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DOE/CS/30569-29
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OPEN CYCLE LITHIUM CHLORIDE COOLING SYSTEM

Final Report for the Period March 1, 1982—May 28, 1983

By
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May 1983

Work Performed Under Contract No. AC03-81CS30569

Colorado State Univeristy
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OPEN CYCLE LITHIUM CHLORIDE
COOLING SYSTEM

FINAL REPORT

for the Period March 1, 1982 to May 28, 1983

Prepared for the
Solar Heating and Cooling Research and Development Branch
Office of Conservation and Solar Applications
U.S. Department of Energy

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ABSTRACT

A lithium chloride open cycle absorption chiller has been designed, built and tested. Solution reconcentration takes place in a small counter-current packed column supplied with solar heated air. Removal of non-condensable gases that enter the chiller dissolved in the strong solution and the make-up refrigerant streams is accomplished by a liquid-jet ejector and a small vacuum pump. Cooling capacities approaching 1.4 tons and COP levels of 0.58 have been achieved at non-optimum operating conditions.

Test results from preliminary system operation suggest that mass transfer processes in both the packed column reconcentrator and the absorber are controlled by concentration gradients in the lithium chloride solution. Liquid phase controlled mass transfer dictates an operating strategy different from the previously assumed gas phase controlled process to obtain maximum rates of evaporation in the packed column. Determination of optimal operating conditions leading to decreased electrical power consumption and improved cooling capacity and coefficient of performance will require further analysis and testing.

BACKGROUND

Open cycle LiCl absorption air conditioning has recently been investigated as a means of solar cooling. A 10-ton system based on this concept has been designed and operated in the Soviet Union [1].

Conventional closed cycle aqueous lithium salt absorption chillers are usually configured as in Figure 1. Refrigerant (water) is vaporized in the evaporator and absorbed by the salt solution in the absorber, which is cooled to remove the latent heat and the heat of dilution. The cooled, diluted solution from the absorber is pumped through the heat exchanger to a higher pressure in the generator, where heat is added to boil off the refrigerant. Refrigerant vapor is liquefied on the cooled condenser coils and flows through an expansion valve back to the low pressure evaporator. Hot strong solution flows from the generator through the heat exchanger, back into the absorber.

In an open cycle chiller, the generator and condenser are replaced by a reconcentrator that evaporates refrigerant from the weak solution to the atmosphere as shown in Figure 2. Refrigerant must be made up from a water source. Provision must also be made to remove air that is dissolved in the make up refrigerant and the strong solution.

The reconcentrator used in the Soviet system, which has operated successfully for more than 5 years, is a gently sloping black roof over which the solution is trickled. The roof is open to the atmosphere and exposed to direct solar radiation. Roof evaporation works well in a hot, low humidity atmosphere, but not in humid climates. Dilution of the solution by rainfall and dust contamination are additional problems. To alleviate these problems, a glass cover over the solution regenerator

Figure 1. Closed Cycle Absorption Chiller Schematic

Figure 2. Schematic of the open cycle absorption air conditioning system.

has been proposed and separately tested [2]. This design is more expensive than the simple sloped roof, and both can conveniently use the collector area only for the specific purpose of solution concentration. To utilize these solar systems for space or domestic hot water heating requires either heat exchangers that can tolerate the corrosive salt solution or extensive modification of the collector itself to handle alternative heat transfer fluids during the heating season.

To make the open cycle chiller compatible with a conventional solar air heating system and to extend its application to more humid climates, the use of a packed column reconcentrator was investigated in a theoretical study [3]. In the column, solution flows by gravity over a suitable packing while solar heated air is forced counter-current to it in direct contact. Reconcentrated solution is collected below the packing and humidified air is exhausted to the atmosphere. In the absence of adequate mass transfer data from the literature, it was determined that packing depths of less than 30 cm and column diameters of less than one meter should be sufficient for evaporation of 15 kg of water per hour (approximately 3 tons of refrigeration). The study considered a wide range of air humidities and air temperatures of 70 to 90°C that could be obtained from a flat plate solar collector heating ambient air.

To substantiate the theoretical analysis and to verify the assumed heat and mass transfer coefficients, an experimental investigation of a packed column reconcentrator was undertaken [4]. A scaled down version of the 3 ton column was fabricated and instrumented to measure the coefficients required for the full size column design. Experimentally

measured heat transfer coefficients compared fairly well to those calculated from published correlations, but measured mass transfer coefficients varied significantly from the assumed values used in the theoretical study.

A second experimental study was undertaken to investigate the problem of removing non-condensable gases from the evaporator-absorber of the chiller [5]. Dissolved air in the incoming solution and refrigerant streams is released in the low pressure evaporator-absorber chamber. The presence of non-condensable gases degrades the performance of both the evaporator and the absorber. Maintenance of a steady vacuum in the chamber required vacuum pumping power of less than 15 W when the maximum theoretical non-condensable gas load was introduced. Because water vapor contamination of the vacuum pump oil was observed, modifications that would probably increase the pumping power were recommended for the gas removal system.

The present study was undertaken to prove and demonstrate the concept of open-cycle absorption cooling in a complete system in which chilled water is produced, and the solution is continuously regenerated in a packed column by water vaporization into solar heated air.

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OBJECTIVES

The objectives of this project were to design, build and test an open cycle LiCl absorption chiller with packed bed regeneration by solar heated air. The specific objectives of machine operation were to demonstrate the open cycle absorption cooling concept by:

1. achieving cooling,
2. further investigating the operational performance of the packed column reconcentrator,
3. testing the non-condensable gas removal system,
4. investigating absorber performance, and
5. testing system cooling performance.

This report includes the results of the above investigations and recommends some system design improvements for the next generation prototype.

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EXPERIMENTAL APPARATUS AND OPERATION

Introduction

As indicated in the above background material, the major thrust of the present research has been the design, assembly, and testing of a complete nominal 3-ton open cycle cooling system. An overall schematic diagram of this cooling system, which combines the packed bed regenerator and chiller subsystems is shown in Figure 3.

Component Description

Packed Column

Based on the previous packed column studies, the column was designed to reconcentrate dilute LiCl solution by vaporization of water at a rate of 15 kg per hour. The column is 81 cm in diameter, packed with 2.5 cm raschig rings to a depth of 23 cm as shown in Figure 4. The solution distributor is a dish with 104 gravity-fed holes, each irrigating an equal area of the packing as illustrated in Figure 5. Air enters the base of the column, passes upward through the packing, counter-current to the dilute lithium chloride, through the risers in the liquid distributor and exits through the mesh-type mist eliminator, to the exhaust duct. The packing support is constructed of epoxy-coated steel bar and mesh, while the rest of the column is fabricated of glass-reinforced polyester. The base of the column holds a variable quantity of strong solution as a means of storing cooling capacity. Solution flows from the base of the column for chiller operation. A portion of the solution is also recirculated from the base of the column back to the distributor to ensure adequate wetting of the packing.

Figure 3. Cooling System Schematic

NOTE: ALL DIMENSIONS IN cm

Figure 4. Packed Column

Figure 5. Liquid Distributor

Evaporator-Absorber

The evaporator-absorber chamber of the chiller is fabricated from a modified ARKLA WF-36 lithium bromide chiller. The chamber contains concentric 1.91 cm (3/4 inch) copper helical tube banks externally wetted by overflow trays as shown in Figure 6. Evaporator tube area is 1.34 m² and absorber tube area is 1.91 m². Unevaporated refrigerant (water) is collected and recirculated to the refrigerant distributor tray, and make-up water replaces evaporated refrigerant to hold the level constant in the solution reservoir. Solution level is controlled in the base of the absorber, from which it is pumped to the regenerating column and through the non-condensables purging system. For off-season bench testing, the evaporator load is supplied from an electric water heater, and the absorber cooling water is supplied from mains.

During preliminary testing of the evaporator-absorber, solution contamination of the refrigerant loop was observed. The refrigerant loop was reconfigured for the remainder of the testing. In the reconfigured mode, refrigerant is not recirculated and any unevaporated refrigerant spills into the base of the absorber. This modification increases the load on the packed column for a given amount of evaporation, but does not significantly affect the operation of the evaporator or the absorber.

Non-Condensable Gas Removal System

Non-condensable gases are removed from the evaporator-absorber by a liquid-jet ejector at a point about three-fourths the way up the absorber tube bank. The ejector is operated by LiCl solution that entrains the gases and pumps them up to an intermediate pressure separator as shown

Figure 6. Evaporator-Absorber

in Figure 3. The increased pressure of the air-water vapor mixture from the absorber and its direct contact with the LiCl solution decrease the concentration of water vapor in the non-condensable gases in the separator. The non-condensable gases can then be removed from the separator by a vacuum pump without water vapor contamination of the vacuum pump oil. Solution flows from the separator to the solution pump inlet as shown in Figure 3.

Instrumentation

The chiller is instrumented as shown in Figure 3. Reference temperatures are measured by copper-constantan thermocouples. Temperature differences are measured by multi-junction thermocouples (thermopiles). Pressures in the evaporator-absorber and in the separator are measured by 0-50 mm Hg absolute pressure dial gages. Liquid flows are read from rotameters calibrated for water and corrected for solution specific gravity where appropriate. Hot air flow rate to the column is calculated from an electronic manometer reading across a pitot rake. Air removed from the separator is metered by trapping in a water-filled, inverted graduated cylinder. Humidities of air entering and leaving the column are calculated from wet and dry bulb thermometer readings with 0.1°C resolution. Thermocouples, thermopiles and electronic manometer outputs are recorded by an automatic data logger. All other measurements are recorded manually. Solution samples are taken intermittently for later measurement of specific gravity. Concentration is calculated from the measured specific gravity.

