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Oliver Ziegler
Technische Universität Dresden, Germany, oliver.ziegler@tu-dresden.de

Ullrich Hesse
ullrich.hesse@tu-dresden.de

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Theoretical and Experimental Investigations on the Recirculation of the Solution (Mix-Flow) in a NH₃-H₂O-Absorption Resorption Refrigeration System

Oliver ZIEGLER[1]*, Ullrich HESSE[1]

[1]Technische Universität Dresden,
Bitzer-Chair of Refrigeration, Cryo and Compressor Technology,
Dresden, 01062, Germany
+49 351 463 32603, oliver.ziegler@tu-dresden.de
+49 351 463 32548, ullrich.hesse@tu-dresden.de

* Corresponding Author

ABSTRACT

Ammonia water absorption resorption refrigeration systems do not need a rectification column like conventional absorption refrigeration systems do. In this case such systems need a recirculation of the solution. This so called "mix flow" is necessary to prevent the dilution of the solution in the "cold cycle" of the system which is caused by the relative high concentration of water in the hot vapor. The influence of this recirculation of the solution on the performance of the system could be shown in previous studies and is now modelled and investigated more in detail. As a result of the theoretical investigation, a preferred internal piping layout is proposed. In addition, the theoretical foundations are verified with experimental investigations at an existing test cycle.

1. INTRODUCTION

Compared to conventional ammonia-water-absorption chillers, the construction of resorption chillers is characterized by the fact that a second solution cycle is installed instead of the conventional components, evaporator and condenser (cf. Figure 1). The background to this configuration is that no rectification of the hot vapor is necessary, since the water-containing hot vapor is absorbed by the ammonia-water-solution in the resorber (also called cold absorber) and ammonia is therefore not present anywhere in the plant as a pure substance. A resulting advantage is that the design pressure of NH₃-H₂O resorption plants is significantly lower (< 0.6 MPa) than that of conventional absorption plants (> 1.2 MPa), which has an effect on investment costs and plant safety.

During the continuous operation of the resorption plant, the relatively high water concentration in the hot vapor causes a mass shift between both solution cycles and thus a dilution of the solution in the cold cycle. As a result of the decrease of the NH₃ concentration in the cold cycle, the evaporating pressure level in the cold desorber increases which is reflected in a reduction of the achievable cooling capacity. Thus it must be ensured that a mass balance between both cycles is achieved, which is based to the concentration of the water in the hot vapor. This is realized by a recirculation of the solution from the cold to the hot cycle (also called mix flow).

In previous experimental studies the effects of too low or too high mix flows on the parameters of a resorption system could be shown by ZIEGLER et al. 2017 [1]. The theoretical fundamentals for adjusting the correct volume flow of the mix flow will now be presented and compared to the previous experimental results.

2. NH₃-H₂O RESORPTION REFRIGERATION SYSTEM

The structure of a NH₃-H₂O resorption refrigeration system (RRS) is based on two interconnected solution cycles. The setup of these two cycles is similar to the setup of an absorption cycle. In terms of the temperatures inside the desorbers the two cycles are referred as "hot" and "cold" cycle. The scheme of a resorption system with its components is shown in Figure 1.
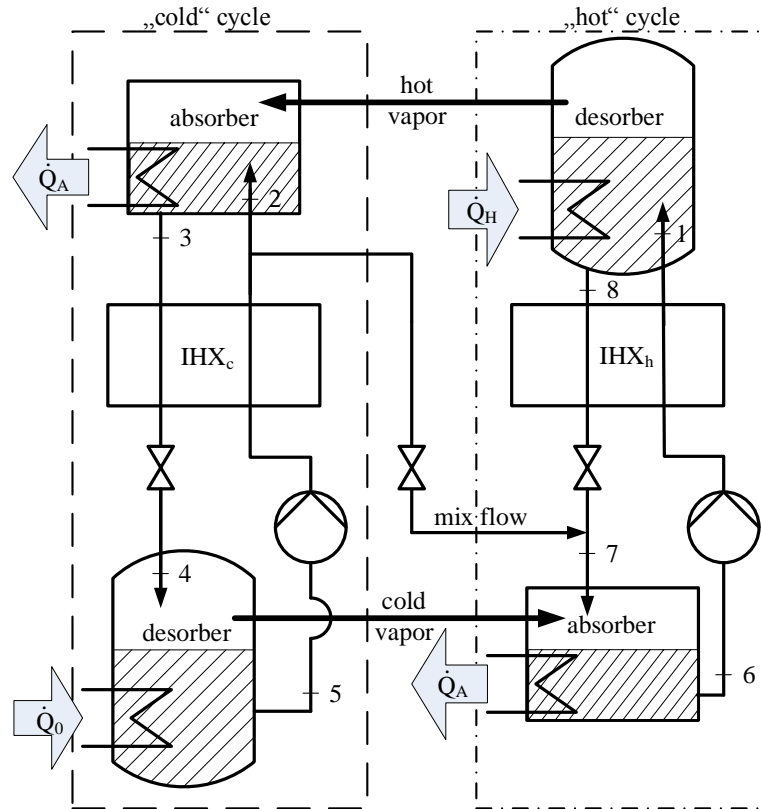
Figure 1: scheme of the resorption system

