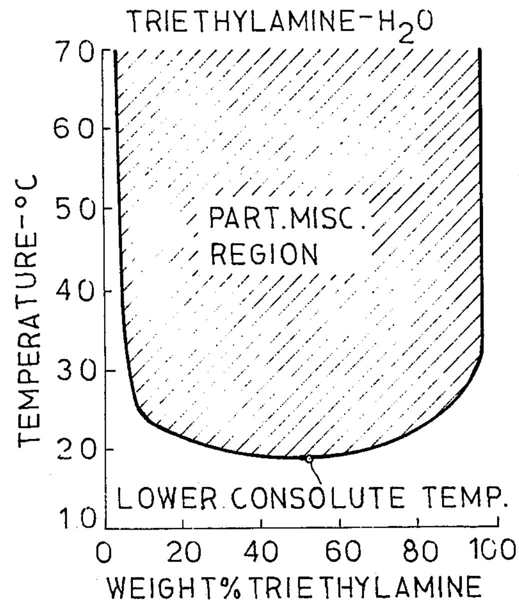



FIG.12b

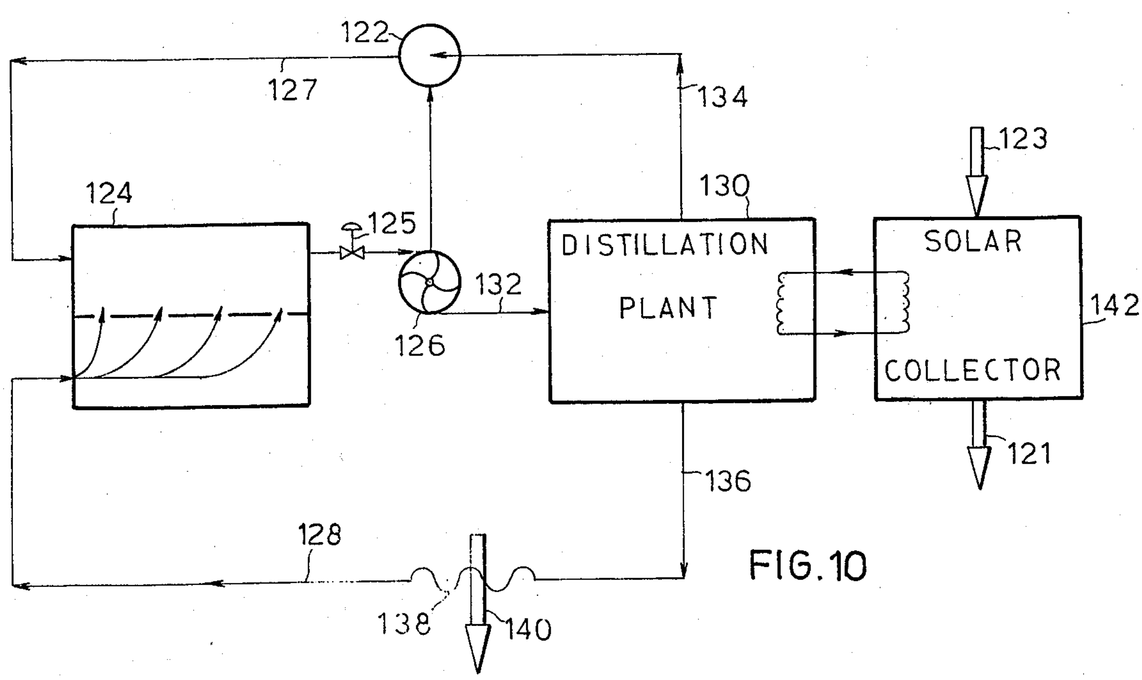



FIG.10

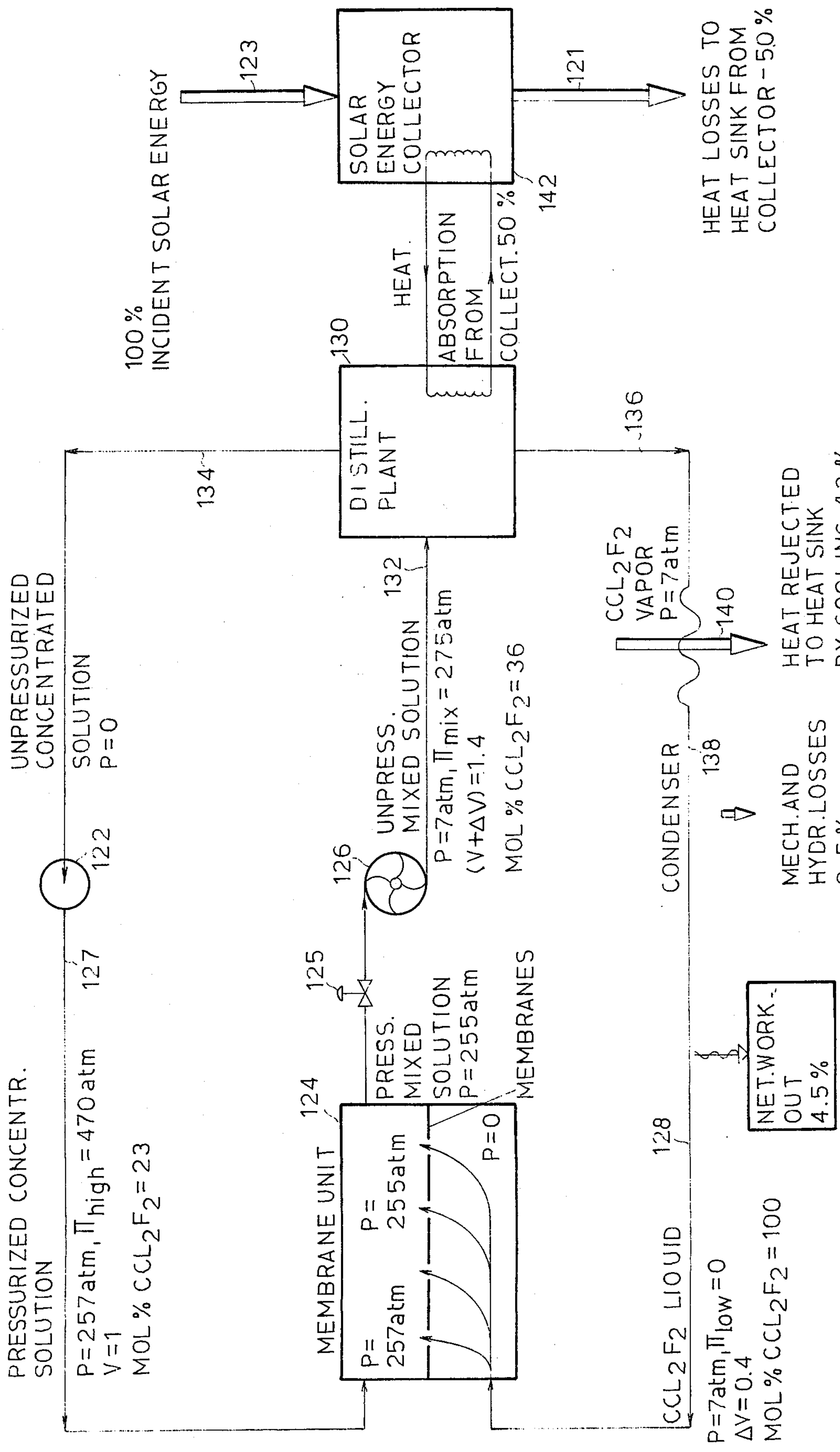

100 % INCIDENT SOLAR ENERGY
123
SOLAR ENERGY COLLECTOR
121
HEAT LOSSES TO HEAT SINK FROM COLLECTOR - 50 %
142
HEAT. ABSORPTION FROM COLLECT. 50 %
130
DISTILL. PLANT
134
136
UNPRESSURIZED CONCENTRATED SOLUTION P=0
132
UNPRESS. MIXED SOLUTION
P = 7atm, Π mix = 275atm
(V+ΔV) = 1.4
MOL % CCL2F2 = 36
CCL2F2 VAPOR P = 7atm
140
HEAT REJECTED TO HEAT SINK BY COOLING 43 %
122
126
CONDENSER
138
MECH. AND HYDR. LOSSES 2.5 %
127
125
PRESSURIZED CONCENTR. SOLUTION
P = 257atm, Π high = 470atm
V = 1
MOL % CCL2F2 = 23
PRESS. MIXED SOLUTION P = 255atm
MEMBRANES
124
MEMBRANE UNIT
P = 257atm
P = 255atm
P = 0
NET WORK OUT 4.5 %
128
CCL2F2 LIQUID
P = 7atm, Π low = 0
ΔV = 0.4
MOL % CCL2F2 = 100

