

Hot Ball Desalination

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ABSTRACT

A new process is described which is expected to achieve a significant reduction in the cost of producing potable water from seawater or brackish water. Basically the new process is a form of flash distillation employing solid balls as heat transfer agents. The process achieves significant improvement over previously disclosed distillation processes in that both thermal efficiency is improved and equipment costs are reduced. A solid ball-liquid heat exchanger is employed to heat the entering brine. The relatively cool solid particles are then reheated by transporting them against the pressure gradient through the flash stages. In the flash stages, the salt-free vapor is condensed directly on the surface of the solids, and collected and transported with the pressure gradient through any remaining flash stages. The heated solids are then further heated, e.g. by direct contact with gases from a fired heater. Thus, heat is transferred only twice, elaborate heat exchange equipment is eliminated, and maximum brine temperatures up to the critical are possible.

INTRODUCTION

The problem of producing potable water from seawater or brackish water is under active investigation today in many parts of the world. Methods which have received considerable attention include: reverse osmosis, extraction, electrodialysis, and many variations of distillation.

To date, distillation processes are the most common, but have been restricted to relatively low maximum operating temperatures owing to scale formation on conventional heat transfer surfaces. The result has been, according to Kogan (1968), that operating costs, (primarily the cost of energy required for the process), and equipment costs have been reasonably well optimized so that decreasing one increases the other, and vice versa, result-

ing in an overall increase in costs. Thus, a significant reduction in the cost of desalted water is dependent on a major change from current practice, such as operating at higher maximum temperatures, or using radically different heat transfer methods.

Attempts have been made to achieve these changes. Woodward (1966), as well as others, has suggested a vapor reheat process in which conventional heat exchange surfaces are replaced by external liquid-liquid heat exchangers and an elaborate stage internal construction to accomplish heat transfer between the flashing brine stream and a heating fresh water stream.

Woodward describes the advantages of vapor reheat distillation over conventional multi-stage flash distillation as follows: "Almost all heat-transfer surfaces or phase barriers are eliminated; scaling problems are reduced to a minimum; heat-transfer coefficients are high but heat must be transferred three times."

Briefly, this process involves flashing a hot brine stream in a series of stages and reheating the condensed vapor stream which flows through the stages countercurrent to the flashing brine. The hot fresh water then exchanges heat with a high-boiling immiscible liquid in an external liquid-liquid heat exchanger. Next, the immiscible heat exchange liquid transfers heat to the entering brine in a second liquid-liquid exchanger. Heat is supplied to the process by further heating the hot brine (usually in a conventional heat exchanger) prior to entering the first flash chamber.

One difficulty with this process is the very low relative velocities (and resultant low heat transfer coefficients) achieved in liquid-liquid exchangers. For even a moderate sized plant, these exchangers become quite large, and thus, quite expensive.

Kogan (1966, 1968) proposed an adaptation of Woodward's process in which the liquid-liquid exchangers were

replaced by solid-liquid exchangers. This had the distinct advantage of reducing the size and cost of the external heat exchangers. In addition, Kogan (1968) has claimed an improved stage internals design. However, his process is still dependent upon elaborate stage internals. Thus, although progress has been made, there remains significant room for improvement in producing potable water with distillation processes thus far revealed.

Briefly, the *Hot Ball Desalination* process employs one solid-liquid heat exchanger to heat the entering brine. The relatively cool solid particles are then reheated by transporting them in a countercurrent fashion, (against the pressure gradient), through the flash stages. In the flash stages, the salt-free vapor is condensed directly on the surface of the solid particles and then collected and transported with the pressure gradient through any remaining flash stages. The relatively hot solids are then further heated in order to make the process operable. (Some of these features are similar to those described by Heinz, (1969)).

This improves on previous distillation processes in the following ways:

1. Heat is transferred only twice. Flash vapor is condensed directly on a moving solids bed. Heat is then exchanged between the solids and the entering brine.
2. Vapors are condensed directly on the solids which eliminates much of the cost associated with stage internals which has been an appreciable part of the overall plant cost. All conventional heat-transfer tubing is eliminated.
3. Heat is added directly to the solids permitting increased maximum brine temperatures.
4. In the flash chambers, solids are transported against the pressure gradient, while the condensed vapor stream is transported with the pressure gradient. Thus, most interstage pumping is eliminated.