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CALCULATION PROCEDURES

Packed Column

The fundamental processes of heat and mass transfer that take place in the packed column have been assumed to be gas phase controlled. Assumptions involved in the definition of heat and mass transfer coefficients are:

1. heat and mass transfer coefficients can be considered to be constant at some average value over the height of the column,
2. the specific heat of air can be considered to be constant throughout the column,
3. Dalton's Law can be used for the partial pressure of water vapor in air, and
4. the liquid temperature is constant across the column, with the interface temperature equal to the bulk temperature.

Heat and mass transfer coefficients can then be related to column parameters as follows:

$$h_{Ga} = \frac{G'(C_p)_{av}}{Z} \ln \frac{T_{G1} - T_L}{T_{G2} - T_L}$$

$$k_{Ga} = \frac{G'}{Z} \ln \frac{H_{int} - H_{G1}}{H_{int} - H_{G2}}$$

$$\text{where } H_{int} = \frac{P_L M_w}{(P - P_L) M_a}$$

The units of volumetric heat and mass transfer coefficients are kW/m³°C and kg water/s m³ (kg water/kg air) respectively. The variables are defined in the nomenclature section. Liquid vapor pressures are calculated from

equilibrium properties of LiCl solutions based on measured temperature and concentration. Air humidities are calculated from measured wet bulb and dry bulb temperature readings. The overall data reduction procedure for the column is flowcharted in Figure 7.

Evaporator-Absorber

The heat and mass transfer processes that take place in the absorber are somewhat more complicated than those that occur in the packed column, because of absorber cooling. Solution temperature and concentration profiles are assumed to be linear across the absorber. The absorber mass transfer coefficient is defined as

$$k_{MASS} = \frac{MABS}{PDEL \times AREA}$$

MABS is the mass of vapor absorbed and is calculated by summing the masses of refrigerant vaporized in the evaporator, refrigerant flashed on evaporator entry and solution flashed on absorber entry. The absorption driving force (PDEL) for calculating mass transfer is the difference between the pressure of water vapor in the evaporator-absorber and the vapor pressure of the solution at the assumed average bulk temperature and concentration. The pressure of water vapor in the evaporator-absorber is assumed to be equal to the measured evaporator bulk pressure. The overall data reduction procedure is flowcharted in Figure 8. The data analyzed are for steady-state conditions for the absorber. The criteria for steady-state are changes in solution inlet temperature and evaporator temperature difference of less than 0.2°C over 3 2-minute scans and flow rate changes of less than 2 percent over the same period.

Figure 7. Packed Column Data Reduction Procedure

Figure 8. Evaporator-Absorber Data Reduction Procedure

RESULTS AND DISCUSSION

Packed Column

To ensure adequate wetting of the packing, solution flow rates through the column were kept near 0.25 kg/s. This flow rate should provide 90% wetting with a gas flow of 0.6 kg/s approaching only 55% of the flooding velocity [6]. The flooding velocity is the gas velocity that causes dynamic liquid hold-up in the packing to such an extent that air flow passages are blocked by liquid.

The maximum evaporation rate obtained for the column was without the chiller in operation which allowed high solution temperatures to be reached. Evaporation of 11.6 kg/hr was obtained with air inlet temperature of 68°C and air inlet humidity of 0.0016 kg/kg. Solution inlet temperature and concentration were 49°C and 44 wt % with measured heat and mass transfer coefficients of 8.19 kW/m³°C and 2.37 kg/s m³ kg H₂O/kg dry air, respectively. The coefficients for this particular operating condition are consistent with those obtained at other operating points, so the high evaporation rate can be attributed to the high air and solution temperatures which provide high mass transfer driving forces.

The column was typically operated at air and solution flow rates of 0.6 and 0.25 kg/s, respectively, with air inlet temperatures and humidities ranging from 40 to 68°C and 0.0016 to 0.0054 kg/kg. At ambient air temperatures of 27°C these absolute humidities would represent relative humidities of 6% to 20%. Solution temperatures varied from 32 to 49°C and inlet solution concentration varied from 38 to 46 wt %.

The heat and mass transfer coefficients in the column at various air flow rates and at two different solution concentrations are presented

in Figures 9 and 10. In general, both coefficients increase with increasing air flow rate and decrease with increasing solution concentration.

The gas phase heat transfer coefficient from the literature [7] has been correlated with other variables in the following equation:

$$h_G a_h = 5.34 G^{0.7} L'^{0.07} e^{0.0041(t_f + 17.8)}$$

where t_f is the gas film temperature in °C and L' is the superficial liquid mass velocity. Air flow rate is clearly the dominant variable. The moderate scatter in the data is probably due partially to experimental errors and to variation in L' and t_f with which the coefficient has not been correlated here. Dependence of the heat transfer coefficient on solution concentration is probably due to the increasing viscosity of the solution with concentration. A viscosity increase may result in a decrease in the effective area for heat and mass transfer by increasing liquid film thicknesses and solution hold-up in the packing. Increased film thicknesses and the corresponding increase in liquid thermal resistance would bring the liquid-phase resistance to heat transfer closer to that of the gas phase. The assumption that the heat transfer process in the column is gas-phase controlled may have to be modified at high solution concentrations.

Figure 9. Column Heat Transfer Coefficient

Figure 10. Column Mass Transfer Coefficient

The heat and mass transfer analogy would show that the mass transfer coefficient, like the heat transfer coefficient, should depend on air flow rate. An increase in solution concentration decreases the diffusivity of water molecules in the solution. Lower mass transfer coefficients at high concentration can be explained by the existence of a concentration gradient in the liquid film. Water vapor evaporates from the liquid-gas interface in the column, increasing the solution concentration at that surface. Water must diffuse through the laminar liquid film from less concentrated solution to reach the interface. The lower diffusivity and greater film thickness associated with a greater solution bulk concentration tend to reduce the rate of water diffusion to the interface. The concentration gradient in the film becomes steeper at high solute concentrations in the bulk solution. The interface humidity is therefore not actually in equilibrium with the bulk solution concentration, but rather with a somewhat higher concentration at the interface.

In the previous experimental packed column study, for a mean concentration of 27.4 wt% and an air flow rate of 0.45 kg/s, the mass transfer coefficient was measured as $4.21 \text{ kg/s m}^3 \text{ kg H}_2\text{O/kg dry air}$ [4]. This is significantly higher than the mass transfer coefficients measured at higher concentrations in the present study. As for heat transfer at high concentrations, the assumption that mass transfer is completely gas-phase controlled does not seem to be appropriate. Figure 10 indicates that the high concentration operation of the column is in the regime where mass transfer is shifting to a completely liquid phase controlled process. The decreasing slope of the high concentration line indicates that mass transfer is becoming independent of air flow rate. This observation suggests that a different strategy for column operation should be used. Higher temperature operation of the column would provide increased driving force

for mass transfer along with increased solution diffusivity and decreased solution viscosity. Although the measured gas-phase heat transfer coefficient would decrease for this situation, the overall rate of evaporation may increase. The depth of packing should probably be increased to compensate for the decreased mass transfer coefficient compared to previous studies.

Experimentally measured pressure drops for air flow through the column are presented in Table 1. Also listed are the measured values of the pressure drop coefficient α from this study as compared to a value given by Leva [8] for 2.5 cm Rashig rings. The expression

$$\Delta P = \frac{\alpha G'^2 Z}{\rho}$$

is suggested for dry packing pressure drop calculations. The measured and published coefficients are quite consistent. Pressure drop measurements were not made for wetted packing, but it was noted that the air inlet velocity to the column did not change significantly for the same blower with wet versus dry packing. The column pressure drop is quite high and is of the same order as the pressure drop through the solar collector array.

The high packing pressure drop and hence the high fan power requirement could be substantially reduced by changing the type of packing. Stacked Rashig rings, for example, can have pressure drop coefficients of one-fourth that of the dumped rings used in the present column [8]. Since the mass transfer process in the packed column does not seem to be entirely gas-phase controlled, increased column pressure drop should not be a necessary condition for increased mass transfer. Other types of packing are available that should exhibit lower pressure drop than Rashig rings without decreased mass transfer.

For liquid-phase controlled mass transfer in the packed column, the increased driving force associated with higher temperature column operation may be more important for total evaporation rate than high air flow rates. Higher temperature air would require lower flow rate through the collector and column with reduced pressure drop.

Table 1. Packed Column Pressure Drop

G' kg/m ² s	V m ³ /s	T_{G1} °C	P kPa _a	ΔP_{meas} Pa	α_{meas} m ⁻¹	α_{LEVA} m ⁻¹
1.16	0.697	66.8	84.0	196.	548	534
0.775	0.466	66.3	84.0	71.2	447	534

Evaporator-Absorber

The absorber was supplied with strong solution from the solar heated regenerator over a relatively wide range of temperatures, flow rates, and concentrations. Chilled water and cooling water flow rates and temperatures varied, so scatter is observed in the evaporator-absorber performance data. Evaporator-absorber operational results for a range of inlet conditions are summarized in Table 2. The chiller was operated at two solution flow rates of approximately 0.07 and 0.14 kg/s. Temperatures of chilled water leaving the chiller ranged from 13°C to 21°C. Solution inlet temperature and concentration varied from 23 to 41°C and from 40 to 46 wt%, respectively.

The chilled water temperatures obtained from the preliminary testing of the machine are somewhat higher than normally expected from a lithium salt absorption chiller. Continued testing of the open cycle chiller should focus on operating at chilled water temperatures below 10°C as a more useful operating condition for space cooling.