FIG.11

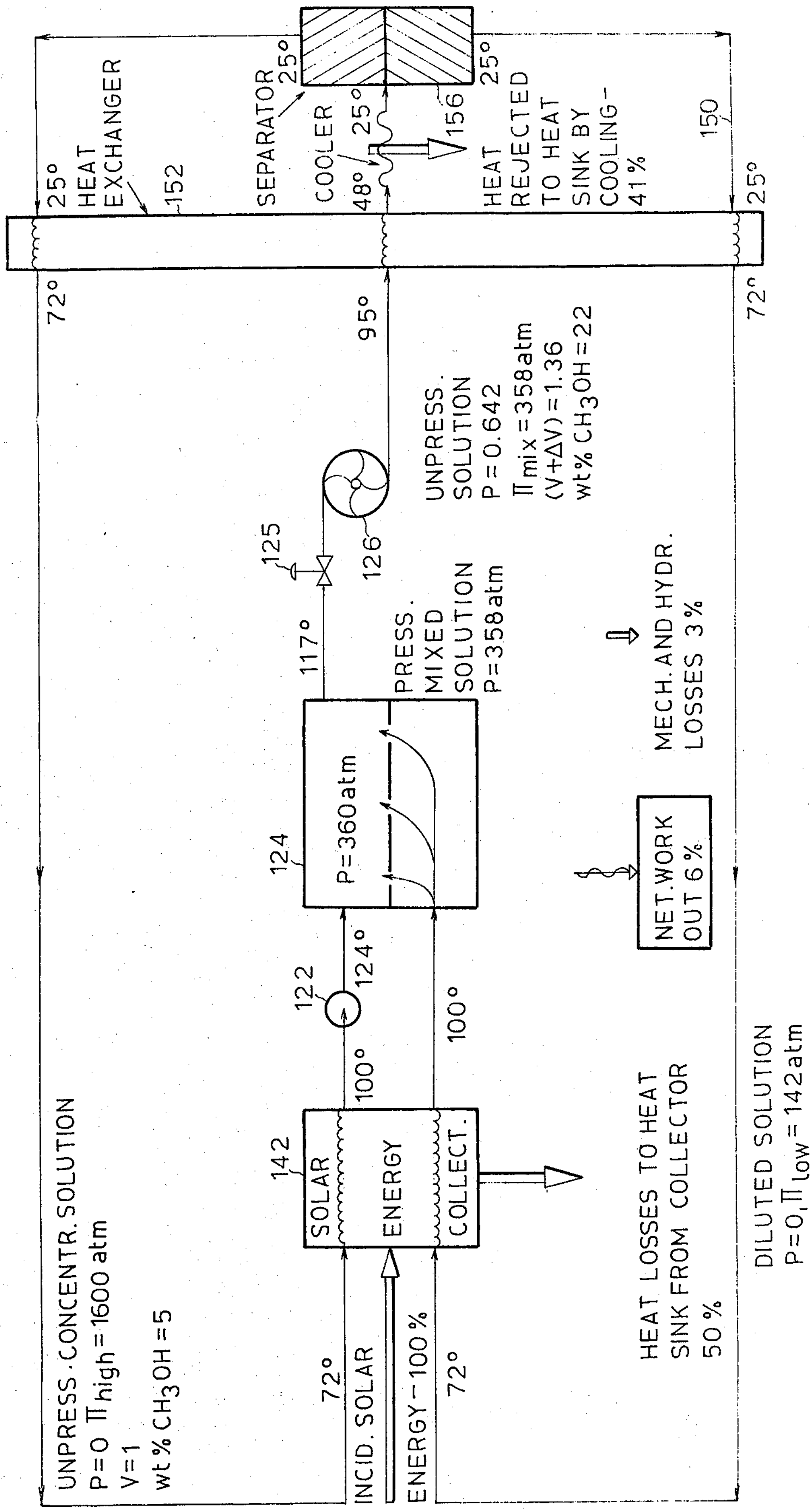

UNPRESS. CONCENTR. SOLUTION
P=0 $\Pi_{high}$=1600 atm
V=1
wt % $CH_3OH$=5
INCID. SOLAR ENERGY−100 %
72°
72°
142
SOLAR ENERGY COLLECT.
HEAT LOSSES TO HEAT SINK FROM COLLECTOR 50 %
100°
100°
122
124°
124
P=360 atm
PRESS. MIXED SOLUTION P=358 atm
117°
125
126
UNPRESS. SOLUTION
P=0.642
$\Pi_{mix}$=358 atm
(V+ΔV)=1.36
wt % $CH_3OH$=22
95°
NET WORK OUT 6 %
MECH. AND HYDR. LOSSES 3 %
DILUTED SOLUTION
P=0, $\Pi_{low}$=142 atm
ΔV=0.36
wt % $CH_3OH$=65
72°
72°
25°
HEAT EXCHANGER
152
SEPARATOR
25°
COOLER
48°
25°
156
25°
HEAT REJECTED TO HEAT SINK BY COOLING− 41 %
150
25°

FIG.13

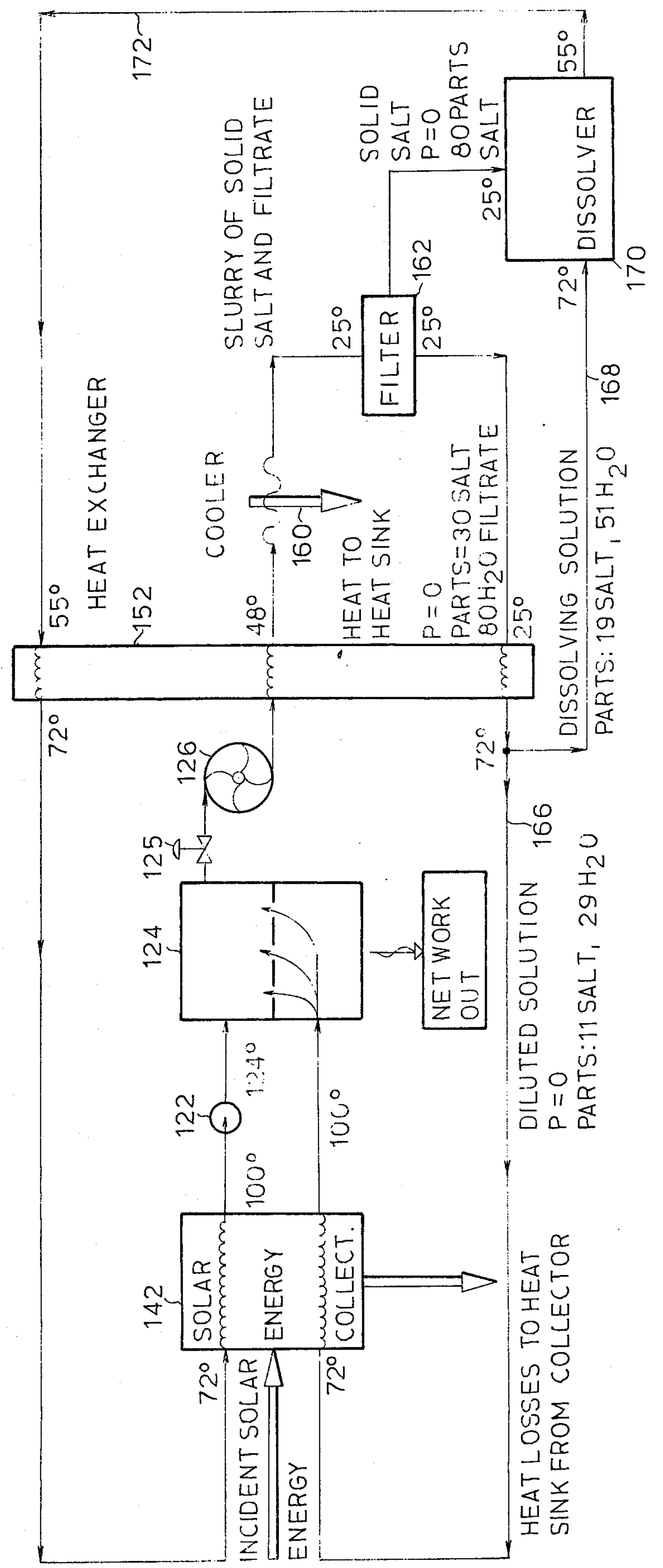

172
HEAT EXCHANGER
152
55°
72°
COOLER
160
48°
SLURRY OF SOLID SALT AND FILTRATE
25°
FILTER
162
25°
SOLID SALT P=0 80 PARTS SALT
25°
DISSOLVER
170
55°
72°
168
HEAT TO HEAT SINK
P = 0 PARTS=30 SALT 80 H2O FILTRATE
25°
DISSOLVING SOLUTION PARTS: 19 SALT, 51 H2O
72°
166
126
125
124
NET WORK OUT
122
124°
100°
100°
DILUTED SOLUTION P = 0 PARTS: 11 SALT, 29 H2O
142
SOLAR ENERGY COLLECT.
72°
72°
INCIDENT SOLAR ENERGY
HEAT LOSSES TO HEAT SINK FROM COLLECTOR

FIG. 14

# METHOD AND APPARATUS FOR GENERATING POWER UTILIZING PRESSURE-RETARDED-OSMOSIS

## BACKGROUND OF THE INVENTION

The present invention relates to a method and apparatus for generating electrical or mechanical power, and particularly to a method and apparatus which generates the power by exploiting certain naturally available sources which have not heretofore been exploited at all, or solar energy which has heretofore been exploited only by the use of very expensive and/or low-efficiency devices.

With the widely-predicted energy crisis looming in the not too distant future because of the dwindling supplies of fossil fuels, it has become vital to develop new sources of energy not based on fossil fuels. The present invention provides a new method and apparatus for generating power by utilizing two liquids having different osmotic pressures, and particularly by exploiting the osmotic interaction between such two liquids.

The different osmotic pressure liquids that may be used for generating the power can be obtained from a number of sources, both naturally occurring and man-made, for example: (1) sea water (liquids of higher osmotic pressure) fed by fresh water (liquid of lower osmotic pressure) from a river; (2) brines or salt water bodies (e.g., The Dead Sea) fed by sea water or fresh water; (3) water "fueled" by artificially adding salt therein from naturally occuring salt bodies (to produce the higher osmotic pressure liquid) interacting with water not so "fueled"; and (4) evaporation ponds (both naturally occurring and man-made) interacting with lower-concentration water solutions.