The driving energy for the process is (waste) heat ($\dot{Q}_H$) supplied to the "hot" desorber at a high temperature level between 70-95°C. This results in a desorption of the refrigerant (ammonia) from the solution in the hot desorber. The gaseous refrigerant is fed to the "cold" absorber. The poor solution in the hot cycle (8) is cooled down in an internal heat exchanger (IHX$_h$) and the pressure is reduced by an expansion valve before it is supplied to the "hot" absorber (7). Due to the negative solution enthalpy during the absorption of ammonia in water, it is necessary to cool both absorbers in order to achieve a better solubility. In the "cold" absorber the refrigerant vapor is absorbed by the solvent with heat removal ($\dot{Q}_A$). In the IHX$_c$ of the "cold" cycle, the rich solution coming from the cold absorber (3) is cooled and then throttled and fed to the cold desorber (4). Due to the high concentration of ammonia in the solution in the cold desorber an expulsion of the refrigerant at very low temperatures (< 0°C) and low pressure (0.07-0.15 MPa) takes place. The energy required for the expulsion is the available cooling capacity of the resorption system ($\dot{Q}_0$). A solvent pump increases the pressure and transports the poor cold solution (5) through the cold IHX$_c$ back to the absorber (2), where the refrigerant vapor is resumed again. The cold refrigerant vapor is fed to the hot absorber, where the vapor is absorbed by the solution similar to the process in the cold absorber. The enriched solution in the hot absorber (6) is fed by a further solvent pump to the hot desorber (1).

The working process of the resorption cycle was described by Grund et al. 2015 [2] as follows and is also shown in Figure 2 in a log p -1/T-diagram.
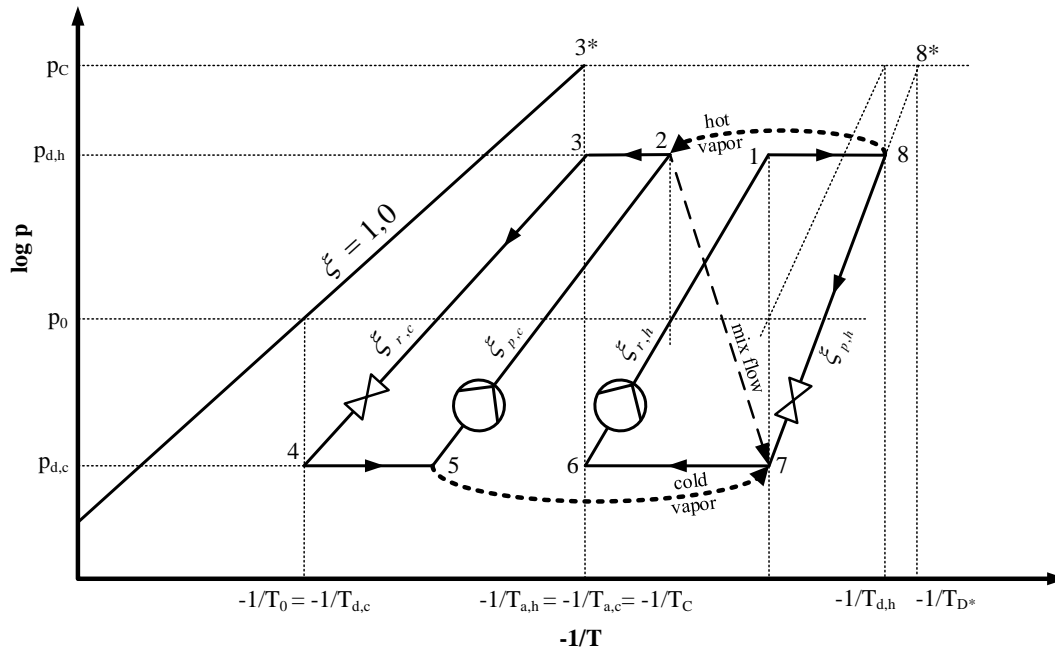
Figure 2: The resorption process in the log p - 1/T diagram