Table 2. Evaporator-Absorber Test Results

COOLING CAPACITY	STRONG SOLUTION MASS FLOW	STRONG SOLUTION INLET TEMPERATURE	STRONG SOLUTION CONCENTRATION	BULK PRESSURE	CHILLED WATER MINIMUM TEMPERATURE	MASS TRANSFER COEFFICIENT	WEAK SOLUTION SUB-COOLING
kW	kg/s	°C	wt %	mmHg	°C	kg/s m ² mmHg	°C
3.97	.0804	36.27	44.46	5.90	15.80	.00106	6.26
4.33	.0811	35.80	43.86	6.25	16.50	.00116	5.62
4.89	.0813	38.00	41.71	8.10	12.97	.00093	6.25
5.46	.0775	31.60	40.07	8.65	17.27	.00061	6.89
6.34	.0739	35.60	45.29	8.50	20.37	.00054	8.15
4.71	.1424	36.30	41.44	6.80	13.03	.00136	5.72
4.86	.1429	36.60	41.65	6.90	13.60	.00140	5.57
5.35	.1440	40.77	46.25	7.40	16.80	.00147	4.65
5.99	.1488	33.30	41.23	7.50	16.90	.00122	5.73
7.43	.1480	39.70	44.36	8.15	19.43	.00122	7.21

More testing of the evaporator-absorber is needed to more fully understand and develop quantitative descriptions of the processes that occur. This testing should be done with tightly controlled chilled water temperature so that the effects of strong solution temperature and concentration on cooling capacity can be detected. When the controlling resistance for mass transfer can be more clearly identified, steps can be made to reduce it in future absorber designs.

The resistance of a laminar solution film in closed cycle LiBr absorption chillers is reduced by "heat transfer additives" that are put into the solution to produce two-phase flow and break up the laminar layer. The open cycle system may not be compatible with these additives since they are usually somewhat volatile and would be lost during solution reconcentration. Non-volatile additives may therefore need to be identified and used, possibly in conjunction with other methods of breaking up the laminar layer or increasing the exposed absorbent surface area. The transfer resistance of a film of non-condensables that might be surrounding the absorber seems to be less significant than that of the liquid film.

A measure of absorber performance that is commonly used in the absorption cooling industry is the amount of sub-cooling of the weak solution as it leaves the absorber. Sub-cooling is the difference between the actual temperature of the weak solution and the temperature at which it would be if its vapor pressure were equal to the absorber bulk pressure.

For comparison, the Carrier model SAM10A nominal 3-ton chiller has solution sub-cooling of less than 3°C for solution flow rate of approximately 0.7 kg/s [9]. The open cycle chiller operates with solution sub-cooling of about 6°C and as high as 8.9°C for flow rates of 0.14 kg/s . Assuming similar solution heat capacities, the excess solution cooling in the LiCl system is 6 times that of the Carrier system even when the Carrier machine is producing almost 2 times as much cooling for the load.

A portion of the sensible heating of the weak solution will be done in the heat exchanger, but more heating will be required in the regenerator. Solution sub-cooling and solution flow rate should both be minimized to reduce the sensible heating requirements of the regenerator as well as the required capacity of the cooling tower.

Non-Condensable Gas Removal System

The performance of the non-condensables removal system has been better than expected. Removal of air from the evaporator-absorber and maintenance of a good vacuum were found to require a lower solution flow rate in the ejector than had been planned. The reduced fluid flow requirement can be attributed to an actual pumping load of 0.2 cc/s of non-condensables compared to an assumed theoretical maximum load of 0.7 cc/s (at Standard Temperature and Pressure). The separator chamber was operated at 27.5 mm Hg . The effect on non-condensables removal of varying this pressure was not investigated. The chamber could conceivably be controlled at a higher pressure to further reduce the required vacuum pump power while still providing adequate gas removal. During normal operation of the system over more than a month, no water vapor contamination of the vacuum oil was detected.

Chiller System

Unlike the closed cycle absorption chiller, the open cycle chiller is not constrained to operate at equal capacities for cooling and regeneration over the short term. Different capacities can be achieved by the components because of the strong solution storage and because the source of refrigerant for evaporation is independent of refrigerant driven off in the packed column. Operation at the same capacities at any point in time is purely coincidental. The effects of independent component operation are net increases or decreases of strong solution temperatures and concentrations. For practical system operation, the concentration swings must be controlled between certain predetermined limits, and the column or chiller operation must consequently cease until the concentration has been raised or lowered to an acceptable level.

A possible operating cycle for the chiller is superimposed on the equilibrium diagram for LiCl in Figure 11. The cycle shows weak solution from the absorber heated from 28.5 to 60°C as it flows through the heat exchanger to the packed bed. The solution temperature then rises from 60 to 65°C as it is reconcentrated from 43 to 44 wt %. The strong solution is cooled to 35°C as it passes back through the heat exchanger. As the solution is diluted by the absorption of water vapor, it is further cooled to 28.5°C. The equilibrium refrigerant temperature at absorber inlet conditions is 6°C. The departure of the absorber outlet solution from equilibrium with 6°C refrigerant is 4.5°C, which is the sub-cooling of the solution.

Because of the decoupled components of the system, it is extremely difficult to measure a system coefficient of performance (COP) for short

Figure 11. Equilibrium Diagram for LiCl with Cooling Cycle Superimposed

term operation. By matching the capacities of the chiller and column operational data obtained at different times, performance factors for the system are estimated for the conditions in Table 3. Performance is compared for a case of coupled chiller and column operation with high sensible heating load and a case of decoupled packed column operation with very low sensible heating load. For both situations, cooling capacity is calculated from column evaporation capacity. Higher air temperature and flow along with lower inlet humidity also contribute to increased evaporation capacity and increased performance factors for the decoupled case. The measures of solar cooling performance are defined as:

thermal coefficient of performance

$$\text{COP} = \frac{\text{cooling delivered}}{\text{heat input to regenerator}}$$

cooling system performance factor

$$\text{CSPF} = \frac{\text{cooling delivered}}{\text{heat transferred to air in collector}}$$

solar performance factor

$$\text{SPF} = \frac{\text{cooling delivered}}{\text{solar incident on collector surface}}$$

electrical coefficient of performance

$$\text{ECOP} = \frac{\text{cooling delivered}}{\text{electrical power to pumps and fans}}$$

For the worst case operating conditions of high solution and air flow rates, the electrical power consumption for the system was measured to total 2.23 kW as itemized in Table 2. Because of the bench test mode of the current testing, the total power consumption does not include the cooling tower fan and pump or the chilled water circulation pump. The

TABLE 3. CHILLER SYSTEM PERFORMANCE

	Coupled	Decoupled
<u>Packed Column</u>		
Air flow rate (kg/s)	0.534	0.574
Air inlet temperature (°C)	52.7	68.1
Air inlet humidity (kg/kg)	0.0037	0.0016
Recirculated solution flow rate (kg/s)	0.107	0.256
Recirculated solution temperature (°C)	40.3	48.1
Weak solution flow rate (kg/s)	0.141	---
Weak solution temperature (°C)	31.5	---
Solution concentration (°C)	43.3	44
Heat transfer coefficient ($\text{kw/m}^3\text{°C}$)	7.45	8.19
Mass transfer coefficient (kg/s m^3)	2.50	2.37
Heat transferred to column (kW)	5.86	8.97
Mass of refrigerant evaporated (kg/s)	0.00137	0.00322
<u>Evaporator-Absorber</u>		
Cooling capacity (kW)	3.42	8.03
Electrical power (kW)	2.23	2.23
<u>Collector</u>		
Heat transferred to air (kW)	23.16	34.24
Energy incident on collector (gross area)(kW)	55.90	64.01
<u>Performance Factors</u>		
COP	0.58	0.90
CSPF	0.15	0.23
SPF	0.06	0.13
ECOP	1.53	3.60

maximum power usage in the system is by the air blowers and the solution pumps. As discussed in the previous section, absorber performance enhancement would decrease the required solution flow and the associated pumping power. The lowered sensible heating load on the packed column would decrease the air flow requirements, which when combined with an improved packing choice for pressure drop, would substantially lower the regenerator fan power requirement. With optimization of fan and pump selection along with absorber performance equivalent to open cycle systems, the parasitic power consumption could probably be halved. Increased cooling capacity associated with absorber improvements would also contribute to an improved ECOP.

Table 4. Electrical Power Consumption

Solution pumps	690 W
Column recirculation pump	100 W
Hot air blowers	1430 W
Vacuum pump	12 W
TOTAL	<u>2232 W</u>

Conclusions

1. An open cycle absorption cooling system has been successfully operated with solar heated air as the primary energy source.
2. Cooling capacities approaching 1.4 tons have been obtained while producing chilled water at 13°C.
3. Chiller COP levels have reached 0.58, under non-optimum conditions.
4. Removal of non-condensables by a combination of ejector purging and low-power vacuum pumping has been successfully obtained.
5. Regeneration of lithium chloride absorbent in a small packed column has been accomplished by use of solar heated air at temperatures as low as 37°C, and not exceeding 69°C.
6. Regeneration capacity can be increased and high electrical power consumption should be significantly reduced by a new packed column operating strategy.
7. Electrical power requirements can be decreased and system capacity and COP can be increased by absorber performance enhancement.

RECOMMENDATIONS FOR FUTURE WORK

In the near future, slight modifications should be made to the chiller system to allow continued controlled testing in an effort to better understand its operation. Recommendations of parameters such as solution temperatures, concentrations and flow rates for best system operation can then be made. The non-condensables removal system should be slightly modified to reduce the fluid recirculation and the associated pumping power. A different packed column operational strategy should be tested because of the apparent liquid phase controlled mass transfer. Higher packed column temperature operation should increase evaporation as well as decrease fan power.

