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Heat transfer in falling film evaporators during the industrial process of apple juice concentrate production

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Abstract Falling film, shell-tube type evaporators are commonly used heat exchangers for the production of fruit juice concentrate. The main problem in the design of the exchanger is a reliable estimation of wall heat transfer coefficients for all effects in real operating conditions. Most literature sources for the overall heat transfer coefficients are based on laboratory measurements, where the tubes are usually short, no fouling exists and the flow rate is carefully adjusted. This paper shows the heat transfer estimated in real industrial operating conditions, compared to literature sources. Paper is based on the author's own experience in designing and launching several evaporators for juice concentrate production into operation. As a summary, the design heat transfer coefficients are provided with relation to sugar content in juice concentrate.

Keywords: Heat transfer; Falling film evaporator; Industrial

Nomenclature

A	–	area, m^2
b	–	sugar content in solution, °Bx
c_p	–	solution specific heat, $\mathrm{kJ\,kg^{-1}\,K^{-1}}$
c_{pw}	–	water specific heat, $\mathrm{kJ\,kg^{-1}\,K^{-1}}$
d_{in}	–	tube internal diameter, mm
d_{out}	–	tube external diameter, mm

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h	–	specific enthalpy, $\mathrm{kJ\,kg^{-1}}$
I	–	enthalpy, kJ
$\dot{I}$	–	enthalpy flux, kW
k	–	wall heat transfer coefficient, $\mathrm{kW\,m^2\,K^{-1}}$
l	–	length, m
$\dot{m}$	–	mass flow rate, $\mathrm{kg\,s^{-1}}$
Nu	–	Nusselt number
n_r	–	number of tubes
ns	–	number of effect
p	–	pressure, MPa
Pr	–	Prandtl number
q	–	unitary heat, $\mathrm{kJ\,kg^{-1}}$
$\dot{Q}$	–	heat flux, kW
r	–	evaporation enthalpy, $\mathrm{kJ\,kg^{-1}}$
Re	–	Reynolds number
t	–	temperature, °C
v	–	specific volume, $\mathrm{m^3/kg}$

Greek symbols

α	–	convection coefficient, $\mathrm{kW\,m^{-2}\,K^{-1}}$
Γ_v	–	vapour mass flow intensity (flowrate per tube circumference) $\mathrm{kgs^{-1}m^{-1}}$
δ	–	wall thickness, mm
λ	–	heat conductivity, $\mathrm{kW\,m^{-1}\,K^{-1}}$
μ	–	dynamic viscosity, Pa s
ρ	–	density, $\mathrm{kg\,m^{-3}}$
ν	–	kinematic viscosity, $\mathrm{m^2\,s^{-1}}$

Subscripts

BPE	–	boiling point elevation
i	–	effect number
in	–	inside
c	–	concentrate
$cond$	–	condensate
s	–	steam
v	–	vapour

1 Introduction

The falling film evaporators, despite having been used in the industry for at least 50 years, are still subject to numerous papers concerning heat and mass transfer. There are a number of papers presenting formulas for heat transfer based on laboratory investigations in ideal conditions, where every parameter is strictly controlled and measured [1,2]. These formulas shall be however compared to industrial experimental results to validate results for the design purpose.

The real values for heat transfer coefficients for all evaporator effects have been shown in [3], as well as used for optimisation [4,5]. The values for heat transfer are also needed and used for computational fluid dynamics simulation [6]. Other numerical models for evaporation in cases of falling film may also be found in [7,8].

2 Theoretical estimation of heat transfer coefficients

To estimate heat transfer coefficients, the dependence of the juice properties on temperature and sugar content have to be estimated. For apple juice density the formula was applied [9,6]:

$$\rho = (1005.3 - 0.22556\, t) - 2.4304\, \frac{t}{1000} + 3.7329\, b + 0.01781937\, b^2 \ . \tag{1}$$

For juice dynamic viscosity [10,6]:

$$\mu = 4.3 \times 10^{-4} exp \left(3.357 \frac{b - 0.3155\,(t-50)}{116.8 - [b - 0.3155\,(t-50)]} \right) \ . \tag{2}$$

Juice thermal conductivity [9,6]:

$$\lambda = \left[0.574 + 1.699 \times 10^{-3}\, t - 3.608 \times 10^{-6}\, t^2 - 3.528 \times 10^{-3}\, b \right] \ . \tag{3}$$

Juice specific heat formula developed by the author on the basis of [11]:

$$c_p = 0.975\, c_{pw} \left[1.007 - 0.3826\, b - 0.1587\, b^2 \right] \ . \tag{4}$$

The convective heat transfer coefficient estimation is based on the criterial dimensionless numbers defined as [12]:

the Prandtl number:

$$\mathrm{Pr} = \mu \, \frac{c_p}{\lambda} \ , \tag{5}$$

the Reynolds number:

$$\mathrm{Re} = \frac{\dot{m}}{n_r \mu d_i \pi} \ . \tag{6}$$

According to [12] the Nusselt number in this case (without nucleation) is a combination of the laminar and turbulent contributions assuming that $\mathrm{Pr} < 50$. The Nusselt number for laminar flow yields

$$\mathrm{Nu}_{lam} = 0.9\mathrm{Re}^{+0.33} \ . \tag{7}$$

The Nusselt number for turbulent flow is

$$\mathrm{Nu}_{turb} = 0.00622\mathrm{Re}^{0.4}\,\mathrm{Pr}^{0.65} \ . \tag{8}$$