The solution cycle of a conventional absorption chiller is modelled on the polyline 6-1-8-7. The evaporation and condensation process of a regular absorption machine is also represented by a solvent cycle. The resulting process is represented by the polyline 2-3-4-5. Based on the prevailing temperature levels in the desorbers each polyline can be assigned to the "hot" (6-1-8-7) or the "cold" (2-3-4-5) cycle. The highest temperature in the entire resorption process ($T_{d,h}$) prevails in the hot desorber and depends on the available heat source. The outlet temperatures of the solution of the two absorbers ($T_{a,h}$, $T_{a,c}$) can be considered as almost identical, due to the usage of the same source of cooling water with a parallel throughflow. The lowest temperature of the entire system occurs at the entrance of the cold desorber ($T_{d,c}$).

The advantages of the resorption technology can be illustrated in the log p / -1/T diagram shown in Figure 2. The reference for the comparison is a conventional absorption system with one solution cycle and the components evaporator and condenser. If the absorption (or resorption) of ammonia in the cold absorber ($2\rightarrow3$) is compared to the liquefaction of NH_3 at the same temperature ($T_{a,h} = T_C$), a substantially lower pressure in the absorber and resorber is required than it would be the case in the condenser ($p_C > p_{a,c}$). This means for the entire process, that a lower high-pressure level is required in comparison to a conventional absorption machine. Conversely this means, that a much lower pressure in the cold desorber ($p_{d,c}$) is necessary than in an evaporator (p_0) in providing the cooling capacity at the same temperature level as in the compared process ($T_0 = T_{d,c}$). The expulsion of the refrigerant in the cold cycle partially takes place at pressures below the atmospheric pressure. For the available working agents of absorption/resorption the pressure difference between an evaporator and the cold desorber ($p_0 - p_{d,c}$) will always be smaller than the difference between a condenser and the cold absorber ($p_C - p_{a,c}$). Also the pressure difference in the resorption process is smaller than the difference in the absorption process (Plank & Niebergall 1959 [3]).

The prevailing temperatures in the cycle, the mentioned temperature ranges for the expulsion and the absorption can be recognized with the assumption of a constant concentration (ξ) of refrigerant in the solution between the absorbers and desorbers of each cycle. The temperature limit for the absorption in the cold cycle is point of state 3 in Figure 2 and for the hot solution cycle the point of state 6 is the temperature limit. Also the temperature range where the expulsion takes place is limited by the points 1 and 8 on the hot and 4 and 5 on the cold side. Furthermore it can be seen that a lower temperature for the expulsion in the resorption process ($T_{d,h}$) is required than for the desorption of a conventional absorption process (T_{D*}) with the same temperature level ($T_C = T_{a,h}$). These considerations apply in the case of the same concentration of the solution inside the compared processes. Further it can be seen that a conventional absorption process with the same temperature of the desorption and absorption has a smaller degassing range. This results in bigger volume flow rates of the reference cycle. The irreversible losses of an absorption cycle can be reduced due to the fact that the absorption and desorption process takes place with a temperature glide. That

means a lower flow of cooling water is required and higher temperatures of the return flow of the cooling water can be realized.

3. RECIRCULATION OF THE SOLUTION (MIX FLOW)

2.1 Background

As a result of the elimination of rectification, there is a mass shift between the cycles during the operation of resorption systems, which among other effects has an impact on the attainable cooling capacity of the RRS. To counteract this mass shift, a recirculation of the solution (mix flow) from the cold to the hot cycle is used. The size of the mix flow depends on various boundary conditions. The significant variables are the heating temperature $T_{d,h}$ and the pressure $p_{d,h}$ at the outlet of the hot desorber. Assuming that the two-phase NH_3-H_2O mixture is in equilibrium at the outlet of the desorbers, the heating and cooling temperatures in particular determine the proportion of water in the hot vapor. The water contained in the hot vapor thus leads to a dilution of the solution in the cold cycle, since due to the low temperature, no water is returned to the hot cycle via the cold vapor. To illustrate this, Table 1 shows the NH_3 concentrations for some boundary conditions, which were calculated using the fluid property library LibAmWa of the University of Applied Sciences Zittau-Görlitz.