Theoretically all of the foregoing sources could be used for generating power in accordance with the invention, as will be more fully described below. Osmotic power stations utilizing power sources such as the Dead Sea would even appear to be competitive in price with present power stations. When evaporation ponds are used, they can be man-made at any desired location to convert the solar energy at that location to mechanical or electrical power. They can be used therefore instead of the known solar energy converters (e.g., photovoltaic devices and solar heat collectors including heat engines) and would even appear to be substantially less expensive than such devices.

## BRIEF SUMMARY OF THE INVENTION

The present invention employs Pressure-Retarded-Osmosis (PRO) for the production of useful energy from the above naturally available energy sources. The invention also enables the PRO process to be employed where the above natural energy sources are not available. A number of both types of systems are described below for purposes of example.

The basic Pressure-Retarded-Osmosis (PRO) method and apparatus for generating power is generally characterized by an arrangement for performing the steps of: applying a hydraulic pressure to a first liquid of a first osmotic pressure and introducing same into a first pathway which is at least partially defined by one face of a semipermeable membrane; introducing a second liquid having a lower hydraulic pressure and a lower osmotic pressure into a second pathway which is at least partially defined by the opposite face of the membrane; maintaining the hydraulic pressure difference between liquids on the opposite faces of the membrane at a pressure difference which is less than the osmotic pressure difference between the liquids, at every point in the two pathways, thus effecting by Pressure-Retarded-Osmosis a passage of at least part of the second liquid through the semipermeable membrane, forming a pressurized mixed solution of greater volume than said first liquid introduced into the first pathway; and converting the potential energy stored in the pressurized mixed solution to useful energy, such as electrical or mechanical energy.



The semipermeable membrane used is preferably one that passes only solvent and no solute (perfectly semipermeable), but it may also be an imperfect one, i.e., one that passes some solute.

The first part (Part **1**) of the description below (particularly FIGS. **1–6**) concerns the basic technique as it may be applied with respect to each of the four naturally-occurring and man-made water sources mentioned above. The second part (Part 2) of the description below (particularly FIGS. **7–14**) concerns the manner of enabling such PRO power plants to be applicable for the production of power where the above natural energy sources are not economically available.

One limitation in the systems described in Part 1 (FIGS. **1–6**) is that such systems can be used, as a practical matter, only in certain areas of the World. the only osmotic power plant scheme described in Part 1 below which can be considered to approach general applicability for production of economic power is that of FIG. **6**, in which water evaporation ponds are used for re-concentration of the mixed solution, i.e., brine diluted in the Pressure-Retarded-Osmosis (PRO) unit. However, as will be discussed in Part 2, the area required for evaporation or distillation would be in the order of 1,000 square kilometers for a 1,000 megawatt power plant utilizing evaporation ponds. This area is so large that it would restrict the use of osmotic power plants to areas which in effect already possess evaporation ponds, such as the Dead Sea and Great Salt Lake.

Another limitation on the general use of the osmotic power plants described in Part **1** (FIGS. **1–6**) is that an expendable source of a low-osmotic pressure aqueous solution, such as ocean or river water, must be available in the region of the evaporation ponds. Large quantities of this solution are expended. For example, in the 1,000 megawatt plant using evaporation ponds described below in FIG. **6**, a daily quantity of 4,300,000 cubic meters of water of low osmotic pressure aqueous solution must be expended. Even if large quantities of such solutions are available, such as in sea water, charges for pumping, filtering, etc., may be exorbitant.

Part 2 of the description below (FIGS. **7–14**) therefore describes systems for making the basic PRO techniques of general applicability, by eliminating both of the above limitations, namely by enabling the solar energy collecting area to be greatly reduced, and by obviating the need of expendable material altogether.

The latter aspects of the invention are accomplished by providing arrangements which, after the potential energy stored in the pressurized mixed solution is converted to useful energy such as elecrical or mechanical energy, the following additional steps are performed: recover the first and second liquids by separating from the mixed solution a quantity of second liquid substantially equal to the quantity which passed through the membrane and mixed with the first liquid; restore substantially the original temperatures to the recovered first and second liquids; reapply the above mentioned hydraulic pressure difference between the recovered first and second liquids; and recycle the recovered first and second liquids through the first and second pathways respectively.

## BRIEF DESCRIPTION OF THE DRAWINGS

The invention is herein described, somewhat diagrammatically and by way of example only, with reference to the accompanying drawings, wherein:

FIG. 1 illustrates the basic osmosis process for the case of sea water and fresh water on opposite sides of a semi-permeable menbrane;

FIGS. 2*a* and 2*b* illustrate the basic pressure-retarded osmosis process on which the present invention is founded;

FIG. 3 illustrates a continuous pressure-retarded osmosis process for producing energy from sea water at the mouth of a river, and FIG. 3*a* is a pressure-energy diagram relating thereto;

FIG. 4 illustrates a continuous pressure-retarded osmosis process for producing energy from Dead Sea brine at the mouth of the River Jordan, and FIG. 4*a* is a pressure-energy diagram relating thereto;

FIG. 5 illustrates the pressure-retarded osmosis process for the production of energy in a multi-stage system, and FIG. 5*a* is a pressure-energy diagram relating thereto;

FIG. 6 illustrates the pressure-retarded osmosis process for the production of energy from concentrated brine provided by evaporation ponds.

FIG. 7 is a block diagram illustrating the generalized concept of a PRO heat engine for generating power.

FIG. 8 is a block diagram of a PRO heat engine constructed in accordance with the Carnot Cycle, utilizing solar energy as a heat source;

FIG. 9 is a block diagram illustrating a portion of a PRO heat engine constructed in accordance with the invention, FIG. 3 being essentially the Pressure-Retarded-Osmosis (PRO) section described in FIGS. 2–6 above;

FIG. 10 is a block diagram illustrating a PRO heat engine using distillation in the thermal unmixing section;

FIG. 11 is a block diagram illustrating a PRO heat engine with a distillation plant using dichlorodifluoromethane as the solvent;

FIG. 12*a* illustrates a methanol-hexane binary liquid system whose miscibility is a function of temperature, this system being useful in effecting the thermal separation of a mixed solution containing these two species into a diluted and a concentrated solution respectively;

FIG. 12*b* illustrates a triethylamine-water binary liquid system whose miscibility is also a function of temperature, which system could therefore be useful in effecting the thermal separation;

FIG. 13 is a block diagram illustrating a PRO heat engine using a FIG. 12*a* type binary liquid circulation system (methanol-hexane); and

FIG. 14 is a block diagram of a PRO heat engine using a solution containing a solute whose solubility is a function of temperature.

## PART 1 - BASIC PRO PROCESS (FIGS. 1–6)

### Ordinary Osmosis (FIG. 1)

Referring to FIG. 1 illustrating the ordinary osmosic process, there is shown a vessel 2 divided into two chambers 4, 6 by means of a semipermeable membrane 8. Chamber 4 contains fresh water having zero osmotic pressure and also zero atmospheric pressure. Chamber 6 contains sea water having an osmotic pressure of 25 atmospheres and a hydraulic pressure of zero atmospheres.

Water permeates through the membrane 8 from the fresh water side 4 to the sea water side 6 because of the osmotic pressure differences of $\Delta\pi = 25 - 0 = 25$ atmospheres. The flux, J, of the permeant water, is given by:

$$J = A\,\Delta\,\pi \qquad \text{(Eq. 1)}$$

where:

J is the flux of permeant, i.e., flow rate per unit area (a convenient flux unit is cubic meters per day and square meter $m^3/m^2$ day);

$\Delta\pi$ is the osmotic pressure difference (atmospheres) across the membrane, and is 25 for the case of sea water-fresh water;

A is a constant which depends on membrane properties.

## PRESSURE-RETARDED OSMOSIS (PRO) - FIGS. 2*a* and 2*b*

Now, as shown in FIG. 2*a*, let a hydraulic pressure, P, less than 25 atmospheres, be applied on the sea water side 6, i.e., let work be done on the sea water. Fresh water will still permeate to the sea water, i.e., against the hydraulic pressure gradient, but at a reduced rate given by:

$$J' = A\,(\Delta\pi - P) \qquad \text{(Eq. 2)}$$

where $(\Delta\pi - P)$ is the effective driving force, and represents a more general statement of effective driving force than does Equation 1.

If the process is permitted to continue, the volume under a pressure, P, will increase on the sea water side 6 (FIG. 2*b*). The total brine thus compressed has a potential for furnishing mechanical energy by passage through a water turbine. the energy furnished will exceed the energy (work) originally done on the sea water by a fraction $\Delta V/V$ where $\Delta V$ is the volume of permeant which has passed through the membrane and V is the original volume of sea water. The excess mechanical energy available is given by $P\Delta V$.

The general process of permeating against a hydraulic pressure gradient is called "Pressure-Retarded-Osmosis" (PRO). In the above example this is a transient process and would ultimately stop because of dilution of sea water by the permeant. However, the process could be carried out continuously, and several techniques for continuous conversion are described below with respect to different energy sources.

## PRO PROCESS APPLIED TO SEA WATER AT MOUTH OF RIVER - FIGS. 3, 3*a*

FIG. 3 illustrates a continuous pressure-retarded process for the production of energy from sea water at the mouth of a river, and FIG. 3*a* is the related pressure-energy diagram. As will be shown, large quantities of useful energy can be continuously extracted from sea water close to the mouth of rivers.