Work for the next generation prototype chiller should include a thorough investigation of heat transfer additives and other means of absorber performance enhancement. A packed column redesign should be undertaken to determine the optimal combination of packing geometry and air flow rate in terms of pressure drop, evaporation rate and collector efficiency. A control scheme for the chiller should be designed to provide automatic control of the evaporator-absorber, regenerator, non-condensables removal system, and collector.

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NOMENCLATURE

Packed Column

a_h	= heat transfer area per unit of gross volume (m^2/m^3)
a_m	= mass transfer area per unit of gross volume (m^2/m^3)
C_{pav}	= average heat capacity of gas (KJ/Kg °C)
G'	= mass velocity Kg dry air/s m^2 of open cross section of empty tower
H_{G1}, H_{G2}	= humidity of air at inlet and outlet conditions for packed tower (Kg H_2O /Kg dry air)
H_{int}	= interface humidity (Kg H_2O /Kg dry air)
h_G	= heat transfer coefficient (KW/ m^2 °C)
$h_a a_H$	= volumetric heat transfer coefficient (kW/ m^3 °C)
k_G	= mass transfer coefficient for gas phase (Kg H_2O /s m^2 (Kg H_2O /dry air))
$K_a a_m$	= volumetric mass transfer coefficient (kg H_2O /s m^3 (Kg H_2O /Kg dry air))
M_a	= molecular weight of air (g/g mole)
M_w	= molecular weight of water (g/g mole)
P_L	= vapor pressure of liquid (mmHg)
P	= total pressure (mmHg)
$T_{G,n}$	= bulk temperature of gas phase (°C)
T_{G1}, T_{G2}	= temperature of air in, temperature of air out (°C)
T_L	= average liquid temperature (°C)
V	= volumetric air flow rate (m^3 /sec)
Z	= packed tower height (m)
ρ	= density of air (Kg/ m^3)

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NOMENCLATURE

Evaporator-Absorber

VLOAD	chilled water flowmeter reading (gpm)
VCW	cooling water flowmeter reading (gpm)
VSOLN	strong solution flowmeter reading (% f.s.)
MREF	refrigerant mass flow (kg/s)
PBULK	evaporator-absorber bulk pressure (mm Hg)
T2	chilled water return temperature to load (°C)
T6	cooling water temperature (°C)
T1	strong solution inlet temperature to absorber (°C)
TP27	thermopile reading across chilled water loop (mV)
TP26	thermopile reading across cooling water loop (mV)
TREF	make-up refrigerant temperature (°C)
X1	strong solution concentration (wt %)
TIME	time of reading
DATE	date of run
TSUMP	solution sump temperature (°C)
QEVAP	cooling capacity (kW)
QABS	absorber cooling requirement (kW)
MEVAP	mass of refrigerant evaporated (kg/s)
MEFLASH	mass of refrigerant flashed (kg/s)
MAFLASH	mass of water flashed from solution (kg/s)
MABS	mass of vapor absorbed (kg/sec)
X2	concentration of strong solution after flash (wt %)
TF	temperature of strong solution after flash (°C)
P2	vapor press of strong solution after flash (mm Hg)
MSOLN	mass flow rate of strong solution (kg/s)
X0	solution concentration out of absorber (wt %)
T0	solution temperature out of absorber (°C)
P0	solution vapor pressure out of absorber (mm Hg)
PDEL	average pressure difference across absorber (mm Hg)
TSUB	solution sub-cooling out of absorber (°C)
KMASS	mass transfer coefficient (kg/s m ² mm Hg)
XBAR	average solution concentration (wt %)
TBAR	average solution temperature (°C)
VISC	average solution viscosity (cp)

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Printed in USA.

SOLAR REGENERATION OF LITHIUM CHLORIDE SOLUTIONS IN OPEN CYCLE ABSORPTION COOLING

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ABSTRACT

Reconcentration of a lithium chloride solution in an open cycle absorption chiller can be accomplished by passing solar heated air through a packed column to which the dilute solution is supplied. Following a theoretical study of heat transfer and water vapor transfer rates in the column, experimental measurement of those rates was made. Heat transfer and mass transfer coefficients are correlated with rates of air and liquid flow, and with temperatures of air and liquid supply.

Performance data are presented and commercial design and operating requirements are suggested.

INTRODUCTION

The concept of open cycle LiCl absorption air conditioning has been examined as a possible solution to the need for a cooling system which can be combined with and operated by a solar air heating system. A 10-ton cooling system based on this principle has been operated in an apartment building near Ashkabad in the USSR [1]. In contrast with conventional absorption refrigeration systems in which both the absorbent and refrigerant circulate in a closed cycle, only the absorbent, a lithium chloride solution, is recirculated in this open cycle. Water, the refrigerant, is continually evaporated to the atmosphere, and must therefore be continuously added to the cycle. The condenser and generator normally used in a conventional closed cycle system are not necessary, so there is a substantial reduction in equipment requirements.

The reconcentration of the lithium chloride solution in the USSR air conditioner is accomplished by trickling it across a gently sloping black roof (Figure 1) which is directly exposed to solar radiation. The absorbed solar energy and a hot, dry atmosphere permit adequate reconcentration of the solution for return to the absorber. In the absorber, water vapor from the evaporator is again absorbed, and the cycle continues. The refrigeration effect is obtained by low temperature (less than 5°C) evaporation of the added water.

Figure 1. A Simplified Schematic of An Open Cycle Absorption Chiller (Drawing from "The Analysis and Simulations of An Open Cycle Absorption Refrigeration System" by R.K. Collier)[2]

Figure 2. Schematic of the open cycle absorption air conditioning system.

In order to broaden the application of the open cycle system to less arid locations, the possibility of reconcentrating the absorbent solution in a packed bed (stripping column) by means of solar heated air was investigated in a theoretical study by Leboeuf and L&F [7]. Figure 2 is a schematic diagram of the USSR system as modified for use with solar heated air. The roof used for reconcentrating the LiCl solution in the USSR installation is replaced by a packed column through which the LiCl solution trickles countercurrent to a rising stream of solar heated air.

To substantiate the theoretical analysis and to verify the mass transfer and heat transfer coefficients assumed in that analysis, an experimental investigation of evaporation rates from LiCl solutions in a packed column was undertaken. The objective was to obtain data which would permit reliable sizing of a packed column for use in a complete solar cooling system of the open cycle LiCl type.

Theory

Considering liquid film resistances to heat and mass transfer in an air-water system to be negligible, the following equations relating height of a packed column to film coefficients in the gas phase are obtained:

$$h_{G,a_h} = \frac{G'(C_p)_{av}}{Z} \ln \frac{T_{G1} - T_L}{T_{G2} - T_L} \quad (1)$$

$$k_{G,a_m} = \frac{G'}{Z} \ln \frac{H_{int} - H_{G1}}{H_{int} - H_{G2}} \quad (2)$$

(Symbols are defined in Figure 3 and the nomenclature list).

The basic differential equations are:

For mass transfer

$$G'dy = -k_{G,a_m}(y_n - y_{int})dZ \quad (3)$$

For heat transfer

$$G'C_p dT_G = -h_{G,a_h}(T_{G,n} - T_{int})dZ \quad (4)$$

The following assumptions are involved in the integration of the basic differential equations:

1. Heat and mass transfer coefficients are constant over the height Z and are independent of temperature and humidity changes. Coefficients calculated from experimental data are average values.
2. The specific heat of air is considered constant at an average value.
3. In expressing partial pressure and vapor pressure in terms of humidity, Dalton's law is employed.
4. The liquid temperature is considered constant, and the interface value is equal to the main body temperature.

Interface humidity is calculated from the equation

$$H_{int} = \frac{P_L M_A}{(P - P_L) M_B} \quad (5)$$

where P_L is the vapor pressure of lithium chloride solution at the concentration and temperature under consideration, and P is the total pressure. M_A and M_B are molecular weight of water and air.

Figure 3. Heat and Mass Transfer in Packed Column

EXPERIMENTAL APPARATUS

Figure 4 shows the experimental system for measurement of heat and mass transfer rates. The 0.5 m by 1.3 m packed column was shop-fabricated from copper sheets (Figure 5). It is provided with piping connections for supplying liquid to the top of the column and withdrawing liquid from the bottom section. Air is introduced at the bottom through a 30 cm square duct. A perforated base plate supports the packing of 1-inch (25mm) ceramic Raschig rings. The solution is distributed uniformly over the packing by means of a perforated copper plate located immediately below the liquid entrance in the upper section of the column (Figure 6). The packed bed height can be varied up to a maximum of 75 cm.

Figure 4. Experimental Apparatus for Measurement of Rates of Heat Transfer and Water Vapor Transfer in a Stripping Column in which a Lithium Chloride Solution is Being Contacted with Heated Air.

Figure 5. Schematic Diagram of Experimental Packed Column

Figure 6. Liquid Distributor Plate

Air was supplied to the packed column at variable rates, temperatures, and humidities from a system consisting of a centrifugal blower, a propane gas air heater, a humidifier, and flow meter. Humidity could be controlled by the combination of moisture from propane combustion and by direct water spray into the air supply duct. The rate of air supply to the column was measured in the supply duct, and air temperatures and humidity were measured at the column entrance and exit. Humidity measurements were made by means of a recording dew point hygrometer.

The desiccant solution flow rate was measured at the column inlet, and temperatures of solutions entering and leaving the column were recorded. The concentration of LiCl in solution was determined by accurate specific gravity measurement.

EXPERIMENTAL PROCEDURE

Experimental runs were performed with air flow rates of 8.5 to 14.2 m³/min (300 to 500 cfm). The liquid flow rate was 5.67 to 11.15 liters/min (1.5 to 2.95 gallons/min) with air inlet temperatures of 60°C to 105°C (140°F to 220°F).