PROCESS DESCRIPTION

These and other features of the present process will become apparent by referring to Figure 1 and the accompanying detailed description of the present process. Figure 1 is a simplified flow diagram showing the essential features of the improved desalination process. In this diagram, brine flow is indicated by a solid line, desalted water flow by a dashed line, and solids flow by a double solid line.

Cold brine enters the process at 1 and then may or may not be combined with recirculated brine at junction 2. The cold brine stream is then compressed, 3, to a pressure that will insure that the brine will remain liquid upon reaching its maximum temperature in exchanger 4.

Cold brine enters exchanger 4 and after passing through a suitable flow distribution system, 5, is heated in countercurrent fashion by hot solid particles which are

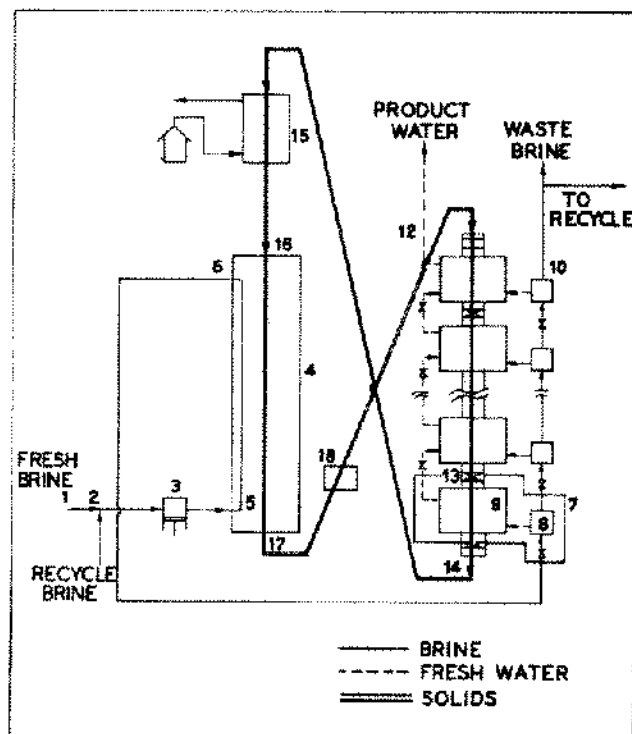


Figure 1. Simplified Process Flow Diagram, Hot Ball Desalination.

raining down through the brine under the force of gravity. Heated brine is then withdrawn from exchanger 4 at the top of the exchanger, point 6. Maximum brine temperatures may vary over a wide range, anywhere from 200°F to 700°F is anticipated.

Hot brine then enters the first of a series of flash stages, 7, in which the pressure of the brine stream is reduced by a suitable device in such a manner that the brine stream is partially vaporized. The decrease in pressure is accompanied by a decrease in temperature of the vapor-liquid mixture. The vapor, which is salt-free, is separated from the liquid brine by gravity, 8, and then enters a second part of the stage, 9, where it is contacted by a moving bed of relatively cool solid particles.

Brine from the first flash stage next enters a second flash stage which is maintained at a lower pressure than the first stage and the process described above is repeated. After passing through a series of such stages, the brine emerges from the last stage relatively cool at point 10. The brine is then disposed of as waste brine, or partially recirculated back to the entering brine stream at point 2.

The salt-free vapor in the first flash stage is contacted by a moving bed of relatively cool solid particles in a manner such that countercurrent heat transfer is approximated. As a result, the vapor is condensed on the surface of the particles. The condensed salt-free liquid is then separated from the solid particles by gravity, using some

device such as a screen. The condensed liquid then leaves the first flash stage, and after passing through a suitable pressure-reducing device, enters the second stage which, as previously indicated, is at a lower pressure and temperature than the first stage. Part of this liquid is thus also vaporized upon entering the second stage. In the second stage, the salt-free liquid is kept separate from the brine stream. The two vapor streams are combined and condensed by contact with relatively cool solid particles. This procedure is repeated in subsequent stages until a relatively cool product water stream emerges from the final stage at point 12. The number of such flash stages is variable and will depend on economic considerations regarding equipment costs and heat requirements.