In FIG. 3 the pressure-retarded osmosis (PRO) apparatus is generally designated 10, and the semipermeable membranes are generally designated 12. The membranes define a first pathway (arrow 14) for the liquid of higher osmotic pressure, e.g., sea water having an osmotic pressure of 25 atmospheres. The hydraulic pressure of the sea water is raised by pump **16** before it is introduced into pathway **14** of the PRO apparatus. The sea water exits from the apparatus into a hydroturbine **17** which converts the energy gained by it into electrical energy. The river water is inletted at the opposite side of the PRO apparatus so as to flow through path **18** in (preferably) counter-flow to path **14** of the sea water, and in contact with the opposite face of the semipermeable membranes **12.** The river water is maintained at essentially zero pressure.

Pressure-retarded osmosis is thus effected through the semipermeable membranes **12** in the direction of arrow **19,** the permeant thereby increasing the volume of the sea water in pathway **14.** The sea water in that path thus gains energy which is in excess of that expended in pressurizing the sea water by pump **16,** and this energy is converted to electrical power by hydroturbine **17.**

As one example, 1 cubic meter of sea water (as a basis) at zero atmospheres hydraulic pressure and 25 atmospheres osmotic pressure (Point A in FIGS. **3** and **3***a*) is compressed by pump **16** to 10 atmospheres hydraulic pressure (Point B), thus absorbing mechanical energy equal to (1) (10) = 10 cubic meter − atmospheres ($m^3$ atm) or 0.28 kilowatt hours (KWH) (Area ABEF in FIG. **3***a*). (For the sake of simplicity in illustrating this process no allowance is made in the discussion for inefficiencies in mechanical equipment, frictional losses in fluid flow, etc.) The sea water is then passed through pathway **14** of the PRO apparatus **10** at the hydraulic pressure of **10** atmospheres in counterflow to river water at zero hydraulic pressure flowing through pathway **18** on the other side of the membranes **12.** The sea water absorbs 0.6 $m^3$ of permeant through the membranes. Thus 1.6 $m^3$ of diluted brine leave the PRO apparatus at a hydraulic pressure of 10 atmospheres (Point C).

As the sea water passes through hydroturbine generator **17,** its hydraulic pressure is released to zero (Point D) in the process of delivering (1.6) (10) = 16 $m^3$atm or 0.45 KWH of energy (Area CDEF in FIG. **3***a*). The net energy delivery for the 0.6 cubic meters of permeant is (0.6) (10) = 6 $m^3$ atm (Area ABCD in FIG. **3***a*). On the basis of one cubic meter of permeant; 10 $m^3$ atm or 0.28 KWH are deliviered, i.e., the energy/permeant ratio is 0.28 KWH/$m^3$ permeant.

The possible power available is limited, in principle, only by the amount of water available from the river source for permeation through the membranes. For example, the Mississipi river delivers 56,000,000 cubic meters of water each day to the Gulf of Mexico. If it is assumed that ¾ of this is permeated through PRO stacks the daily energy available would be 12,000,000 KWH, i.e., a power output of about 500 megawatts.

The pump may be driven directly with the hydroturbine as shown by the broken line mechanical connection **15** in FIG. **3,** instead of having a separate electric motor for the pump. Such direct coupling might reduce capital costs and also reduce sources of energy loss due to electric motor inefficiency.

The most attractive feature of this application of PRO (and some subsequent ones to be described) is the fact that the "fuel" system sea water - river, is provided without man-made apparatus, i.e., the solar energy is converted to a (slightly) high osmotic pressure solution by natural means. Unfortunately the capital cost of PRO plants for energy delivered by the use of the sea water-river water combination would usually be excessive for two reasons, both traceable to the relatively low osmotic pressure of sea water:

First, the energy/permeant ratio, kilowatt hours per cubic meter of permeant, is too low, being only 0.28 KWH/$m^3$. Second the capital cost of the PRO apparatus would be too high. Present equipment for reverse osmosis, another osmotic precess utilizing appreciable hydraulic pressures, will cost in the order of 60 dollars per daily cubic meter of water permeated ($D/$M^3$), but at effective driving forces ($\Delta\Pi - P$) in the order of 40 atmospheres. However in the sea water-river case the average effective driving force would be in the order of 10 atmospheres (see FIG. **3**).

This low value of effective driving would given an undesirably low value of the Flux, J, and thus increase the capital cost of the membrane equipment. Therefore, it is assumed that the ratio of the capital cost ($) to permeant flow rate, $m^3/_d$ would be 150 $D/$m^3$ for the sea water-river water case. Based on the above values the capital cost per kilowatt can be calculated:

$$(150)\left(\frac{1}{.28}\right)(24) = 13000\ \frac{\$}{KW}$$

where 24 is hours per day.

This cost is far too high when compared with present power plant capital costs in the order of 200 $/KW.

From the above calculation on the use of PRO with sea water it is clear that to make PRO economical as an energy conversion method, the energy/permeant ratio must be increased and/or the capital cost/permeant flow rate ratio must be decreased. Both of these results can be accomplished by increasing the hydraulic pressure, but the water-receiving solution must also have a correspondingly higher osmotic pressure, i.e., greater than the hydraulic pressure, as required in PRO. For the satisfaction of this requirement, naturally available brines, in strongly saline bodies of water such as the Dead Sea, are ideal if a lower osmotic pressure solution is also available such as sea water, brackish springs, or a river draining into the saline body.

## PRO PROCESS APPLIED TO DEAD SEA BRINE AT MOUTH OF JORDAN -FIGS. 4,4*a*

FIGS. **4** and **4***a* show an economical scheme for using PRO together with such an osmotic sink solution, as can be provided by the Dead Sea and other naturally salty bodies of water. FIG. **4** illustrates the flow diagram, and FIG. **4***a* illustrates the pressure-energy diagram.

In FIG. **4,** the pressure-retarded osmosis apparatus is generally designated **20,** and includes the semipermeable membranes **22** dividing the apparatus into a first flow pathway **24** and a second (preferably) counterflow pathway **28,** each pathway being partially defined by the opposite faces of the semipermeable membranes **22.** The first pathway **24** is for the Dead Sea brine whose hydraulic pressure is raised by pump **26** before being introduced into the inlet of that pathway. The outlet of pathway **24** leads to a hydroturbine **27** which converts the energy gained by the first liquid to electrical power. The second pathway **28** is for the lower osmotic pressure liquid supplied for example by the Jordan River. The permeant from the latter water pathway passes through membranes **22** by pressure-retarded osmosis, as shown by arrow **29**, thereby increasing the volume of the liquid flowing through pathway **24**, producing therein a quantity of energy which is in excess of that expended by pump **26**.

As one example, 1 cubic meter of Dead Sea brine ($\pi = 940$ atm) at zero pressure gauge (Point A in FIGS. 4 and 4*a*) is compressed to a hydraulic pressure of 200 atmospheres (Point B) after which it is passed through the PRO apparatus **20** at this pressure in counterflow to the Jordan River water at zero hydraulic pressure on the other side of the membrane. (In the calculations on Dead Sea brine, it is assumed that the osmotic pressure-concentration relations will be the same as for magnesium chloride). Each cubic meter of Dead Sea brine receives 0.8 cubic meters of permeant **29** at 200 atmospheres pressure (Point C) after which the diluted solution ($\pi = 515$ atm) passes through the hydroturbine generator **27** where its hydraulic pressure is reduced to zero (Point D) in delivering a net energy output for the 0.8 $m^3$ of permeant, of 160 $m^3$ atm (Areas ABCD in FIG. 4*a*). On the basis of one cubic meter of permeant 200$m^3$ atm or 5.6KWH are delivered, i.e., the energy/-permeant ratio is now 5.6 KWH/$m^3$ permeant.

As can be seen in FIG. 4 the effective driving force, ($\Delta\pi - P$), at the Dead Sea Brine and Jordan River inlets of the PRO unit are 738 and 114.5 atmospheres respectively. Because of these high effective driving forces, which will give high flux values for permeant, the capital cost of the PRO unit should now be much lower than in the sea water case where effective driving forces were very low. It is assumed that for Dead Sea Brine use the cost will be 60 $/$m^3$ of permeant. This is in the order of magnitude of present reverse osmosis equipment capital costs.

Thus capital cost per kilowatt hour now becomes:

$$(60)\cdot\left(\frac{1}{5.6}\right)(24) = 240\ \$/KW$$

To this cost must be added the cost of the hydroturbine generator, estimated at 75 $/KW. The total installed cost will then be 315 $/KW. This figure is low enough to merit comparison of PRO with existing power plants. Assuming that the capital cost is paid for at 8 percent per annum, the contribution of the capital cost to the cost of the power is:

$$\frac{(.08)(315)}{(365)(24)} = 0.003\ \$/KW$$

From the foregoing analysis it is clear that, generally speaking, the higher the hydraulic pressure the cheaper will be the capital cost per kilowatt. However, as the hydraulic pressure is increased there must be, as can be seen from considering FIG. 4*a*, a simultaneous decrease in the permissible ratio of diluted Dead Sea Brine to Dead Sea Brine entering the apparatus. High values of this ratio are desirable to allow for hydraulic pressure losses due to friction in the system and for the fact that the pump and hydroturbine have efficiencies less than 100 percent. (Incidentally, it can be demonstrated that for positive net energy delivery from the system the ratio of diluted to entering brine must exceed 1/(hydraulic efficiency) (pump efficiency) (Hydroturbine efficiency) where hydraulic efficiency is the ratio of the hydraulic pressure entering the hydroturbine to the hydraulic pressure leaving the pump).