After starting the blower, the duct gate opening was adjusted to give the desired flow rate, as indicated by orifice meter reading. The propane gas heater was then started, and the gas pressure to the burner was adjusted so that the inlet air temperature to the packed column was at the desired level.

A previously prepared lithium chloride solution of the selected concentration was then added to the packed column where it collected in the lower (sump) section. Circulation was established by turning on the pump and adjusting the control valve.

Hygrometer chart readings were observed as an air sampling valve was alternately switched between inlet and outlet sampling points. Liquid and air temperatures and air humidity at inlet and outlet of the packed column were continuously monitored by strip chart recorders.

No heat was added to or removed from the liquid, other than that transferred from the hot air. As operation continued, the temperature of the circulating solution increased until a nearly constant value was reached. Steady state operation, usually reached in about one hour, was indicated by no further changes in temperature and humidity. A liquid sample was then collected, and liquid and air temperatures and air humidities were recorded. At a later time, the specific gravity of the solution was measured by weighing an exactly known volume of solution in a picnometer.

RESULTS

Complete data and results of the investigation are shown in Tables 1 and 2. Coefficients of heat transfer and water vapor transfer measured at several air flow rates over a range of air temperatures and humidities are tabulated. Also listed are the computed values of the heat transfer coefficient based on the McAdams [4] empirical relationship:

$$h_{cah} = 1.78(G')^{0.7}(L')^{0.07}e^{(0.0035)t_f} \quad (6)$$

used by LeBoeuf and Lof [3] in the previous theoretical study of this process. In this equation, $t_f = t_L + (t_{G1} - t_L)/2$. Although this equation is based on experimental measurements with pure water in a column packed with 1-inch rings, the degree of agreement with the results of the present work with lithium chloride solutions is within expected limits.

The mass transfer coefficients shown in Table 1 are seen to vary over a wide range, reasons for which are suggested below.

Table 2 contains data showing the specific influence of air flow rate on the measured heat transfer coefficient. In Figure 7, the same information is correlated graphically.

DISCUSSION

The experimental apparatus is designed for the evaporation of about 11 lbs (5 kg) of water per hour (equivalent to one ton refrigeration capacity), but it was operated at evaporation rates up to about 18 lbs (8.16 kg) per hour. Results in Table 1 show that this evaporation rate is feasible at indicated

Table 1. Data and Results of Experiments - Bed Height - 0.75 ft (229 mm)

Air Flow Rate (cfm)	Liquid Flow Rate (gpm)	Air Inlet Temp. °F	Air Outlet Temp. °F	Air Inlet Humidity lb H ₂ O/lb air	Air Outlet Humidity lb H ₂ O/lb air	Liquid Inlet Temp. °F	Liquid Outlet Temp. °F	Mean Liquid Concentration Wt% LiCl	Interface Humidity lb H ₂ O/lb air	Evaporation Rate lbs/hr	Heat Transfer Coefficient (h _{ca}) Btu/hr ft ² °F	Theoretical Heat Transfer Coefficient Btu/hr ft ² °F	Mass Transfer Coefficient lb/hr ft ³ (k _{ga})	Ratio Heat Transfer Mass Transfer
426	1.5	228	126	.0160	.0249	101	102	21.4	.0294	16.6	461	537	1242	.37
416	1.5	228	126	.0175	.0270	100	100.5	22.4	.0318	17.7	450	537	1243	.36
342	1.5	185	116	.0144	.0288	93	94	12	.0358	18.5	317	422	932	.34
342	1.5	182	124	.0167	.0288	95	95	8.4	.0395	18	247	421	692	.35
342	1.5	186	112	.0120	.0237	96	97	17.8	.0349	17.5	396	425	656	.60
342	1.5	178	104	.0116	.0203	97	98	27.4	.0260	13	568	420	844	.67
342	1.5	180	120	.0167	.0271	94	94	9.6	.0375	15.5	269	421	631	.43

Table 2. Effect of Airflow Rate on Heat Transfer Coefficient - Bed height - 0.75 ft (229mm)

Air Flow Rate cfm	Liquid Flow Rate gpm	Air Inlet Temp °F	Air Outlet Temp °F	Mean Liquid Concent. Wt% LiCl	Liquid Inlet Temp °F	Liquid Outlet Temp °F	Heat Transfer Coeff. Btu/hr ft ² °F	Theoretical Heat Transfer Coefficient Btu/hr ft ² °F
290	2.2	185	108	34.1	102	105.5	363	392
290	2.2	181	105	32	98	100	322	386
302	2.2	210	130	35	119	120	184	443
302	2.2	210	124	34.9	111	111	447	436
315	2.95	205	114	27	107	107	549	445
315	2.95	202	113	27	106	105	581	431
325	2.2	216	112	25.9	102	102	577	447

Conversion factors for Table 1 and Table 2:

100 cfm = 2.83 m³/min
 1 gpm = 3.785 L/min
 °F = 1.8 (°C) + 32
 1 lb H₂O/lb air = 1 kg H₂O/kg air
 1 lb/hr = 0.4535 kg/hr
 1 Btu/hr ft²°F = 17.29 W/(m²·K)
 1 lb/hr ft³ = 16.02 kg/(hr·m³)

temperatures and flow rates in a packed tower containing approximately 2 cubic feet (0.057 m³) of one-inch (25 mm) Raschig ring packing.

Table 1 shows that the experimental measurements of heat transfer coefficients are in the same range as those predicted for the air/water system by Eq. (6). Differences between h_{ca} predicted by Eq. (6) and determined experimentally can be attributed partly to experimental error and partly to the difficulty in operating at fully adiabatic conditions. In spite of thermal insulation, heat loss was found to be significant. Figure 8 shows that after steady-state operation of the dry column had been achieved (after 24 minutes in the test depicted), heat loss from the column resulted in cooling of the air about 10°C. The outdoor location of the installation and varying atmospheric conditions made heat loss estimates uncertain.

Another cause of discrepancy between measured and theoretical heat transfer coefficients was variation in inlet air temperature. Fluctuation in fuel supply to the air heater resulting from on-off control operation caused as much as ±5°F (±2.8°C) variation in air inlet temperature during a run. It is estimated that the values of the heat transfer coefficients are subject to errors of about 10 percent.

If heat loss from a packed column in a practical cooling system is minimized by use of adequate insulation (R-20 is suggested), satisfactory operation can

Figure 7. Effect of Airflow Rate on Heat Transfer Coefficient

be achieved with air supply at temperatures as low as 140°F (60°C). In areas of high atmospheric humidity, operation at higher air temperatures, up to 180° (82°C) for example, permits use of a shorter column. The experimental results show that conventional flat plate solar air collectors and pebble-bed heat storage units are suited to the conditions required for effective operation of the lithium chloride concentration system.

Figure 8. Time-Temperature Comparison in Dry Packed Column

Mass transfer coefficients and heat transfer coefficients are related to each other by the equation [5],

$$h_G/k_G = c_p \left(\frac{Sc}{Pr} \right)^{2/3} \quad (7)$$

and for the special case of air and water [5],

$$\frac{h_G}{k_G} = 0.24 \text{ Btu}/(\text{lb}\cdot\text{F}) = 1005 \text{ J}/(\text{kg}\cdot\text{K})$$

Experimental ratios of heat transfer and mass transfer coefficients usually show considerable variation, even in systems of pure liquids in packed columns. Variations are due to several factors:

- The ratio of $\frac{h_{Ga}}{k_{Ga}}$ is affected by the ratio $\frac{a_h}{a_m}$ which is not necessarily equal to unity in a packed column. Mass transfer and heat transfer areas may be unequal because of liquid channeling, wall effects, and incomplete wetting of the packing.
- Mass transfer coefficients can be substantially affected by end effects in a column. Vapor transfer can take place at the bottom and top of the column where air and liquid are in contact outside the packed zone. These end effects vary from column to column and cannot be generalized.
- If dry areas are present in the packing due

to poor liquid distribution or low liquid flow rate, these areas are available for heat transfer but not for mass transfer.

d) An error of 0.001 in measuring the humidity driving force for mass transfer can affect the value of the mass transfer coefficient by as much as 30% [5]. Hence slight error in liquid temperature or humidity measurement or in concentration of solution can cause considerable error in the computed mass transfer coefficient.

e) The apparent psychrometric ratio, $\frac{h_{Ga}}{k_{Ga}}$ for a packed column is neither constant nor equal to the humid heat [5]. Rather it varies from 0.23 to 0.58 Btu/(lb·F), (963 to 2429 J/(kg·K)) in air-water systems, depending on flow rates and packed heights. This ratio will also vary in a packed column in which lithium chloride solutions are being concentrated.

The increase in heat transfer coefficient with increase in air rate shown in Figure 7 indicates the desirability of employing the highest air mass flow rates consistent with pressure drop and flooding velocity constraints. Pressure loss measurements in the operating experimental column showed 0.10 to 0.25 inch water gage (0.025 to 0.0625 kPa) pressure drop at air mass flow rates of 290 to 385 cfm (8.2 to 10.9 m³/min). These losses would normally be consistent with requirements and capabilities in solar air heating systems.

As anticipated, it was also found that variation in lithium chloride concentration caused virtually no change in the experimental heat transfer coefficient.

The heat transfer coefficients shown in Table 2 and plotted in Figure 7 differed as much as 25 percent from values computed by Eq. (6), the average difference being about 15 percent. Differences in the coefficients reported in Table k, at lower liquid flow rates and a wider air temperature range, averaged about 30 percent for reasons previously outlined.

The height of a packed column should be designed with an adequate margin for uncertainty in the mass transfer coefficient. Fortunately, the coefficient is sufficiently high and the column size (for a 3-ton cooling capacity, for example) is sufficiently small, that cost is not significantly increased by oversizing.