The relatively cool solid particles which enter the first flash stage at point 13 are heated by condensing vapors to a temperature approaching that of the condensing vapor. The hot solid particles leave the first flash stage at point 14.

The hot solid particles then enter a heat exchanger, 15, where they, for example, are heated by contact with hot gases from a fired heater. The solids leave exchanger 15 at a temperature sufficiently high to insure that exchanger 4 will be operable.

The hot solids then enter exchange 4 at point 16. After passing through a suitable flow distribution device, the solids rain down through exchange 4 in direct countercurrent contact with the entering, relatively cool, brine. By this process, the solids are cooled, and the brine is heated. The overall brine and solids flow rates, and solids heat transfer properties are selected to maintain minimum temperature differentials at each end of exchanger and thus, minimize overall heat requirements. The cooled solids leave exchanger 4 at point 17.

After leaving exchanger 4, the solids may be washed in a suitable device at point 18. Washing may be required to eliminate contamination of the product water by brine carried with the solids into the flash chambers.

The relatively brine free and cool solids then enter the last (lowest temperature) of the series of flash chambers where they are contacted in countercurrent fashion by relatively hot salt-free vapors. As previously described, the solids are thus heated by the condensing vapors. The solids then flow (by gravity when possible) into the next higher temperature stage and the process is repeated. Finally, the solids enter the first (or highest temperature) stage at point 13, and the entire process is repeated.

EXAMPLE, 100,000 GALLON PER DAY PLANT

The process will be illustrated by considering a plant which produces 100,000 gallons per day of salt-free water.

In this example, 124,000 pounds per hour of fresh brine

enter the process at 86.4°F. The brine is compressed to an absolute pressure of 247.3 pounds per square inch and then enters the main heat exchanger at essentially 86.4°F and 247.3 psia. (The theoretical work of compression is 0.99×10^5 Btu/hr.)

The brine is then heated in the main heat exchanger by countercurrent contact with raining solids to a temperature of 400°F. The solids (for this example, taken to be 3/32 inch diameter iron spheres) enter the main exchange at 402°F and at a mass flow rate of 1.06×10^6 pounds per hour. The solids then leave the main exchanger at 88.4°F. The duty for the main exchanger is 39.6×10^6 Btu/hr. The temperature difference at both the hot and cold ends of the main exchanger is 2°F. However, owing to the effect of temperature on the heat capacities of water and iron, most of the heat will be transferred over larger temperature differences.

The hot brine next enters the first or highest temperature of the flash chambers. The pressure of the first flash chamber is set at 225.6 psia. Part of the brine stream vaporizes upon entering the first chamber until the equilibrium temperature of 392.0°F is achieved. The salt-free vapor (flow rate of 1,290 pounds per hour) is then separated from the brine and condensed at 392°F by contact with relatively cool iron spheres. The iron spheres enter the first flash chamber at 382.2°F and are heated to 390°F by the condensing vapors. The average temperature difference for heat transfer in the first chamber is thus about 6°F. The heat transferred in the first chamber is 10.7×10^5 Btu/hr.

Pressure in the second flash chamber is maintained at 201 psia. Brine leaving the first chamber (112,710 pounds per hour) is again partially vaporized until the equilibrium temperature of 384.2°F is reached. In this case, 1,230 pounds per hour of salt-free vapor are produced. In addition, condensed salt-free water from the first chamber is partially vaporized upon entering the second chamber. The vapor produced this way is 13 pounds per hour. Thus, a total of 1,243 pounds per hour of salt-free vapor are condensed in the second chamber. The condensing vapors heat the solids from 374.5°F to 382.2°F, and the total quantity of heat exchanged in the second chamber is 10.4×10^5 Btu/hr.

In like manner, the behavior of subsequent stages could be described. In this example, the plant has 52 flash chambers. The pressure of the 52nd chamber is 0.956 psia, and both salt-free water product (34,700 pounds per hour) and waste brine (89,300 pounds per hour) leave the 52nd chamber at 100.3°F. Iron spheres enter the last chamber at 88.4°F.