These two apparently mutually exclusive requirements, high hydraulic pressure and high ratio of diluted to entering Dead Sea Brine, can both be met by staging.

## PRO 3-STAGE PROCESS - FIGS. 5, 5*a*

FIGS. 5 and 5*a* show a possible 3-stage unit. Each stage consists of one of the pressure-retarded osmosis apparatuses described above (these being designated **30, 40** and **50**, respectively) and a hydroturbine (**37**, 47, 57) at the output end of the higher osmotic pressure liquid pathway. A pump **36** at the inlet end of the first stage raises the hydraulic pressure of the high osmotic pressure liquid (e.g., Dead Sea Brine) to a very high/hydraulic pressure, for example 350 atmospheres. This pressure is reduced to about 200 atmospheres in hydroturbine **37** at the end of the first stage, and is inletted at this pressure into the second stage **40.** The hydroturbine **47** at the outlet of the second stage drops by pressure further, for example to about 108 atmospheres, before the liquid is introduced into the third stage **50,** dropping to 0 atmospheres at the outlet of hydroturbine **57** of the third stage.

The low osmotic pressure liquid pathway is not illustrated in FIG. 5, but it will be appreciated that it would be as described in the previous examples.

Thus, in each stage water from the low osmotic pressure liquid permeates into the high osmotic pressure liquid, causing the latter to gain energy; while the hydraulic pressure of the high osmotic pressure liquid is very high at the first stage, and is successively lowered while energy is delivered by the respective hydroturbine. By this means, hydraulic pressures as high as 350 atmospheres may be utilized, and the final ratio of diluted to entering Dead Sea Brine may be about 2.2, as shown.

In some parts of the World solid salt is stored in large quantities on or below the surface. This salt could be used as "fuel" for energy conversion in accordance with the invention by adding the salt to the water to produce the higher osmotic pressure liquid. In this case the costs would be greater than for the previous processes, because of the cost of dissolving the salt, and also the cost of disposing the diluted salt solution. Nevertheless, this use of salt as "fuel" could very well still be commercially feasible in some locations.

In the methods described above, naturally occurring bodies of water or salt or used for producing the power in accordance with the invention. The invention, however, may also be exploited by artificial or man-made means, by the use of artificial solar or evaporation ponds which collect the solar energy and produce the high osmotic pressure solution.

## PRO PROCESS APPLIED TO EVAPORATION PONDS - FIG. 6

FIG. 6 illustrates the flow diagram of a process in accordance with the invention applied to brines of Dead Sea concentrations produced by evaporation ponds. In this case, the pressure-retarded osmosis apparatus **60** is fed by concentrated brine from the outlet end of an evaporation pond **70,** this brine being first pressurized, e.g., to about 200 atmospheres by pump **66.** The diluted brine exiting from apparatus **60** is passed through hydroturbine **67,** where its hydraulic pressure drops to **0,** and is then introduced into the inlet and of the evaporation pond **70.** The lower osmotic pressure solution may be river water which is concentrated in the apparatus, after which it may be discharged or salvaged for other use.

The new feature provided by the evaporation pond system of FIG. **6** amounts to passing from point D to point A in the diagram of FIG. 4*a*, and this will add somewhat to the costs of the process. The unit cost of the evaporation pond is estimated at 100 dollars per daily cubic of water evaporated. The cost of the evaporation ponds per kilowatt hour will then be:

$$(100)\left(\frac{1}{5.6}\right)(24) = 428\ \$/KW$$

The total capital cost per kilowatt hour will be: 315 + 428 = 743 $/KW

The above capital cost, although considerably more expensive than the cost of presently built power plants, is much cheaper than the estimated cost of solar energy power plants.

The efficiency of the process from the standpoint of solar energy conversion, can be calculated from the scheme using evaporation ponds. In this scheme 5.6 KWH are delivered per cubic meter of water permeated. In gram calories this is $(4.8)(10)^6$ gm Cal/m$^3$. To evaporate a cubic meter of water requires $(580)(10)^6$ gm Cal/m$^3$ of solar energy. Therefore the thermal efficiency of the process is $(4.8/\mathbf{580})(100) = 0.83$ percent. The thermal efficiency could be improved by operating at a hydraulic pressure much closer to the osmotic pressure, but this would require a very low ratio of diluted Dead Sea Brine to Dead Sea Brine entering the PRO apparatus.

In the above described processes, the low osmotic pressure solution is concentrated. Thus, the process could also be used not only for the production of power, but also for the recovery of concentrated solution or solutes.

Also, while the energy gained by the higher osmotic pressure liquid is converted to power by passing same through a hydroturbine generator, it will be appreciated that this energy can be stored in the liquid itself, e.g., by using it for pumping same to a higher elevation until needed, or until subsequently used for generating power.

## PART 2 - PRO PROCESS OF MORE GENERAL APPLICABILITY (FIGS. 7–14)

### PRO HEAT ENGINE, GENERALIZED CONCEPT (FIG. 7)

As indicated earlier, the systems shown in FIGS. **3–6** to illustrate the basic PRO process would be of limited applicability because of the need for large solar energy collecting surfaces and the need for expendable material. Now described are systems which eliminate both of the above limitations and therefore make the PRO process of more general applicability.

The new system, as shown in its most generalized form in FIG. **7,** consists of two sections. First, there is the pressure-retarded osmosi (PRO) section **102,** which delivers work available from the free energy decrease occurring during mixing of a diluted and a concentrated solution. Second, there is the thermal unmixing section **104,** in which heat absorption and rejection supplies energy for free energy recovery, i.e. unmixing



and temperature restoral to the two solutions after which they are returned via lines **106, 108** to the PRO section **102,** thus completing the cycle. The thermal unmixing process can involve intermediate gaseous or solid phases in the course of restoring the original solutions or can be conducted so that the original liquid solutions are directly produced.

The above description fits in every respect the definition of a heat engine, i.e., a man-made device which makes it possible for a working substance to undergo a cyclic process in the conversion of heat into work. Thus the invention is believed to represent the first use of any osmotically-operated power process as the work producing component of a heat engine. The invention is also believed to represent the first use of pressure-retarded osmosis as the work-producing component of a heat engine.

The heat source, diagrammatically indicated by arrow **110** in FIG. **7,** for effecting the thermal unmixing in section **104,** could theoretically be any energy source such as fossil fuels or nuclear energy. However, because of the dwindling energy resources from fossil fuel, only solar energy will hereinafter be mentioned.

With regard to solar energy collectors, several promising types exist. These include solar ponds, i.e., ponds in which the solar energy maintains a gradation in temperature by means of a gradation in salt concentration, and solar heating devices such as those incorporating selective coatings for trapping solar energy in the form of heat. With both of these types the heat can be extracted by the passage through the unit of a heat-transfer fluid. Solar distillation plants are also energy collectors. In addition to this service they also perform the unmixing functions.

The heat sink into which the heat from the thermal unmixing section **104** is rejected as indicated by arrow **112,** is preferably of the liquid type such as sea water or river water. However ambient air can also serve as an appropriate heat sink. It is assumed herein that any of these potential sinks can provide a heat rejecting temperatures not exceeding 25°C.

In FIG. **7,** the net work output from the PRO section **102** is indicated by arrow **114** and block **116,** the mechanical and hydraulic losses from the PRO section being passed to the heat sink as indicated by arrow **118.**

### PRO HEAT ENGINE AS CARNOT ENGINE (FIG. 8)

FIG. **8,** in which the parts corresponding to FIG. **7** are correspondingly numbered, shows an idealized PRO heat engine operating between temperature limits of 100° and 25° centigrade. It is assumed that its heat source is a solar energy collector **120** which loses (arrow **123**) half the solar energy incident upon it (arrow **121**). This PRO engine is subject to the maximum efficiency imposed by the law of Carnot:

$$E_{max} = 100\left(1 - \frac{273 + 25}{273 + 100}\right) = (1 - .8)(100) = 20\%$$

where 273° is absolute temperature at 0° centigrade.

Based on this limitation and the above assumptions it is seen that, of 100 units of energy incident on the solar energy collector, a maximum of 10 units can be obtained as useful work. The useful work obtained from an actual PRO engine must be measured against this criterion, i.e., the useful work from the PRO engine will always be less than, but should approach as closely as possible to, 10 percent of the energy incident on the solar energy collector.

Because the PRO heat engine obtains its work from the free energy decrease during mixing of the solutions, the maximum work obtainable, i.e., the Carnot work, is equal in magnitude to this decrease in free energy.

Following are a number of important advantages of the PRO heat engine:

1. By making pressure retarded osmosis the work-producing process of a heat engine, the working liquids undergo a cyclic process and are therefore not expended at all. The production of useful energy by PRO becomes of general applicability, requiring only the availability of solar or other energy source and no material resource expenditure. This extends the usefulness of the technique of the above-cited patent specification in which applicability of the systems therein described as examples was limited to locations having an expendable source of low osmotic pressure solution such as river water, brackish water, or sea water.

2. The PRO heat engine can be efficient in the conversion of energy to work. As will be shown, the thermal efficiency of the PRO engine can be in the order of at least 60 percent of that thermodynamically possible. This means that when operating between the limitations of the previous section 100° to 25° centrigrade, and with 50 percent losses in the solar energy, collector, the PRO heat engine will convert $(0.6)(10) = 6$ percent of the incident solar energy to useful work.