It is concluded that theoretical prediction of heat transfer rates in the concentration of lithium chloride solutions can be made by use of McAdams' equation for heat transfer coefficients based on air and water in packed columns. The ratio of heat transfer to mass transfer coefficients shows considerable variation, so it is recommended that column design for lithium chloride concentration be based on heat transfer coefficients.

SUMMARY

- o The influences of air flow rate, humidity, and temperature on the coefficients of heat transfer and mass transfer in a packed column in which water is vaporized from a LiCl solution by heated air have been experimentally determined.
- o Theoretical predictions of transfer coefficients and packed column sizes in an open cycle LiCl solar cooling system have been experimentally verified.
- o The feasibility of operating an open cycle cooling system in which LiCl solutions are concentrated in a packed column supplied with solar heated air has been established.

NOMENCLATURE

a	= area of surface per unit of gross volume (sq ft/cu ft)(m ² /m ³)
a_h	= heat transfer area, ft ² /cu ft (m ² /m ³)
a_m	= mass transfer area, ft ² /cu ft (m ² /m ³)
$C_{p,av}$	= heat capacity of gas (Btu/lb°F)(J/(kg·K))
D_p	= diffusivity (ft ² /hr)(m ² /hr)
G'	= mass velocity lb dry air/hr sq ft of open cross section of empty column (kg/(hr·m ²))
H_{G1}, H_{G2}	= humidity of air at inlet and outlet conditions for packed column (lb H ₂ O/lb dry air)(kg H ₂ O/kg dry air)
H_{int}	= interface humidity (lb H ₂ O/lb air)(kg H ₂ O/kg air)
h_G	= heat transfer coefficient (Btu/hr sq ft°F)(W/(m ² ·K))
k_G	= mass transfer coefficient for gas phase (lb/(hr)(sq ft)(kg/(hr·m ²)))
k	= thermal conductivity (Btu/hr ft°F)(W/(m·K))
L'	= mass velocity, lb liquid/hr sq ft of open cross section of empty column (kg/(hr·m ²))
M_A, M_B	= molecular weights (lb/lb mole)(kg/kg mole)
P_L	= vapor pressure of liquid, (mmHg)(kPa)
P	= total pressure (mmHg)(kPa)
r	= distance from center line of duct of unit radius (for circular ducts), ft (m)
$T_{G,n}$	= bulk temperature of gas phase (°F)(K)
$T_{L,n}$	= bulk temperature of liquid phase (°F)(K)
T_{int}	= interface temperature (°F)(K)
T_{G1}, T_{G2}	= temperature of air in, temperature of air out (°F)(K)
t_L	= average liquid temperature (°F)(K)
x	= fractional linear distance from one end point of the interval $0 \leq x \leq 1$ (for open channels)
Z	= packed column height (ft)(m)
V_1	= velocity (ft/hr)(m/h)
Sc	= Schmidt number ($\mu/\rho D$)
Pr	= Prandtl number ($C_p \mu/k$)
μ	= viscosity (lbm/ft sec)(kg/(m·s))
ρ	= density (lb/ft ³)(kg/m ³)

Subscripts

1	= bottom of column
2	= top of column
av	= arithmetic average
h	= heat transfer as in $h_G a_h$
m	= mass transfer as in $k_G a_m$

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ACKNOWLEDGEMENT

The authors acknowledge with appreciation The Office of Solar Energy Applications, U.S. Department of Energy, for the support of the work on which this paper is based.

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APPENDIX

Evaporator-Absorber Data and Results

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VL740 GEVAP XBAR	VC4 SA43 THAR	VSOLN MEVAP VISC	MREF MEFLASH	PHULK MAFLASH	T2 MABS	T6 X2	T1 TF	TP27 P2	TP26 MSOLN	TREF X0	X1 T0	TIME P0	DATE PDEL	TSUMP TSUR	KYASS
5.75	3.31	1.10	.002983	6.10	16.70	4.83	28.67	673.70	1586.00	6.43	42.67	1307.30	8303.01	23.87	
NO SOLUTION FLASHING															
4.79	5.30	.00192	.000001	0.00000	.00193	42.67	28.67	5.52	.0743	41.59	22.85	4.16	1.250	6.19	.00030
42.13	25.75	10.14													
5.75	3.53	1.09	.003250	6.07	15.48	4.78	28.68	568.30	1461.00	6.28	42.52	1316.15	8303.01	23.88	
NO SOLUTION FLASHING															
4.76	5.24	.00191	.000001	0.00000	.00192	42.52	28.68	5.59	.0738	41.44	22.61	4.14	1.209	6.18	.00083
41.95	25.64	10.02													
5.71	3.51	1.13	.002830	5.83	16.05	5.43	35.53	579.30	1457.00	7.73	44.52	1144.30	8303.02	26.10	
4.89	5.33	.00154	.000002	.00026	.00192	44.56	32.04	5.73	.0807	43.62	24.89	4.08	.873	5.84	.00115
44.14	24.46	11.41													
5.72	3.50	1.14	.002830	5.90	15.80	5.17	35.27	560.70	1489.00	7.50	44.46	1157.00	8303.02	25.90	
3.97	5.43	.00154	.000002	.00029	.00190	44.62	32.13	5.90	.0804	43.59	24.62	4.03	.936	6.25	.00106
44.10	23.41	11.39													
5.92	3.40	1.17	.002830	6.00	15.80	4.97	37.23	520.30	1552.00	7.47	44.40	1212.00	8303.02	26.40	
4.45	5.35	.00175	.000002	.00034	.00214	44.59	32.43	6.00	.0798	43.42	25.34	4.25	.868	5.62	.00129
44.01	23.38	11.14													
5.42	3.40	1.20	.002830	6.10	15.97	4.97	36.83	505.30	1559.00	7.50	44.21	1224.30	8303.02	26.63	
4.32	5.37	.00173	.000002	.00032	.00207	44.39	32.45	6.10	.0814	43.28	25.55	4.36	.869	5.52	.00123
43.83	23.00	10.92													
5.52	3.33	1.19	.002830	6.25	16.50	4.80	35.80	507.80	1503.00	7.55	43.66	1253.15	8303.02	26.48	
4.33	5.52	.00174	.000002	.00035	.00200	44.00	32.37	6.25	.0811	42.93	25.42	4.44	.905	5.62	.00115
43.45	23.40	10.58													
6.31	5.57	1.20	.002830	6.01	16.74	4.53	35.53	516.30	1003.00	7.30	43.50	1310.45	8303.02	25.22	
4.46	7.08	.00177	.000002	.00031	.00200	43.67	31.29	6.01	.0815	42.57	24.22	4.23	.891	5.73	.00123
43.12	27.75	10.56													
6.80	3.01	1.12	.003130	7.20	16.82	5.73	37.27	627.70	1901.00	7.57	45.25	1508.15	8303.07	27.40	
5.27	5.78	.00211	.000001	.00006	.00214	45.28	36.50	7.20	.0762	44.03	28.25	4.88	1.162	5.59	.00098
44.55	32.38	10.75													
6.41	3.95	1.11	.003130	7.25	16.15	5.63	37.22	601.70	1898.00	7.60	45.07	1517.30	8303.07	29.17	
5.06	6.85	.00203	.000001	.00005	.00210	45.11	36.38	7.25	.0759	43.89	27.93	4.83	1.211	5.35	.00091
44.50	32.15	10.65													
6.40	3.02	1.10	.002980	7.15	16.05	5.40	33.14	579.00	1765.00	7.08	44.10	1559.09	8303.07	28.75	
NO SOLUTION FLASHING															
4.87	5.30	.00195	.000000	0.00000	.00196	44.10	33.18	6.50	.0750	42.98	27.66	5.08	1.358	5.72	.00075
43.54	30.42	10.22													
5.76	2.80	1.21	.003800	8.30	16.13	5.00	37.37	673.70	2329.00	7.50	43.32	1214.30	8303.09	29.70	
5.55	7.50	.00227	.000001	.00007	.00234	43.35	36.45	8.30	.0823	42.16	28.30	5.60	1.350	6.64	.00091
42.76	32.37	9.01													
5.77	2.80	1.21	.003800	8.30	15.70	5.43	37.55	662.30	2307.00	7.48	43.22	1221.30	8303.09	29.53	
5.55	7.53	.00223	.000001	.00010	.00232	43.27	35.34	8.30	.0823	42.08	28.10	5.56	1.368	6.74	.00089
42.58	32.22	8.93													