Table 1: NH_3 concentration of hot and cold vapor for selected operating points of a RRS

Measurement Point	Temp [°C]	Pressure [bar]	$\xi_{h,vapor}$ [kg$_{NH3}$/kg]	Measurement Point	Temp [°C]	Pressure [bar]	$\xi_{c,vapor}$ [kg$_{NH3}$/kg]
Hot Desorber Out	70	4	0,947	Cold Desorber Out	-6	0,5	0,997
Hot Desorber Out	80	4	0,910	Cold Desorber Out	3	0,5	0,993
Hot Desorber Out	90	4	0,854	Cold Desorber Out	12	0,5	0,984
Hot Desorber Out	70	5	0,961	Cold Desorber Out	-6	1	0,999
Hot Desorber Out	80	5	0,933	Cold Desorber Out	3	1	0,998
Hot Desorber Out	90	5	0,890	Cold Desorber Out	12	1	0,994
Hot Desorber Out	70	6	0,970	Cold Desorber Out	-6	1,5	1,0
Hot Desorber Out	80	6	0,947	Cold Desorber Out	3	1,5	0,999
Hot Desorber Out	90	6	0,913	Cold Desorber Out	12	1,5	0,997
Hot Desorber Out	70	7	0,976	Cold Desorber Out	-6	2	1,0
Hot Desorber Out	80	7	0,958	Cold Desorber Out	3	2	0,999
Hot Desorber Out	90	7	0,929	Cold Desorber Out	12	2	0,998

2.2 Calculation of the mix flow for a continuous operation

In order to be able to operate the RRS continuously, a different mix flow results depending on the external boundary conditions, provided that the masses in both cycles are kept constant

$$m_c = m_h = \frac{m_{fill}}{2} \tag{1}$$

These masses consist of a fraction of water and a fraction of ammonia

$$m_c = m_{c,H2O} + m_{c,NH3} \tag{2}$$
$$m_h = m_{h,H2O} + m_{h,NH3} \tag{3}$$

This can be used to determine the concentration of ammonia in the respective partial cycle

$$\xi_{h,NH3} = \frac{m_{c,NH3}}{(m_{c,H2O} + m_{c,NH3})} \tag{4}$$

$$\xi_{c,NH3} = \frac{m_{h,NH3}}{(m_{h,H2O} + m_{h,NH3})} \tag{5}$$

Assuming that the volumes of vessels, apparatus and pipes in the partial cycles are identical

$$V_c = V_h \tag{7}$$

and the liquid levels of the vessels after the desorbers and absorbers are kept constant during operation by suitable level controllers

$$L_{c,d} = L_{c,a} \tag{8}$$
$$L_{h,d} = L_{h,a} \tag{9}$$

the concentration in the respective partial cycle can be determined under certain boundary conditions. The ammonia concentration of the cold cycle $\xi_{c,NH3}$ results directly from the concentration of the hot cycle and the filling concentration ξ_{fill} by means of the following correlation

$$m_{fill} * \xi_{fill} = \frac{m_h * \xi_h + m_c * \xi_c}{2} \tag{10}$$

With the outgassing width $\Delta\xi_{h,d}$ in the hot desorber and the saturation concentration of the solution at the outlet of the hot desorber

$$\xi_{p,h,out,liq} = f(T_{h,d,out}; p_{h,d,out}) \tag{11}$$

all concentrations in the partial cycles are determined

$$\xi_h = \xi_{p,h} + \frac{\Delta\xi_{h,d}}{2} \tag{12}$$

$$\xi_{r,h} = \xi_{p,h} + \Delta\xi_{h,d} \tag{13}$$

$$\xi_{r,c} = \xi_c + \frac{\Delta\xi_{h,d}}{2} \tag{14}$$

$$\xi_{p,c} = \xi_c - \frac{\Delta\xi_{h,d}}{2} \tag{15}$$

Furthermore, it is assumed that the volume flow of the poor hot solution is limited to the value $\dot{V}_{p,h}$ by a throttling element in the type of a flow limiter, which corresponds to a mass flow of $\dot{m}_{p,h}$. With the heat flow $\dot{Q}_{h,d}$, received by the fluid in the hot desorber, the enthalpy change can be determined at a temperature difference between inlet and outlet of the hot desorber $\Delta T_{h,d}$ by means of a suitable fluid property library. For this purpose, the enthalpy of the fluid flows must be determined at the outlet $h_{h,d,out,liq}$ and $h_{h,d,out,vap}$ and at the inlet of the hot desorber $h_{h,d,in,liq}$, assuming phase equilibrium. The received heat flow in the desorber $\dot{Q}_{h,d}$ consists of an evaporation $\dot{Q}_{lat}$ and a warming $\dot{Q}_{sens}$ of the solution, which are influenced by the temperature glide when ammonia is expelled. This results in a latent heat flow of