This thermal efficiency will enable the solar collector area to be drastically reduced in comparison to what is required with the evaporation ponds described as examples in the above-cited patent application. For a 1,000 megawatt PRO heat engine this area need be only in the order of 60–70 kilometers as compared with about 1,000 square kilometers for evaporation ponds.

3. The PRO heat engine possesses practical advantages over existing heat engines. Most existing heat engines utilize the vapor power cycle. In such a cycle, a working fluid under pressure is vaporized by addition of heat from a high temperature heat source. It then does work of expansion in a turbine or engine, after which it is condensed by heat removal to a low temperature heat sink. The liquid is compressed to the original pressure, thus completing the cycle.

Because of the phase changes occurring in the vapor power cycle, it must be conducted within limits which add to the cost of necessary equipment. For example the expansion of saturated steam in a heat engine is accompanied by partial condensation to liquid water. However, this liquid content cannot exceed 10 or 12 percent because of excessive wear on turbine blades or engine pistons. Thus the expansion of the steam is limited. This limitation can be minimized by super-heating the steam prior to expansion, but this adds to the cost of the equipment, and for the low heat source temperatures considered herein of 100°C, no additional efficiency is gained.

An additional limitation of the vapor power cycle using steam lies in the fact that all the vapor must be finally condensed. This means that at the heat rejection temperature of 25°C, a vacuum must be maintained in the condenser, including means for continuous removal of air which might leak in through pump seals, etc.

The PRO heat engine overcomes the above limitations of the vapor power cycle because no vapor exists during the power production part of the cycle (and indeed need not exist in any part of the cycle, as will be seen).

4. Another very important advantage of the PRO heat engine is that it provides a liquid under high hydraulic pressure as the energy producing fluid. By this means the engine can utilize a high pressure hydroturbine. The work is used here to mean any liquid-driven turbine, in contrast to a steam turbine or other turbine driven by gas or vapors other than a steam turbine. Hydroturbines are more efficient than steam turbines or other vapor turbines. Furthermore, a hydroturbine is inherently safer since at high pressures much less energy is stored in a liquid than in a gas.

## PRO SECTION OF HEAT ENGINE (FIG. 9)

FIG. 9 illustrates the PRO section **102** (of FIG. **8**) of the heat engine; FIGS. **10–14** (described below) illustrate different arrangements which may be used for the thermal unmixing section **104** (of FIG. **8**).

The pressure-retarded osmosis (PRO) section, consisting of a pump **122**, the membrane unit **124**, and a hydroturbine **126**, is shown in FIG. **9** as it would operate under ideal conditions. A concentrated solution, by which is meant one having a high osmotic pressure ($\pi_{high}$), and having a volume of V cubic meters ($m^3$) is pressurized by pump **122** to a hydraulic pressure P atmospheres (atm) requiring a work input of PV cubic meter atmospheres ($m^3$atm), after which it is pumped via line **127** into the high pressure side of the membrane unit **124.** Simultaneously a diluted solution, by which is meant one having a low osmotic pressure, ($\pi_{low}$), and having a volume of $\Delta V$ $m^3$ is pumped (by a pump not shown) via line **128** into the low hydraulic pressure side of the membrane unit **124.** The diluted solution permeates through the membranes against the hydraulic pressure P because it is arranged that everywhere in the unit $P>\Delta P$ where $\Delta P$ is the osmotic pressure difference (atm) between the solutions on each side of the membrane. This is the fundamental principle of pressure-retarded osmosis, as described above.

A volume $(V + \Delta V)$ $m^3$ of mixed solution is sent to hydroturbine **126** at the pressure P atm. Thus the hydroturbine delivers $P\ (V + \Delta V)$ $m^3$ atm of work (via connection **129**) in the course of reducing the pressure of the mixed solution of zero. The net output of work is equal to the difference between the output from the hydroturbine and the input to the pump, i.e., the net work is $(P\Delta V)(m^3)$ atm.

It is important to understand that net work is obtained only from $\Delta V$, the volume of permeant liquid passing through the membranes. In order to minimize the size of the membrane unit it may be stated as a first guideline:

Guideline **1**: the ratio should be maximized of net work delivered to volume of liquid passed through the membranes.

This is accomplished by using a high hydraulic pressure. However P must be less than $\Delta P$ everywhere in the unit, as described above, and the minimum $\Delta P$ occurs between the diluted solution and the mixed solution. Therefore it follows as a corollary to Guideline **1** that the osmotic pressure difference between the mixed solution and the diluted solution should be high.

It should be realized that Guideline **1** is also appropriate for the thermal unmixing section (**104,** FIGS. **7** and **8**). If the ratio is high of net work delivered to volume of permeate passed through the membranes, then less mixed solution must be separated by the thermal unmixing techniques.

Thermal unmixing techniques may be divided into several categories depending on the nature of the intermediate phases employed in the unmixing. Thus it is possible to utilize vapor and solid intermediate phase as well as to divide the mixed solution directly into diluted and concentrated solutions.

The best thermal unmixing technique would seem to be that which minimizes the thermal energy input requirement. A guideline for carrying out this requirement can be cleaned from **FIG. 8.** Since this is a Carnot cycle, the thermal efficiency is maximized. Therefore the ratio of thermal energy input to net work is minimized. Since the net work is equal to the free energy decrease, it is possible to say that the ratio of thermal energy input to free energy decrease is also minimized. This value is 50/**10** = 5 for the temperature chosen. In any actual plant, the ratio will be higher but this value of five may be considered as a target at which to aim.

Therefore it may be stated as a second guideline:

Guideline 2: Within thermodynamic limitations, the ratio of thermal energy input to free energy decrease should be minimized in the PRO heat engine.

THERMAL UNMIXING BY DISTILLATION (FIGS. **10** and **11**)

The use of distillation as the thermal unmixing process is shown in general terms in the heat engine of FIG. **10**, parts corresponding to those in FIG. **9**, being correspondingly numbered. The distillation plant **130**, divides the unpressurized mixed solution from input **132** into a first output **134** of V $m^3$ unpressurized concentrated solution having a high value of osmotic pressure ($\pi_{high}$), and a second output stream **136** of a diluted solution in the form of a vapor. The concentrated solution in stream **134** is joined to the PRO section of FIG. **9** at the pump **122**. Meanwhile, the diluted solution vapor in stream **136** is condensed in condenser **138** and cooled, thus rejecting part of the incoming thermal energy to the heat sink as shown by arrow **140**, after which the diluted solution is joined via line **128** to the PRO section at the low pressure side of the membrane unit. The remainder of the cycle is as described above in the discussion on the PRO section, as a result of which the unpressurized mixed solution of FIG. **9** is sent to the distillation plant for thermal unmixing, and the cycle completed. The heat for the distillation plant is supplied from a solar energy collector **142.**

Water is a poor choice as a solvent in a distillation plant in the thermal unmixing section of a PRO heat engine. Specifically it fails badly to meet Guidline 2 because of its very high volumetric latent of vaporization of 580 cal/cm$^3$. Even with concentrated salt solutions for which the free energy of separation will be in the order of 5 cal/cm$^3$, the ratio of heat input to free energy of separation will be in the order of 580/5 = 120. This factor would increase the solar collector area requirement for a 1000 megawatt plant from the theoretical minimum value of about 4 km$^2$ (assuming an average insolation flux of 520 calories per day on each square centimeter) to 500 km$^2$, if water were used in a simple distillation process such as solar distillation. If it is more realistically assumed that the solar collector efficiency is about 50 percent, 1000 km$^2$ of solar collector area would be required. Other losses might increase the area requirement further.

The second guideline may in principle be approached more closely in distillation plant by employing the use of "heat multiplying" distillation plants such as multiple effect or multistage flash distillation plants. In these the arrangement is such that one kilogram of steam (or its thermal equivalent) entering from the heat source is capable of vaporizing, say 10 pounds of steam, thus increasing the efficiency of the process 10-fold. However there is a limitation on the heat multiplying capability. In each effect or stage of the plant, the vapor condenses at the boiling point of pure water characteristic of the pressure in the effect. However the concentrated brine in the same effect boils at a higher temperature because of its salt content. This phenomenon in known as boiling point elevation (BPE). Its effect is cumulative from effect-to-effect and the possible heat multiplying capability of the plant is limited by the available source-to-sink temperatures such that:

$$\text{Max. dist. plant heat multiplying capability} = \frac{\text{Source temp.} - \text{Sink. temp.}}{\text{BPE}}$$

Now we know that boiling point elevation increases directly with osmotic pressure difference such that we may say:

$$\text{Max. dist. plant heat multiplying capability} = \frac{(\text{Source temp.}-\text{sink temp.})}{\text{Osmotic pressure difference}}$$

However it was stated as a corollary to the first design guideline that the osmotic pressure difference between the mixed solution and the diluted solution should be as high as possible. The complete implementation of this guideline would reduce the heat multiplying capability of the plant drastically. This capability is further reduced by the fact that for solar energy heat sources discussed herein, the difference between source and sink temperature is low. Therefore a distillation plant to be described subsequently is not of the heat multiplying type, and it is explained why such a plant would not be feasible with solar collectors as the energy source. However it is recognized that heat-multiplying distillation plants might be efficacious under other conditions.