Evaporator-Absorber Data and Results

5.75	2.40	1.21	.003410	8.30	15.23	5.88	37.53	635.50	2272.00	7.40	43.10	1229.30	8303.09	29.35	.00088
42.58	32.19	.00214	-.00001	.00011	.00224	43.16	36.19	8.30	.0623	42.01	28.18	5.62	1.339	6.55	
5.75	4.74	1.20	.003600	7.75	14.18	5.53	36.74	634.80	1383.00	7.23	42.71	1245.30	8303.09	27.74	.00095
42.21	30.52	.00214	-.00000	.00018	.00232	42.80	34.55	7.75	.0815	41.62	28.48	5.21	1.271	6.61	
5.75	4.75	1.18	.003600	7.53	13.58	5.32	36.16	624.80	1305.00	6.98	42.45	1300.30	8303.09	26.82	.00095
41.94	29.52	.00210	-.00000	.00019	.00230	42.55	33.74	7.53	.0801	41.36	25.50	4.99	1.271	6.80	
5.77	4.83	1.14	.003600	7.48	13.13	5.14	36.23	603.00	1297.00	6.83	42.19	1315.30	8303.09	26.53	.00094
41.73	28.24	.00204	-.00000	.00023	.00227	42.31	33.33	7.48	.0800	41.14	25.14	4.95	1.263	6.73	
5.79	5.75	1.18	.003410	7.30	12.80	5.07	36.17	607.80	1080.00	6.73	41.89	1331.30	8303.09	25.79	.00103
41.44	28.51	.00205	-.00000	.00028	.00235	42.04	32.58	7.30	.0799	40.84	24.64	4.90	1.199	6.52	
5.79	2.70	1.20	.003600	7.80	15.87	6.27	34.37	676.00	2222.00	7.63	43.95	1026.05	8303.10	29.10	
NO SOLUTION FLASHING															
43.35	31.10	.00228	.00000	0.00000	.00228	43.95	34.37	7.05	.0818	42.76	27.82	5.21	1.669	6.78	.00072
5.78	2.93	1.19	.004770	7.55	13.65	5.95	36.60	603.30	2040.00	7.35	43.37	1053.05	8303.10	29.43	.00083
42.37	30.43	.00204	-.00000	.00013	.00218	43.44	34.91	7.55	.0810	42.30	25.95	4.80	1.375	7.53	
5.79	2.78	1.15	.003830	7.70	13.43	5.97	37.57	576.70	2144.00	7.53	43.04	1108.05	8303.10	29.17	.00099
42.59	31.13	.00195	-.00000	.00021	.00217	43.15	34.88	7.70	.0809	42.02	27.49	5.38	1.159	5.99	
5.78	3.10	1.14	.007070	8.00	13.10	5.77	38.27	563.30	2105.00	6.93	42.68	1124.05	8303.10	29.03	.00057
42.25	29.95	.00191	-.00001	.00025	.00215	42.81	35.11	8.00	.0801	41.69	24.78	4.66	1.670	8.95	
5.74	2.79	1.20	.003930	8.15	12.98	5.58	39.08	554.40	2310.00	7.26	41.90	1159.05	8303.10	28.42	.00089
41.51	30.53	.00189	-.00001	.00034	.00223	42.08	34.52	8.15	.0810	40.95	26.73	5.54	1.304	5.41	
5.74	2.74	1.20	.003930	8.10	12.97	5.53	34.00	574.30	2262.00	7.23	41.71	1206.05	8303.10	28.13	.00033
41.30	30.35	.00195	-.00001	.00031	.00227	41.97	34.15	8.10	.0813	40.73	26.54	5.56	1.271	6.25	
7.00	4.95	1.09	.004330	7.00	15.00	6.20	23.40	598.70	1097.00	7.13	43.38	1058.00	8303.17	24.53	
NO SOLUTION FLASHING															
42.59	22.98	.00243	.00001	0.00000	.00244	43.38	23.40	3.79	.0738	41.99	22.57	3.97	3.122	9.24	.00041
7.00	4.95	1.10	.004330	7.05	14.97	6.10	23.70	735.30	1082.00	7.03	43.12	1104.00	8303.17	24.67	
NO SOLUTION FLASHING															
42.43	23.27	.00246	.00001	0.00000	.00247	43.12	23.70	3.93	.0748	41.74	22.83	4.11	3.029	8.80	.00043
7.00	4.99	1.09	.004330	7.25	15.00	6.07	24.17	704.70	1111.00	7.00	42.77	1112.00	8303.17	24.90	
NO SOLUTION FLASHING															
42.08	23.54	.00247	.00000	0.00000	.00247	42.77	24.17	4.15	.0740	41.39	23.11	4.29	3.030	8.58	.00043

Evaporator-Absorber Data and Results

7.01	4.99	1.08	.004330	7.37	14.98	5.98	24.60	708.50	1126.00	6.98	42.09	1127.00	8303.17	25.20	
NO SOLUTION FLASHING															
5.16	7.01	.00247	.000000	0.00000	.00247	42.09	24.60	4.48	.0732	40.71	23.52	4.61	2.826	7.66	.00046
41.40	24.05	9.49													
7.01	4.99	1.07	.004330	7.45	15.05	5.90	24.85	704.00	1172.00	6.90	41.73	1135.00	8303.17	24.98	
NO SOLUTION FLASHING															
5.16	7.27	.00244	.000000	0.00000	.00245	41.73	24.85	4.67	.0725	40.36	23.31	4.66	2.786	7.63	.00046
41.05	24.04	9.55													
7.00	4.97	1.05	.004250	7.65	15.00	5.83	25.49	712.50	1145.00	6.88	41.01	1151.00	8303.17	24.63	
NO SOLUTION FLASHING															
5.19	7.13	.00248	.000000	0.00000	.00248	41.01	25.49	5.11	.0710	39.63	23.13	4.84	2.677	7.43	.00049
40.32	24.36	9.45													
6.94	3.54	1.08	.003910	8.30	20.37	6.37	35.60	733.00	1794.00	7.50	45.29	1310.00	8303.24	30.97	
NO SOLUTION FLASHING															
5.34	7.54	.00234	.000001	0.00000	.00253	45.29	35.60	6.84	.0739	43.79	29.27	5.28	2.442	8.15	.00054
44.54	32.44	10.62													
7.00	3.60	1.04	.004000	8.05	18.43	6.17	35.63	687.00	1777.00	7.23	45.07	1315.00	8303.24	30.97	
NO SOLUTION FLASHING															
5.32	7.72	.00237	.000000	0.00000	.00237	45.07	35.63	6.96	.0734	43.67	29.13	5.28	1.927	7.18	.00064
44.37	32.38	10.47													
7.00	3.95	2.10	.003610	7.25	16.13	5.40	30.50	709.00	2145.00	6.77	43.07	1355.00	8303.24	27.90	
NO SOLUTION FLASHING															
5.14	7.45	.00245	.000000	0.00000	.00247	43.07	30.50	5.99	.1427	42.34	27.28	5.20	1.657	5.55	.00073
42.70	24.44	9.85													
6.95	3.92	2.07	.003920	7.25	16.44	5.56	31.64	665.40	2230.00	6.74	42.48	1410.00	8303.24	27.10	
NO SOLUTION FLASHING															
5.72	7.55	.00224	.000000	0.00000	.00230	42.48	31.64	6.69	.1405	41.80	26.24	5.07	1.372	5.94	.00085
42.14	28.94	9.32													
6.40	3.30	2.18	.003810	8.50	20.53	6.23	38.13	734.70	2394.00	7.60	45.20	1300.00	8303.29	30.10	
NO SOLUTION FLASHING															
5.22	7.41	.00250	.000001	0.00000	.00249	45.20	38.13	7.95	.1490	44.46	29.26	5.02	2.018	9.06	.00065
44.43	33.54	10.54													
6.40	3.51	2.16	.004080	8.07	20.24	6.15	38.80	750.50	2446.00	7.58	44.97	1310.00	8303.29	29.40	
NO SOLUTION FLASHING															
5.40	10.27	.00257	.000000	0.00010	.00266	45.00	38.13	8.07	.1476	44.20	28.45	4.87	1.599	8.61	.00037
44.60	33.34	10.43													
6.45	3.50	2.13	.004100	8.10	20.29	6.00	34.43	743.70	2429.00	7.48	44.80	1317.00	8303.29	28.98	
NO SOLUTION FLASHING															
5.35	10.26	.00255	.000000	0.00037	.00261	44.32	37.95	8.10	.1455	44.03	28.00	4.90	1.650	8.90	.00043
44.43	32.94	10.35													
6.40	4.00	2.10	.003880	7.53	19.60	5.80	37.20	744.70	2096.00	7.20	43.72	1344.00	8303.29	27.63	
NO SOLUTION FLASHING															
5.36	10.23	.00253	.000000	0.00023	.00277	43.73	35.54	7.63	.1430	42.96	26.84	4.84	1.396	7.64	.00104
43.37	31.17	9.45													
7.43	5.73	2.17	.004050	7.15	18.43	5.27	39.70	975.00	2593.00	7.93	44.36	1213.00	8303.30	29.83	
NO SOLUTION FLASHING															
7.43	11.40	.00294	.000001	0.00057	.00355	44.53	35.38	7.15	.1480	43.48	29.12	5.35	1.900	4.90	.00207
44.01	32.25	10.16													
6.40	3.75	2.16	.004250	7.20	19.13	5.20	35.70	848.00	2705.00	7.93	44.16	1219.00	8303.30	29.33	
NO SOLUTION FLASHING															
7.20	12.25	.00289	.000001	0.00053	.00349	44.34	35.24	7.20	.1473	43.31	29.18	5.44	1.881	4.74	.00207
43.82	32.21	9.99													