$$\dot{Q}_{lat} = \dot{Q}_{h,d} - \dot{Q}_{sens} \tag{16}$$

The sensible heat flow $\dot{Q}_{sens}$, which is necessary for the temperature increase of the liquid phase, can be calculated by the enthalpy change between inlet and outlet of the hot desorber, assuming that the fluid enters the hot desorber in phase equilibrium with a vapor fraction of

$$x_{h,d,in} = 0 \tag{17}$$
$$\dot{Q}_{sens} = \dot{m}_{p,h} * (h_{h,d,out,liq} - h_{h,d,in,liq}) \tag{18}$$

Now the mass flow of hot vapor can be determined to

$$\dot{m}_{h,vap} = \frac{\dot{Q}_{lat}}{h_{h,d,out,vap} - h_{h,d,in,liq}} \tag{19}$$

The mass flow of the rich solution $\dot{m}_{r,h}$ is not equal to the mass flow of the poor solution after the hot desorber $\dot{m}_{p,h}$, but is higher by the amount of the expelled vapor. This must be determined iteratively. Further correlations are necessary for this. One of the necessary assumptions is that the mass flow of hot vapor is equal to the sum of the mass flows of the mix flow and cold vapor in order not to violate the mass balance of the RRS between the two partial cycles.

$$\dot{m}_{h,vap} = \dot{m}_{c,vap} + \dot{m}_{mix} \tag{20}$$

Since the fraction of water in the cold vapor can be assumed to be

$$\dot{m}_{c,vap,H2O} \approx 0 \tag{21}$$

and the concentration of the mix flow corresponds to the concentration of the poor solution in the cold cycle

$$\xi_{mix} = \xi_{p,c} \tag{22}$$

the required mix flow results from the water fraction of the hot vapor to

$$\dot{m}_{mix} = \frac{\dot{m}_{h,vap} * (1 - \xi_{h,vap})}{(1 - \xi_{mix})} \tag{23}$$

2.3 Results of the calculation for the mix flow of selected operating points of a RRS

To determine the required amount of the mix flow, a simulation model was created that contains the correlations from 2.2. This was necessary, among other aspects, to carry out a pre-dimensioning of the necessary measuring device for determining the hot vapor volume flow. The RRS available at the TU Dresden is operated with the boundary conditions shown in Table 2, which also are used as a basis for the calculation of the mix flow.

Table 2: Boundary and operating conditions of the RRS of the TU Dresden

	Value	Unit
$\dot{Q}_{h,d}$	42	kW
$\Delta T_{h,d}$	10	K
ξ_{fill}	0,4	kg_{NH3}/kg
$\dot{V}_{p,h}$	20	l/min
$p_{h,d,out}$	3…7	bar
$T_{h,d,out}$	70…90	°C
T_{mix}	15	°C

A variation of the temperature at constant pressure or a variation of the pressure at constant temperature therefore results in the hot vapor flow during the operation of the RRS as shown in Figure 3 and 4.
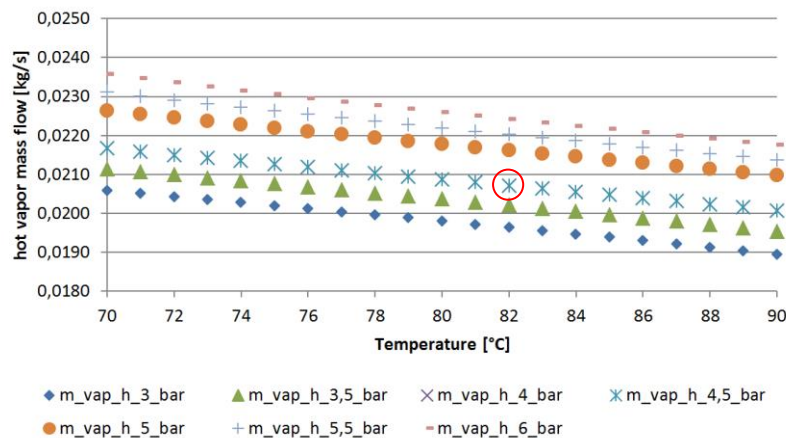