Because of the effect of temperature on both the heat of vaporization of water and the heat capacity of iron, (in this example), the vapor flow rates and amount of heat transferred in each stage is not a constant, but varies from

stage to stage. In this example, the quantity of heat transferred per stage reaches a minimum in stage 31 (about 5.9×10^5 Btu/hr.) and then increases again to about 9.3×10^5 Btu/hr. in stage 52. The total heat transferred in the series of flash chambers is 38.0×10^6 Btu/hr.

The solids leave the first flash chamber at 390°F and are then heated to 402°F by countercurrent exchange with hot gases from a fired heater. The amount of heat transferred to the solids at this point is 16.5×10^5 Btu/hr. which is also the minimum heat requirement for the process.

After reaching 402°F the solids enter the main heat exchanger and subsequently emerge at 88.4°F . The solids are then washed and fed to the 52nd flash chamber at 88.4°F .

The minimum heat required for the process is also equal to the difference between the heat removed from the solids in the main exchanger (39.6×10^6 Btu/hr.) and the heat added to the solids in the flash chambers (38.0×10^6 Btu/hr.). In addition, the heat required for the process may be obtained from an overall energy balance which takes into account the enthalpies of all entering and exit streams, and the external work done on the process. In this example, $\Delta H = 1.72 \times 10^6$ Btu/hr. and $W \approx 0.1 \times 10^6$ Btu/hr. which results in a minimum heat requirement of about 1.6×10^6 Btu/hr.

PROCESS MODEL

The results presented in the previous sections were calculated. Confirming experiments have yet to be made. At this point, it is worthwhile considering the basis for the calculations and some of the consequences.

The three basic elements of the process are the external liquid-solid exchanger, the series of flash chamber-condensers, and the solids heater where energy is introduced into the system. Of the three, it is believed that the last one is reasonably straight-forward and that a version of the "Pebble Heater" described in Perry (1963) can be employed. The other two elements are described below.

External liquid-solid exchanger

Heat exchangers based on the transfer of heat between a liquid and falling solid pellets, although still uncommon, have been proposed in at least two recent desalination processes. As mentioned earlier, Kogan (1966) has suggested their use in a modification of vapor reheat distillation. In addition, Fenske and Coworkers (1970) have recently disclosed that they are making use of "raining solids" exchangers in their current work on a high temperature extraction process for desalting seawater or brackish water. In addition to the elimination of conventional heat transfer surfaces, this type of exchanger has at least two other important advantages: high heat transfer coefficients and high throughput rates, both the result of high relative velocities between liquid and solid.

The estimation of heat transfer coefficients for liquid-solid exchangers is relatively straight-forward. Holman, J. P., et al. (1965), and Bird, R. B., et al. (1962) are among the sources of methods which can be used. In general, the coefficients for the system considered here are believed to be quite high. Values on the order of 1,000 Btu/hr.ft.²°F are expected, which, when combined with the high surface area per volume typical of liquid-solid exchangers, result in volumetric coefficients above 100,000 Btu/hr.ft.³°F. In the case described earlier, 3/32 inch diameter spheres were selected in order to maximize surface area and to reduce transient heating time.

In order to minimize process heat requirements, the temperature difference at each end of the main brine heater must be minimized. Two main factors must be considered in the design of such an exchanger:

1. The relative mass flow rates of brine and solids.
2. The material used in the solid balls.

A most important consideration is the selection of a suitable material for the solids. Under the proper conditions, this will determine the relative mass flow rates of brine and solids.

Figure 2 is an enthalpy diagram up to 500°F for a liquid-solid exchanger in which the liquid is water and the solid balls are iron. In this particular case, it is believed that a 2°F temperature difference may be fixed at both the hot and cold ends.