The disadvantage of water as the solvent phase in a PRO heat engine may be overcome by using a solvent with a low value of volumetric latent of vaporization. Virtually all liquids have a volumetric latent heat of vaporization lower than water. Some of these such as the halogenated organic compounds, and especially those containing fluorine, are exceptional in this regard. For example dichlorodifluoromethane $C\,Cl_2F_2$, a commerically available refrigerant ("Freon-12") has a volumetric latent heat of vaporization of only 44 cal/cm$^3$. By the use of such materials Guideline **2** is approached more closely than with water. From a practical standpoint the use of Freon 12 means that the solar energy collecting area can be an order of magnitude less in area than if water is used, all other things being the same.

FIG. **11** (parts corresponding to FIG. **10** being correspondingly numbered) illustrates a PRO heat engine with a distillation plant using $C\,Cl_2F_2$, dichlorodifluoromethane, as the solvent in the dilute solution. For this calculation the following conditions and/or assumptions were utilized:

1. A relatively non-volatile solute of molecular weight **46** was used; e.g., Ethanol meets this requirement.

2. The osmotic pressure of the mixed solution was 275 atmospheres. This value is sufficiently high that a high hydraulic pressure in the order of 255 atm can be used, according to Guideline 1.

3. The efficiencies of the solar collector, the pressurizing pump, and the hydroturbine were assumed to be 50 percent (thermal), 95 percent (mechanical), and 95 percent (mechanical), respectively.

4. It was assumed that there was a 2 atmosphere pressure drop due to hydraulic friction in any unit of the apparatus.

5. Raoult's Law was assumed to apply to all solutions.

6. The system was operated so that 1 volume of concentrated solution was diluted with 0.4 volume of solvent, $C\ Cl_2F_2$.

7. The density of all solutions was assumed to be one.

8. The temperature of the condensate leaving the condenser was 25°C.

Based on these assumptions, the operating conditions of FIG. **11** were obtained. Of special interest is the requirement that the concentrated solution have a temperature of 94°C, as compared with the condensate temperature of 25°C. The temperature of 94°C is due chiefly to two requirements: first, the mol fraction of $CCl_2F_2$ must be low in the concentrated solution in order to provide a high enough osmotic pressure to the concentrated solution, second, the partial pressure of $CCl_2F_2$ in the concentrated solution must be at least equal to the partial pressure of pure $CCl_2F_2$ in the condensate at 25°C in order to condensate this solvent.

Even with the use of a single-effect distillation plant, the temperature of 94°C is approaching the assumed upper temperature limit from a solar collector. For this reason it is clear why heat multiplying plants may not be feasible with solar collectors. The total temperature difference must be at least the sum of the minimum permissible temperature differences (boiling point elevation) in each stage, and this sum might exceed the temperature difference possible with a solar energy collector.

Of most interest are the figures showing the distribution of energy, based on 100 units of solar energy entering the solar collector. It is seen that the net work is 4.5 units, i.e., the overall efficiency is 4.5 percent. This means that a PRO heat engine using $CCl_2F_2$ in this manner would require 4/**.045** = 90 km² area for a 1000 megawatt plant, as compared with about 1000 km² for a plant using water as the solvent.

As stated above, Freon **12** is an exceptional material. It may be one of the relatively few solvents with a latent heat of vaporization sufficiently low to overcome the basic deficiency when distiliation is used as the thermal umixing technique, namely that the latent heat of vaporization is usually too high in comparison to the free energy of separation.

## THERMAL UNMIXING BY SEPARATION INTO TWO LIQUID PHASES (FIGS. 12*a*, 12*b*, 13)

There may also be used a thermal unmixing system involving separation into two liquid phases, a technique which should be inherently more efficient than distillation since no latent heat of vaporization is involved.

A number of binary liquid systems are distinguished by the fact that their mutual solubility is a strong function of temperature. FIG. **12***a* shows such a system, methanol-hexane. Above a temperature of 42.6°C, known as the upper consolute temperature, the two species are miscible in all proportions. Below this temperature, say at 25°C, the liquids are only partially miscible, and two liquid phases exist in equilibrium, a 5 percent solution (by weight) of the methanol and a 95 percent solution of methanol. These two solutions are called conjugate solutions.



This behavior immediately suggests means of thermal unmixing in a PRO heat engine. Assume that a methanol-hexane solution is 22 percent methanol and at a temperature appreciably higher than 42.6°C. The solution is cooled down to 25°C at which temperature the 5 and 95 percent methanol solutions form. The amount of each solution will be determined by the lever arms y and X such that:

$$\frac{Y}{X} = \frac{\text{weight of 5\% solution}}{\text{weight of 95\% solution}} = \frac{100}{40}$$

These two solutions can be reheated above 42.6° to a temperature region where they are again naturally and completely miscible. At this temperature the total free energy of these two separated solutions is higher than that of the solution obtained by mixing them. (The free energy of a system decreases when it undergoes a natural process). Therefore the decrease in free energy upon mixing can be utilized to produce useful energy by means of the PRO section. After passage through the hydroturbine, the mixed solution is ready for cooling, and the cycle is completed.

The utility of this method is not limited to liquids having an upper consolute temperature. Binary systems such as triethylamine and $H_2O$ exhibit a lower consolute temperature, as can be seen in FIG. **12***b*. With such systems unmixing is accomplished by a temperature rise, followed by cooling of the separated liquids to bring them into the miscible range. Next, mixing occurs in the PRO section to produce useful energy; the mixed solution is heated to separate the concentrated and diluted solutions; and the cycle is completed.

The advantage over distillation of either type of such binary systems in the PRO heat engine is that the phase changes are liquid-liquid and not liquid-vapor. Thus the thermal energy input need not include values required by the latent heat of vaporization, but can approach much more closely to the free energy of separation in accordance with Guideline 2.

FIG. **13** (parts corresponding to those of FIG. **11** being correspondingly numbered) shows a PRO heat engine including a liquid separator, generally designated **156** which employs methanol-hexane as the liquid-system whose miscibility is a function of temperature. The methanol-hexane system was chosen to illustrate this method because its upper consolute temperature of 42.6°C is between the maximum temperature of 100°C, assumed to be the heat source temperature with a solar energy collector **142** and 25°C, assumed to be the lowest temperature to which the system can be conveniently cooled.

The following other conditions and assumptions were utilized:

1. The concentrated solution, diluted solution, and mixed solutions have the compositions of FIG. **12***a*. The weight ratio of concentrated to diluted solution is also as shown in FIG. **12***a*, i.e., 100/**40** = 2.5, and this

is equivalent, for reasons of density difference, to a volume ratio of 1/**0.36** = 2.8.

2. Raoult's Law applies to all solutions in the PRO section.

3. Efficiencies of the solar collector **142**, the pump **122**, and the hydroturbine **126** are as before 50, 95 and 95 percent respectively.

Based on these conditions and assumptions, the conditions of FIG. **13** were obtained. The mixed solution temperature, after passing through the hydroturbine **126**, is 95°C. It passes through a heat exchanger **152** where it is partially cooled to 48°C in the course of preheating the separated solutions. The mixed solution is then cooled to 25°C by heat rejection to the cooler (arrow **154**). This causes the concentrated and diluted solutions to form. These are divided in the separator **156**, after which they pass through the heat exchanger **152** where they are heated to 72°C. They then pass through the solar energy collector **142** where they are heated to 100°C. The concentrated solution is compressed by pump **122** to 360 atm, which warms it to 124°C after which it passes into the membrane unit **124** where it absorbs the diluted solution, so that the temperature is reduced to 117°C. In the membrane unit, the pressure drops to 358 atm on the mixed solution, after which it is depressurized through the hydroturbine **126**, producing work, and reducing the temperature of the solution to 95°C. This completes the cycle.

Also shown in FIG. **13** are the energy distributions. We consider the solar energy incident on the collector as 100 percent. Half of this is lost in the solar collector, and thus 50 percent goes forward to the heat engine. If this PRO engine were ideal, i.e. employed the Carnot cycle, then 10 percent of the solar energy incident on the solar collector would be available for work as shown in FIG. **8.** However the cycle is not a Carnot cycle since temperature absorption and rejection are not all accomplished at the maximum and minimum temperature. respectively. Therefore it is assumed that only 9 percent of the solar energy is available to supply the free energy of separation. This is a higher value than was obtained with distillation and follows from the fact that the separation into the two solutions is accomplished without an intermediate vapor phase, and thus no energy must be utilized to supply the latent heat of vaporization.

Of this 9 percent, 6 percent will be available for useful work and 3 percent will be unavailable due to mechanical and hydraulic losses in the circulating streams, and to the fact that some fraction of the available free energy is lost in providing an adequate driving force in the PRO section for permeant transfer.

## THERMAL UNMIXING BY USING SOLUTE WHOSE SOLUBILITY IS A FUNCTION OF TEMPERATURE (FIG. 14)

It is well known that solubility of salts and other solutes is a function of temperature. This behaviour in which can intermediate solid phase is produced, can be used as a basis for operation of the thermal unmixing section of a PRO heat engine. The technique can be used for solutes whose solubility either decreases or increases with temperature. This is illustrated in FIG. **14** for a solute whose solubility increases with temperature. The solubility characteristics of potassium nitrate are such that it could meet the solution and precipitation requirements shown.