Figure 3: Hot vapor flow for constant pressure and various temperatures
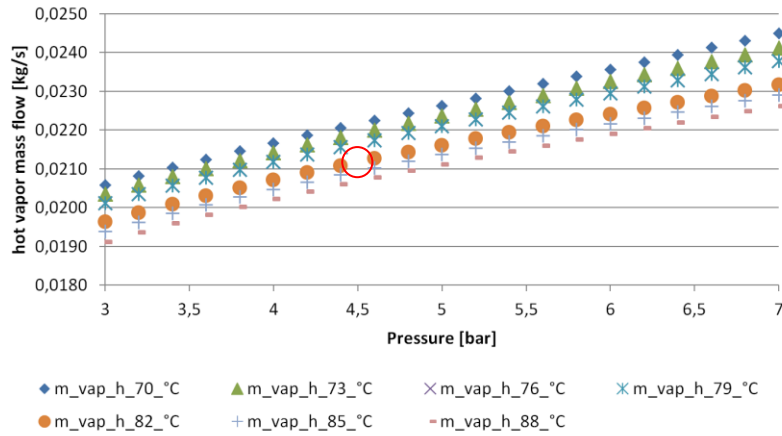
Figure 4: Hot vapor flow for constant temperature and various pressures

As expected, the mass flow of hot vapor increases due to an increase in temperature or a decrease in pressure levels. In connection with the water content of the hot vapor, the necessary mix flow results, which are shown in figures 5 and 6.
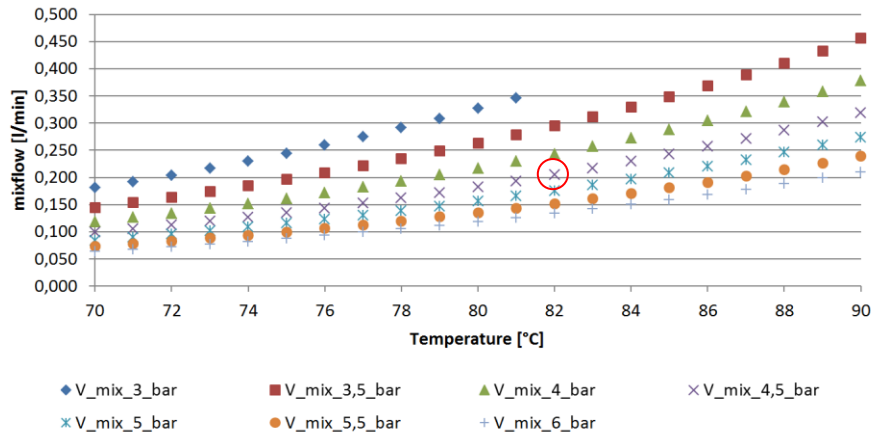

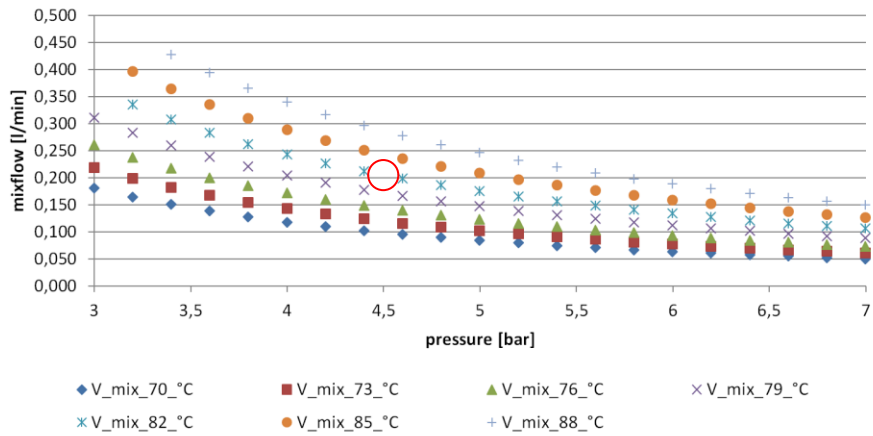**Figure 5:** Mix flow for constant pressure and various temperatures

Figure 6: Mix flow for constant temperature and various pressures

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The required mix flow increases due to an increase in temperature or a decrease in pressure, but seems to adjust asymptotically to a minimum for high pressures or low temperatures.