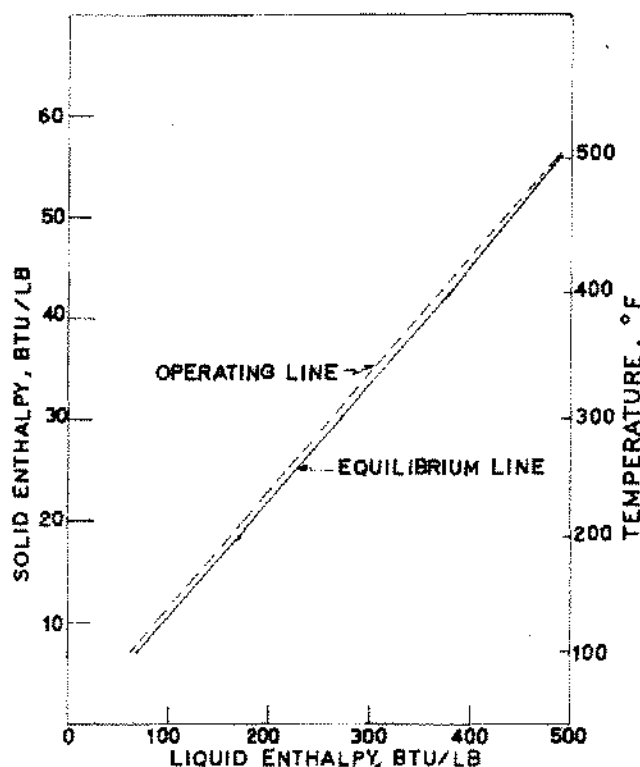


Figure 2. Enthalpy Diagram, Iron-Water Heat Exchanger.

In this figure, the lower line called the "equilibrium" line, is a plot of the enthalpy of the solid vs the enthalpy of the liquid at the same temperature. The "operating" line is then located as a straight line, above the equilibrium line, for the case of solids heating the fluid. The slope of the "operating" line determines the ratio of the total heat capacities of the liquid and solid over the selected temperature range. From this the relative mass flows may be determined. If the "equilibrium" line is straight or concave upward, the temperature differences at each end can be picked, and in theory, there will be no second law-violation in the exchanger. Iron and water show the desired type of behavior. Using the relative mass flows thus determined, about even volume flows result, which is desirable in order to reduce exchanger size. Iron is therefore a desirable material to use.

Flash chamber-condensers

The heart of the process is the series of stages in which the brine flashes and the salt-free vapors are condensed on the surface of the balls.

The method of operation (or model) picked for these chambers, and the one closest to conventional practice, is as described previously.

The heat balance for the series of flash chambers is done sequentially, one stage at a time, starting at the hot end.

These calculations were programmed to be done on a digital computer with saturated steam and water data incorporated as a sub-routine.

A minimum temperature approach of 2°F between condensing steam and the heating solid particles was assumed for these calculations. Given the facts that steam condensation coefficients are high, and the surface area available for condensation is large in each condenser, this temperature approach is probably conservative. In addition, the chamber size becomes quite reasonable.

MINIMUM ENERGY REQUIREMENTS

For any given maximum operating temperature, it is possible to investigate the relationship between number of stages and minimum heat requirements.

The minimum heat requirement, per pound of product, is determined by three factors:

1. The temperature difference allowed at the hot end of the liquid-solid exchanger.
2. The temperature drop taken in the first (hottest) of the flash chambers.
3. The maximum brine temperature.

Figure 3 shows the minimum process heat requirement, in Btu per pound of product water, plotted against number