As shown in FIG. **14** (parts corresponding to those of FIG. **13** being correspondingly numbered), the unpressurized mixed solution, containing 60 parts salt and 80 parts water, is 95°C after passing through the hydroturbine **126.** It passes through a heat exchanger **152** where it is cooled to 48°C in the course of preheating the filtrate (to be discussed) and the unpressurized concentrated solution. The mixed solution, which is now almost saturated with regard to the salts, is then cooled to 25°C by means of the cooler **160.** This causes salt to precipitate. The slurry of salt and solution is sent to a filter **162** which separates the slurry into a solid phase containing 30 parts of salt and a filtrate containing 30 parts of salt and 80 parts of water. The filtrate is warmed to 72°C by passing through the heat exchanger via line **164.** The hot filtrate is divided into two parts. One part, the diluted solution, containing 11 parts salt and 29 parts water, is sent via line **166** to the solar energy collector **142** for final heating. The remainder, 19 parts salt and 51 parts water, is sent via line **168** to the dissolver **170** where it dissolves the 30 parts of salt from the filter.

The solution emerging from the dissolver is the unpressurized concentrated solution, and it contains 49 parts salt and 51 parts water. It is assumed that in the dissolving process the solution temperature drops to 55°C. (The actual temperature attained will depend on the heat of solution, which can be positive or negative). The concentrated solution is passed via line **172** through the heat exchanger **152** where its temperature is raised to 72°C, the same as that of the diluted solution. These solutions then go through the same processes in the PRO section as described for partially miscible liquids, the cycle being completed with the unpressurized mixed solution emerging from the hydroturbine **126.**

This technique for thermal unmixing has the same basic limitation as distillation, namely the changing of one of the components into a phase other than a liquid phase. The energy required for this may be high compared to the free energy of separation. However it appears that by a judicious choice of solute this process can be made efficient.

Many other variations and applications of the invention will be apparent.

What is claimed is:

**1.** A method of generating power comprising: applying a hydraulic pressure to a first liquid of a first osmotic pressure and introducing same into a first pathway which is at least partially defined by one face of a semipermeable membrane; introducing a second liquid having a lower hydraulic pressure and a lower osmotic pressure into a second pathway which is at least partially defined by the opposite face of the membrane; maintaining the hydraulic pressure difference between liquids on the opposite faces of the membrane at a pressure difference which is less than the osmotic pressure difference between the liquids, at every point in the two pathways, thus effecting by Pressure-Retarded-Osmosis a passage of at least part of the second liquid through the semipermeable membrane, forming a pressurized mixed solution of greater volume than said first liquid introduced into said first pathway; and converting the potential energy stored in the pressurized mixed solution to useful energy.

**2.** The method according to claim **1**, wherein said energy is converted by passing the increased-volume pressurized first liquid through a turbine generator to generate electrical power.

3. The method according to claim 1, wherein said first liquid is passed through the first pathway in counter-flow with respect to the second liquid in the second pathway.

4. The method according to claim 1, wherein the first and second liquids are saline water solutions having different osmotic pressures.

5. The method according to claim 4, wherein the first liquid is sea water and the second liquid is water having a lower concentration of salt than sea water.

6. The method according to claim 4, wherein the first liquid is a naturally available salt-water body, and the second liquid is derived from a river or ocean feeding the salt-water body.

7. The method according to claim 4, wherein the first liquid is a body of water to which salt has been artifically added to produce the higher osmotic pressure solution.

8. The method according to claim 1, wherein the first liquid is fed from an evaporation pond, and wherein the evaporation pond is fed from the outlet of the first pathway after the energy in the first liquid has been converted to power.

9. The method according to claim 8, including the further step of recovering the water evaporated from the pond.

10. The method according to claim 8, including the further step of recovering the concentrated solution and/or solutes resulting from the evaporation of water from the pond.

11. Apparatus for generating power from heat comprising: a semipermeable membrane; means for applying a hydraulic pressure to a first liquid of a first osmotic pressure and introducing same into a first pathway which is at least partially defined by one face of the semipermeable membrane; means for introducing a second liquid having a lower hydraulic pressure and a lower osmotic pressure into a second pathway which is at least partially defined by the opposite face of the membrane; means for maintaining the hydraulic pressure difference between liquids on the opposite faces of the membrane at a pressure difference which is less than the osmotic pressure difference between the liquids, at every point in the two pathways, thus effecting by Pressure-Retarded-Osmosis a passage of at least part of the second liquid through the semipermeable membrane, forming a pressurized mixed solution of greater volume than said first liquid introduced into said first pathway; and means for converting the potential energy stored in the pressurized mixed solution to useful energy.

12. Apparatus according to claim 11, wherein said converting means comprises a turbine generator generating electrical power.

13. Apparatus according to claim 11, wherein said means for applying a hydraulic pressure comprises a liquid pump.

14. Apparatus according to claim 13, further including means for applying a portion of said energy to drive the pump.

15. Apparatus according to claim 11, further including an evaporation pond, means feeding the first liquid from the outlet end of the evaporation pond into the inlet end of the first pathway, and means feeding the first liquid from the outlet end of the first pathway after passing through said conversion means into the inlet end of the evaporation pond.

16. A multi-stage system for generating power, each stage comprising apparatus of claim 11, the hydraulic pressure of the first liquid in the first stage being highest and progressively decreasing through all the stages.

17. The method of claim 1 including the further steps of: recovering the first and second liquids by separating from said mixed solution a quantity of second liquid substantially equal to the quantity which passed through the membrane and mixed with the first liquid; restoring substantially the original temperatures to the recovered first and second liquids; reapplying the above mentioned hydraulic pressure difference between the recovered first and second liquids; and recycling the recovered first and second liquids through the first and second pathways respectively.

18. The method according to claim 17, wherein the recovering of said first and second liquids is effected by thermal separation.

19. The method according to claim 18, wherein said thermal separation is effected by distillation.

20. The method according to claim 18, wherein said thermal separation is effected by using as said first and second liquids solutions of two liquid species whose miscibility is a function of temperature.

21. The method according to claim 18, wherein the thermal separation is effected by using as said first liquid a solution of a solvent and a solute whose solubility is a function of temperature.

22. The method according to claim 18, wherein said thermal separation is effected by using solar energy as the energy source.

23. A heat engine including the apparatus according to claim 11, further including: separating means for recovering the first and second liquids by separating from said mixed solution a quantity of second liquid substantially equal to the quantity which passed through the membrane and mixed with the first liquid; temperature restoral means for restoring substantially the original temperatures to the recovered first and second liquids; means for reapplying the above mentioned hydraulic pressure difference between the recovered first and second liquids; and means for recycling the recovered first and second liquids through the first and second pathways respectively.

24. The heat engine according to claim 23, wherein said separating means and temperature restoral means are thermal means.

25. The heat engine according to claim 24, wherein said thermal separating means comprises a distillation device.

26. The heat engine according to claim 25, wherein both said first and second liquids include dichlorodifluoromethane as the solvent, and the solute is a low molecular weight compound such as ethanol.

27. The heat engine according to claim 24, wherein the two liquids are solutions of two liquid species whose miscibility is a function of temperature, and wherein said thermal separating means comprises means for changing the temperature of the mixed solution, such that the mixed solution will separate into the first and second liquids thus substantially recovering them.

28. The heat engine according to claim 27, wherein one of the said liquid species is methanol and the other is hexane.

29. The heat engine according to claim 27, wherein one of the said liquid species is triethylamine, and the other is water.

30. The heat engine according to claim 24, wherein the first liquid is a solution of a solvent and a solute whose solubility is a function of temperature, and wherein said thermal separating means including means utilizing said latter property to precipitate the solute.

31. The heat engine according to claim 30, wherein said first liquid is a solution of water and potassium nitrate, and said second liquid is water.

* * * * *

# UNITED STATES PATENT OFFICE
# CERTIFICATE OF CORRECTION

Patent No. 3,906,250 Dated September 16, 1975

Inventor(s) Sidney Loeb

It is certified that error appears in the above-identified patent and that said Letters Patent are hereby corrected as shown below:

In Column 12, line 37, "P>ΔP where ΔP" should be -- P<Δ¶ where Δ¶ --.

In Column 12, line 45, before "zero" "of" should be -- to --.

In Column 12, line 58, "ΔP" should be --Δ¶ --.

In Column 12, line 59, "ΔP" should be --Δ¶ --.

In Column 13, line 12, "cleaned" should be -- gleaned --.

In Column 16, "95" should be -- 65 -- in lines 6, 14 and 20.

In Fig. 12a, the right region marked "X" should be -- Y --.

In Fig. 13, "P = 0.642" should be -- P=0 --; and "¶mix= 358 atm." should be deleted.

**Signed and Sealed this**

*Twenty-fourth* Day of *January 1978*

[SEAL]

*Attest:*

RUTH C. MASON
*Attesting Officer*

LUTRELLE F. PARKER
*Acting Commissioner of Patents and Trademarks*

# UNITED STATES PATENT OFFICE
# CERTIFICATE OF CORRECTION

Patent No. 3,906,250 Dated September 16, 1975

Inventor(s) Sidney Loeb

It is certified that error appears in the above-identified patent and that said Letters Patent are hereby corrected as shown below:

In the "Foreign Application Priority Data", the first application listed therein should be identified as follows:

-- July 3, 1973 Israel ...... 42658 --.

In Column 7, line 51, "0.003 $/KW" should be -- 0.003 $/KWH --.

In Column 9, line 10, "daily cubic of" should be -- daily cubic meter of --.

In Column 11, line 35, "60-70 kilometers" should be -- 60-70 square kilometers --.

**Signed and Sealed this**

**Thirty-first Day of May 1977**

[SEAL]

*Attest:*

**RUTH C. MASON**
*Attesting Officer*

**C. MARSHALL DANN**
*Commissioner of Patents and Trademarks*