The red circles in Figures 3 - 6 show a specific measuring point, which is discussed in more detail in Section 5.

4. EXPERIMENTAL PLANT

4.1 Test Cycle and Modification
For the verification of the presented calculation results an RRS at the TU Dresden is available, which already was used by GRUND and ZIEGLER et al. for measurements. Due to the deviations of calculation results from the real plant, this was modified and extended to verify them (cf. Figure 8).
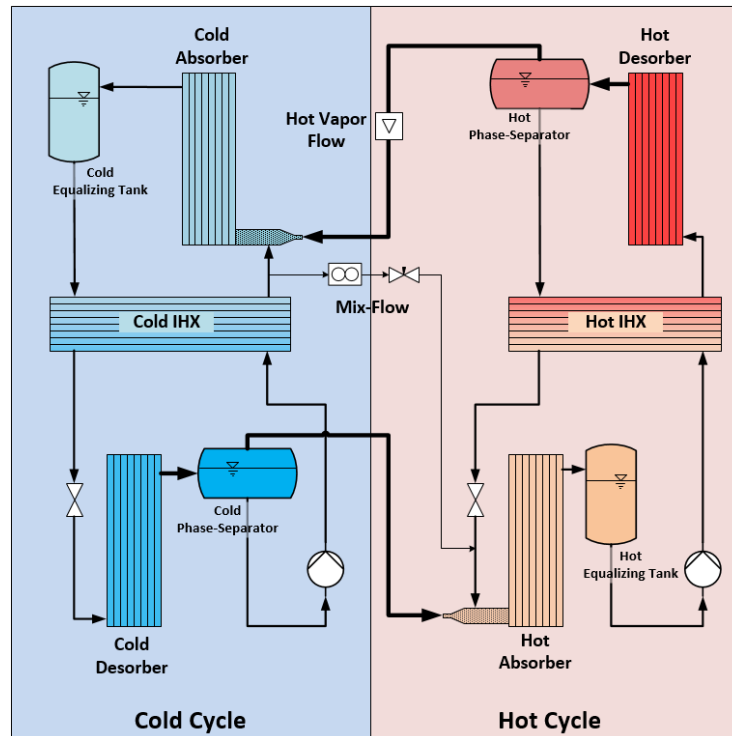

Figure 8: Process Flow Diagram of the Resorption Refrigeration System (TU Dresden)

4.2 Mix Flow
As with the system at the TU Dresden, a bypass from the high pressure area of the cold cycle to the low pressure area of the hot cycle is preferred for the mix flow in order to avoid additional equipment. The driving force for maintaining the flow of the mix flow is the pressure difference between high and low pressure. To adjust the mix flow correctly, it is necessary that the mix flow is a liquid fluid, as the volume flow is measured by using an oval gear meter.

4.3 Hot Vapor Flow
To measure the vapor volume flow, a device was obtained with which the volume flow of the hot vapor can be determined via the differential pressure due to the flow velocity. This V-cone flow meter is used in the hot vapor pipe between the outlet of the hot desorber vessel and the inlet of the cold absorber.

5. EXPERIMENTAL DATA

Previous experimental results from ZIEGLER et al. 2017 [1] with the RRS at the TUD have shown that under comparable boundary conditions a mix flow of approx. 0.2 l/min is necessary (cf. Figure 7 & 8) to ensure a continuous operation of the system, which corresponds to the results of the calculation model here presented.

Furthermore, the second graph, with a flow rate of 0.5 l/min, shows that a mix flow that is not optimally adjusted leads to a shift in concentration (cf. Figure 7) and a mass shift (cf. Figure 8).
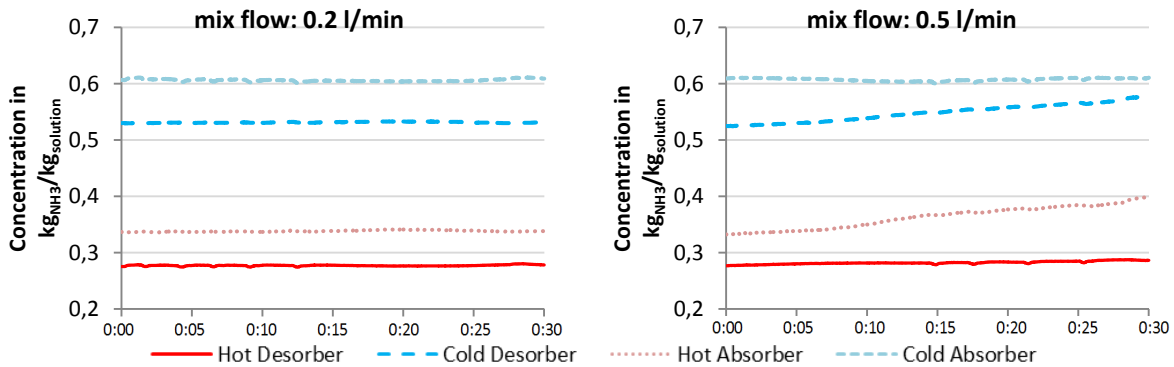