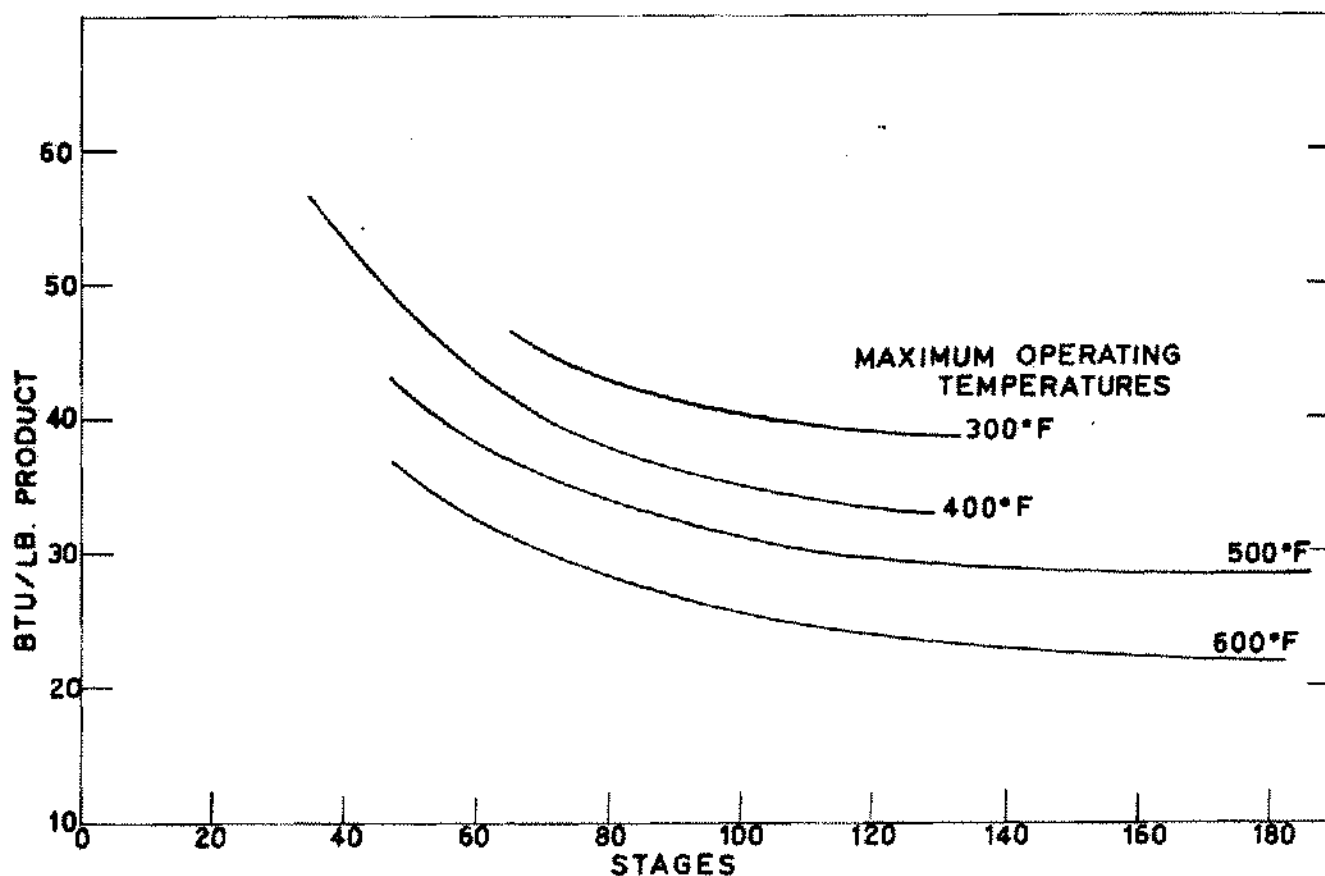


Figure 3. Minimum Process Heat Requirements.

of stages. Maximum brine temperature is shown as a parameter. As can be seen, increasing the maximum brine temperature significantly reduces the process heat requirements. In addition, increasing the number of stages has a pronounced effect on the heat requirement up to about 100 stages. After that point, the curves become rather flat.

The results shown in Figure 3 require further explanation. As mentioned earlier, in the process model considered herein, the amount of heat transferred in each stage is not equal. As shown in Figure 4, the heat transferred per stage actually passes through a minimum roughly halfway through the series of stages. In Figure 4, three curves are plotted for 40, 51, and 68 total stages, all for 400°F maximum brine temperature. Stage 1 is the highest temperature stage. In each case, the total amount of heat transferred by the series is about the same. However, the operation of each individual stage is clearly quite unique.

It is possible to model the process so that roughly an equal amount of heat is transferred in each stage of the series. This has been studied. However, the results are not included here.

CONCLUSIONS

A desalination process has been described which is expected to achieve significant improvements over present flash distillation and vapor reheat processes in the areas of both heat economy and equipment costs. The specific areas of improvement are:

1. Heat is transferred only twice, as compared to three times in previously described vapor reheat processes, which results in improved thermal efficiency.
2. Vapors are condensed directly on the solids. This eliminates much of the cost associated with stage internals which, in turn, has been an appreciable part of the total plant cost.
3. Heat required to operate the process is added directly to the solids which makes possible increased maximum operating temperatures.
4. In the flash chambers, solids are transported against the pressure gradient, while the condensed vapor stream is transported with the pressure gradient. Thus most inter-stage pumping is eliminated.