Evaporator-Absorber Data and Results

5.90	3.25	2.15	.000700	8.60	18.73	5.13	40.00	815.70	2738.00	7.63	43.76	1231.00	8303.30	32.00	
5.93	13.57	.00275	-.00001	.00033	.00310	43.26	37.73	8.60	.1468	42.95	30.42	6.01	1.293	6.13	.00125
43.40	34.07	9.14													
5.90	3.25	2.15	.000700	8.60	18.40	5.00	40.00	824.00	2731.00	7.60	43.57	1238.00	8303.30	31.98	
7.01	10.35	.00281	-.00001	.00035	.00316	43.69	37.49	9.50	.1463	42.75	30.54	6.15	1.227	5.75	.00135
43.21	34.02	8.99													
5.90	3.43	2.13	.004250	7.00	19.07	5.77	38.17	855.70	2823.00	7.67	42.34	1312.00	8303.30	28.17	
7.29	11.54	.00233	.00001	.00065	.00357	43.14	33.22	7.00	.1447	42.09	27.50	5.36	.819	4.43	.00230
42.61	30.36	9.34													
5.90	3.43	2.13	.004120	7.00	17.90	5.73	38.00	851.70	2789.00	7.63	42.52	1316.00	8303.30	27.73	
7.25	11.52	.00231	.00001	.00063	.00361	42.42	32.83	7.00	.1443	41.77	27.13	5.36	.821	4.43	.00230
42.30	29.94	9.20													
5.90	3.53	2.13	.004330	7.00	17.60	5.70	37.90	832.30	2800.00	7.60	42.43	1323.00	8303.30	27.33	
7.05	11.75	.00235	.00001	.00071	.00355	42.64	32.60	7.00	.1445	41.61	26.50	5.25	.876	4.75	.00213
42.12	29.50	9.14													
5.90	2.55	2.12	.004330	7.15	15.87	5.90	37.23	690.30	3447.00	7.70	42.31	1338.00	8303.30	27.33	
5.90	10.05	.00237	.00001	.00064	.00302	42.20	32.42	7.15	.1437	41.33	26.38	5.28	.935	5.00	.00169
41.76	29.40	8.88													
5.95	2.55	2.11	.004330	6.90	13.60	5.90	36.63	562.70	3162.00	7.77	41.55	1350.00	8303.30	26.07	
4.85	9.22	.00135	.00001	.00070	.00365	41.45	31.41	5.90	.1429	41.09	24.93	4.91	.996	5.57	.00140
41.47	23.17	8.93													
5.95	2.50	2.11	.004250	6.80	13.03	5.90	36.30	544.70	3062.00	7.70	41.44	1356.00	8303.30	25.43	
4.71	8.74	.00134	.00002	.00072	.00262	41.55	30.92	5.80	.1424	40.89	24.31	4.78	1.008	5.72	.00136
41.27	27.51	8.39													
5.95	4.14	2.21	.003830	7.40	17.07	5.83	34.83	631.30	1780.00	7.50	41.35	1234.00	8304.07	27.60	
5.41	8.43	.00217	.00000	.00032	.00249	41.94	32.59	7.40	.1497	41.25	26.32	5.46	.972	5.05	.00134
41.53	29.75	8.64													
5.90	3.80	2.20	.004000	7.50	16.90	5.47	34.17	702.30	2035.00	6.73	41.46	1303.00	8304.07	25.77	
5.93	9.33	.00240	-.00000	.00026	.00266	41.53	32.43	7.50	.1489	40.80	26.04	5.36	1.069	5.53	.00130
41.17	23.23	8.43													
5.90	3.80	2.20	.004000	7.55	16.83	5.33	33.33	701.70	2014.00	6.67	41.39	1309.00	8304.07	26.63	
5.93	9.30	.00240	-.00000	.00022	.00262	41.45	32.45	7.55	.1488	40.73	25.87	5.33	1.109	5.73	.00124
41.09	23.14	8.39													
6.80	3.75	2.20	.004000	7.50	16.87	5.33	33.47	710.70	2011.00	6.50	41.30	1316.30	8304.07	25.47	
6.06	9.15	.00243	-.00000	.00019	.00261	41.35	32.22	7.50	.1488	40.64	25.76	5.33	1.085	5.62	.00126
40.99	28.59	8.35													
6.90	3.80	2.20	.004000	7.50	16.90	5.40	33.30	702.30	1992.00	6.53	41.23	1321.30	8304.07	25.30	
5.99	9.20	.00240	-.00000	.00017	.00257	41.29	32.13	7.50	.1488	40.58	25.53	5.29	1.104	5.73	.00122
40.93	28.45	8.33													
5.90	3.71	1.17	.003830	8.05	17.15	5.25	33.00	669.50	1669.00	6.35	41.21	1335.00	8304.07	26.60	
NO SOLUTION FLASHING															
5.70	7.40	.00223	-.00001	0.00000	.00228	41.21	33.00	7.93	.0791	40.06	25.37	5.41	1.379	6.53	.00057
40.63	29.18	8.03													
5.90	3.50	1.16	.004000	8.10	16.90	5.27	31.93	682.70	1665.00	6.30	41.19	1346.00	8304.07	25.77	
NO SOLUTION FLASHING															
5.82	7.25	.00233	-.00001	0.00000	.00232	41.13	31.93	7.46	.0784	40.01	24.41	5.12	1.813	7.53	.00067
40.50	28.17	8.22													

Evaporator-Absorber Data and Results

5.30	3.75	1.16	.004000	8.10	17.03	5.27	31.90	670.70	1658.00	6.33	41.11	1352.00	8304.07	25.53	
NO SOLUTION FLASHING															
5.71	7.05	.00229	-.00001	0.00000	.00029	41.11	31.90	7.44	.0784	39.95	24.25	5.08	1.838	7.63	.00065
40.53	24.02	8.20													
5.90	2.90	1.15	.004000	8.40	17.27	5.30	31.80	645.70	1999.00	6.40	40.95	1358.00	8304.07	25.23	
NO SOLUTION FLASHING															
5.50	6.41	.00221	-.00001	0.00000	.00219	40.95	31.80	7.53	.0784	39.84	24.79	5.30	1.987	7.57	.00058
40.35	24.30	8.03													
5.90	2.90	1.15	.004120	8.55	17.20	5.37	31.90	641.00	2026.00	6.50	40.74	1406.00	8303.07	25.53	
NO SOLUTION FLASHING															
5.46	5.47	.00219	-.00002	0.00000	.00217	40.74	31.90	7.68	.0783	39.64	25.02	5.45	1.985	7.42	.00257
40.19	26.45	7.85													
5.90	2.90	1.15	.004000	8.55	17.13	5.33	31.90	650.30	2096.00	6.43	40.60	1412.00	8304.07	25.60	
NO SOLUTION FLASHING															
5.54	6.46	.00222	-.00002	0.00000	.00221	40.60	31.90	7.76	.0783	39.49	25.23	5.58	1.884	7.03	.00061
40.04	28.57	7.72													
5.90	2.90	1.15	.004000	8.60	17.10	5.20	31.77	641.30	2076.00	6.37	40.39	1420.31	8304.07	25.43	
NO SOLUTION FLASHING															
5.45	7.07	.00219	-.00002	0.00000	.00217	40.39	31.77	7.81	.0776	39.29	25.05	5.59	1.902	7.09	.00060
39.54	29.41	7.60													
5.90	2.90	1.15	.004000	8.55	17.27	5.30	31.60	641.70	2015.00	6.40	40.07	1428.31	8304.07	25.33	
NO SOLUTION FLASHING															
5.46	5.47	.00219	-.00002	0.00000	.00218	40.07	31.60	7.90	.0775	38.98	25.00	5.69	1.857	6.99	.00061
39.32	28.30	7.40													
5.95	3.40	2.10	.004000	7.40	15.80	6.60	40.77	623.30	1835.00	8.23	46.25	1231.30	8304.25	33.93	
NO SOLUTION FLASHING															
4.38	4.43	.00215	.00001	.00029	.00245	42.34	39.45	7.40	.1440	45.57	32.72	5.66	.868	4.65	.00147
40.96	35.44	11.10													
5.95	3.40	2.09	.003430	7.43	16.47	6.60	39.61	625.70	1742.30	8.30	46.07	1240.30	8304.25	33.10	
NO SOLUTION FLASHING															
4.37	4.43	.00215	.00001	.00020	.00236	45.13	34.23	7.43	.1425	45.38	32.00	5.51	.962	5.13	.00123
40.75	35.11	11.04													
5.95	3.40	2.08	.003430	7.50	16.63	6.60	39.50	644.70	1842.00	8.30	45.94	1247.30	8304.25	32.73	
NO SOLUTION FLASHING															
4.34	4.43	.00222	.00001	.00018	.00241	42.00	38.20	7.50	.1425	45.23	31.67	5.46	1.018	5.48	.00124
40.52	34.94	10.95													
5.95	3.40	2.07	.003430	7.55	16.57	6.63	39.60	633.00	1861.00	8.13	45.83	1253.00	8304.25	32.27	
NO SOLUTION FLASHING															
4.49	4.57	.00220	.00001	.00020	.00241	43.84	38.13	7.55	.1418	45.13	31.14	5.34	1.107	5.99	.00114
40.51	34.56	10.93													
5.95	3.40	2.07	.003430	7.55	16.43	6.60	39.77	673.30	1906.00	8.10	45.67	1301.00	8304.25	31.97	
NO SOLUTION FLASHING															
4.34	4.43	.00232	.00001	.00025	.00254	45.75	37.95	7.55	.1417	44.93	30.84	5.33	1.108	5.99	.00122
40.34	34.43	10.83													
5.95	3.40	2.06	.003430	7.48	16.43	6.50	39.93	661.00	1938.00	8.23	45.56	1307.00	8304.25	31.80	
NO SOLUTION FLASHING															
4.34	4.43	.00226	.00001	.00029	.00258	45.65	37.68	7.48	.1406	44.83	30.68	5.31	1.084	5.88	.00125
40.14	34.14	10.81													
5.95	3.40	2.05	.003430	7.50	16.43	6.60	39.77	664.70	1970.00	8.20	45.44	1313.00	8304.25	31.40	
NO SOLUTION FLASHING															
4.34	4.43	.00229	.00001	.00030	.00260	45.54	37.57	7.50	.1402	44.71	30.29	5.24	1.131	6.15	.00120
40.12	33.95	10.76													
5.95	3.40	2.03	.003430	7.50	16.47	6.50	39.40	653.00	1809.00	8.13	44.83	1328.00	8304.25	30.77	
NO SOLUTION FLASHING															
4.34	4.43	.00227	.00001	.00033	.00260	44.94	35.75	7.50	.1386	44.11	29.68	5.28	1.108	5.97	.00123
40.02	33.21	10.58													

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