Figure 7: NH_3 concentrations of the solution in the absorber and desorber vessels during continuous operation of the TUD-RRS with a mix flow of 0.2 l/min and 0,5 l/min under the operating conditions of Table 2 ($T_{d,h} = 82°C$, $p_{d,h} = 4.5$ bar) over a period of 30 min (ZIEGLER et al. 2017 [1])
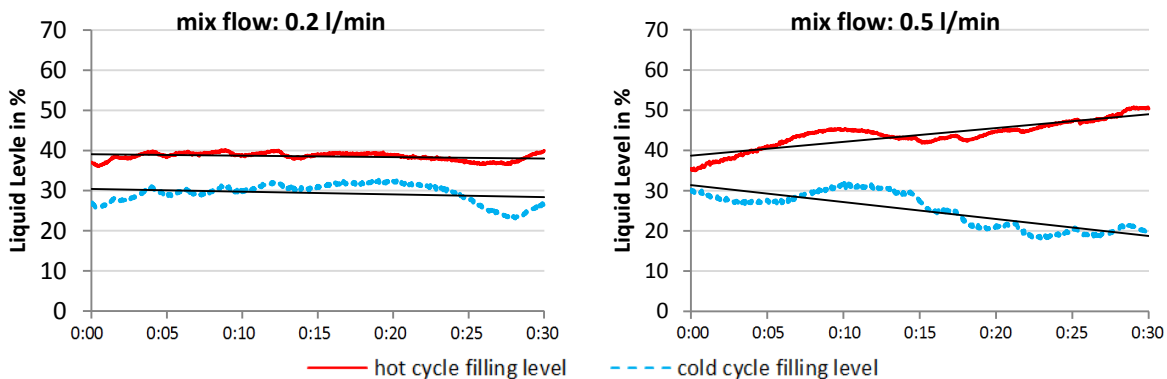

Figure 8: Sum of the liquid levels in the absorber and desorber vessels during continuous operation of the TUD-RRS with a mix flow of 0.2 l/min and 0.5 l/min under the operating conditions of Table 2 ($T_{d,h} = 82°C$, $p_{d,h} = 4.5$ bar) over a period of 30 min (ZIEGLER et al. 2017 [1])

6. CONCLUSIONS

The mix flow, which is required for the continuous operation of RRS, can be determined by using suitable assumptions by the thermodynamic correlations presented here. This places certain requirements on the structure of the RRS, which are necessary for the experimental verification of the calculation results. Based on earlier results with the RRS available at the TU Dresden it was found that the calculations are realistic. When comparing the calculated with the measured mix flow, according to the boundary conditions of Table 2, there is no significant difference for the given case. With both methods the value is determined at approx. 0.2 l/min. Thus the mathematical model has been verified for one operating point. In order to use the calculation model for the dimensioning of the mix flow of further resorption plants, an extensive operating map must be included. This is achieved by implementing a measuring device to determine the vapor volume flow, which is a direct indicator for the necessary mix flow, and then verified by measuring at different boundary conditions (external temperatures and NH_3 concentrations).

NOMENCLATURE

T	temperature	°C
p	pressure	bar
m	mass	kg
ξ	NH$_3$ concentration	kg$_{NH3}$/kg
$\dot{V}$	volume flow	l/min
$\dot{m}$	mass flow	kg/s
L	liquid level	
TUD	Technische Universität Dresden	

Subscript

d	desorber		NH3	ammonia
a	absorber		H2O	water
h	hot		fill	filling
c	cold			
r	rich solution			
p	poor solution			
vapor, vap	vapor			
liquid, liq	liquid			
C	condensation			
0, E	evaporation			

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