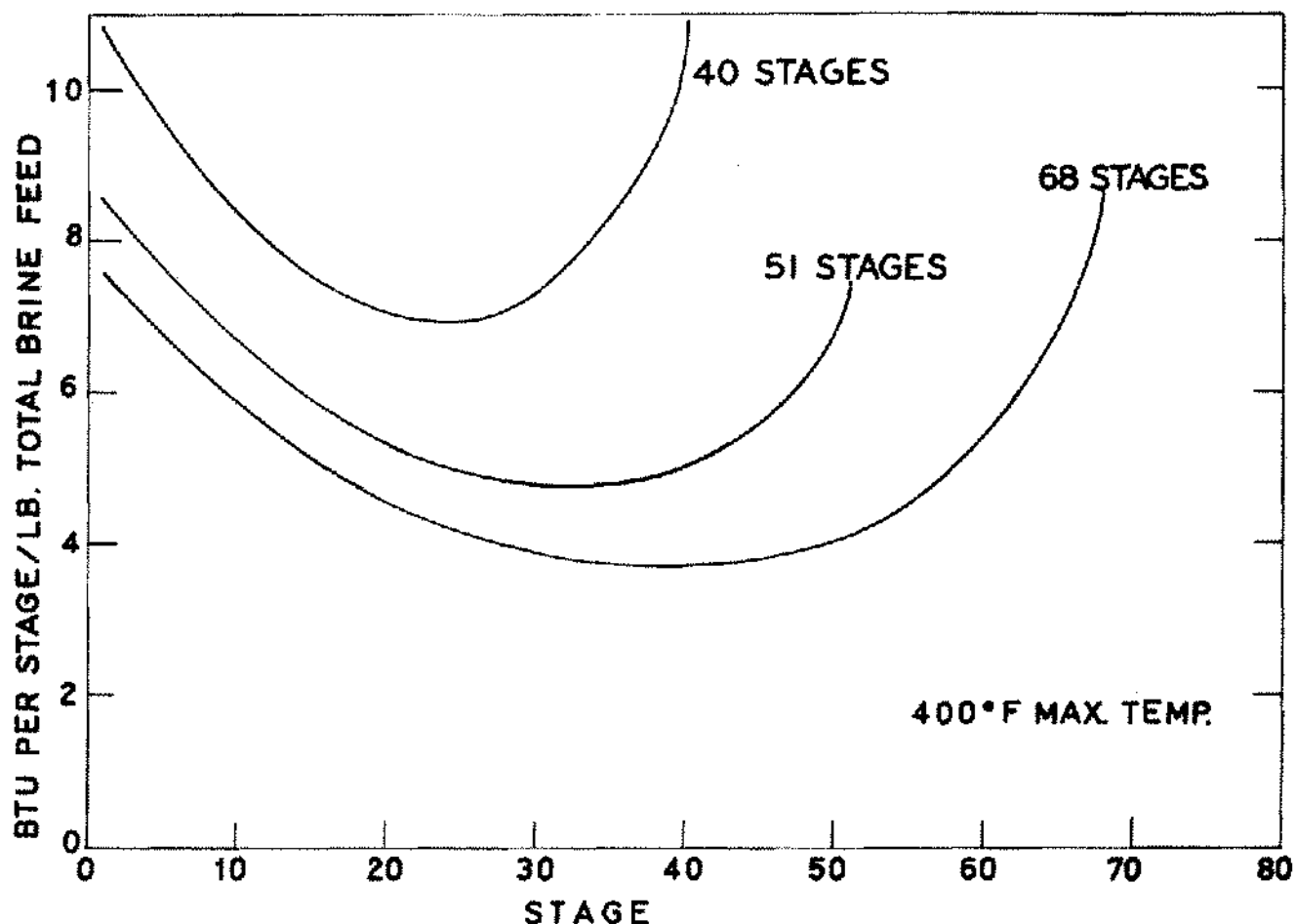


Figure 4. Heat Transferred per Stage.

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