The combined Nusselt number is developed by geometrical summation

$$\mathrm{Nu}_{in} = \left(\mathrm{Nu}^2_{com} + \mathrm{Nu}^2_{turb}\right)^{0.5} \ . \tag{9}$$

The convective heat transfer coefficient:

$$\alpha = \lambda \left(\frac{9.81}{2}\right)^{0.33} \mathrm{Nu} \ . \tag{10}$$

For the convective heat transfer during evaporation inside the tubes of the falling film evaporators [13], the elaborated formula for the Nusselt number is

$$\mathrm{Nu} = 1.663\,\mathrm{Re}^{-0.2648}\,\mathrm{Pr}^{0.1592} \ . \tag{11}$$

Formula (11) is valid within the range: $15 < \mathrm{Re} < 3000$ and $2.5 < \mathrm{Pr} < 200$.

In an earlier paper [14] the Nusselt number was estimated as

$$\mathrm{Nu} = 3.8 \times 10^{-3}\,\mathrm{Re}^{0.4}\,\mathrm{Pr}^{0.65} \ . \tag{12}$$

These formulas were compared with the results of the estimation shown above on the basis of the VDI Heat Atlas. The results are shown in Fig. 1. The calculations presented in Figs. 1 and 2 are for five cases of apple juice falling film evaporator 20 t/h with dimensions shown in Tab. 2. Juice entering each effect under following thermodynamic conditions: case I – 95.2 °C, b = 10 °Bx, case II – 89.4 °C, b = 12.8 °Bx, case III – 83.4 °C, b = 17 °Bx, case IV – 74.4 °C, b = 22.9 °Bx, case V – 56.9 °C, b = 34.9 °Bx, end result – 56.9 °C, b = 70 °Bx. The averaged values for juice thermal properties were used.

The differences between convection heat transfer values obtained using the formulas shown above for each effect of the evaporator are extremely high – this has been noticed also by the author in [13].

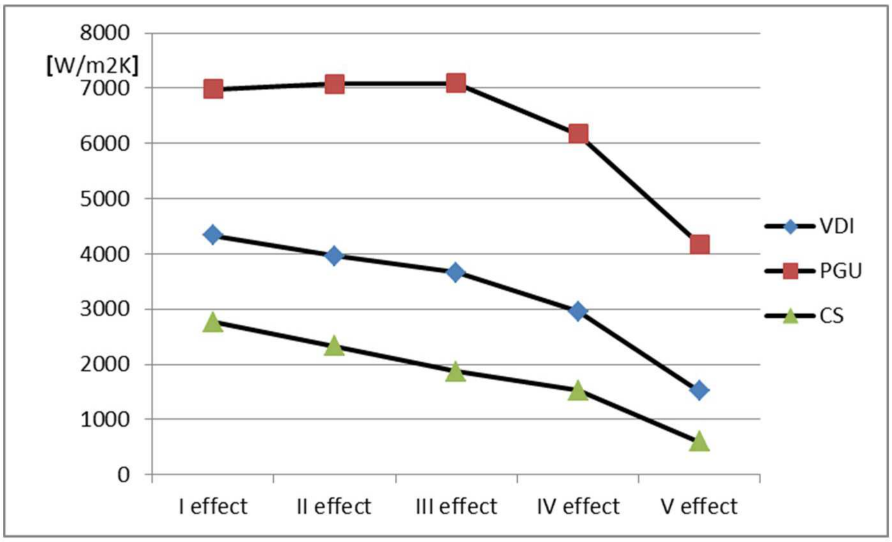

Figure 1: Convective heat transfer coefficients inside tubes calculated according to: the Verein Deutscher Ingenieure, (VDI) [12]; Prost, Gonzalez and Urbicain, (PGU) [13]; Chun and Seban, (CS) [14].

The steam flow heat convection coefficient on the outer side of the vertical exchanger tubes estimated using the [12] formulas:

$$\Gamma_v = \frac{\dot{m}_v}{\Pi\, d_{out}\, n_r} , \tag{13}$$

$$\mathrm{Re}_s = \frac{v}{\mu_v} , \tag{14}$$

$$\mathrm{Pr}_s = \frac{\mu_v\, c_{pv}}{\lambda_v} , \tag{15}$$

$$\mathrm{Nu}_s = 0.925 \left(\frac{\mathrm{Re}_s}{1 - \frac{\lambda_v}{\lambda_l}} \right)^{-0.3333} , \tag{16}$$

$$L_v = \left(\frac{\mu\, v^2}{9.81 \rho} \right)^{0.33} , \tag{17}$$

$$\alpha_v = \mathrm{Nu}_v \frac{\lambda_v}{L_v} . \tag{18}$$

The overall wall heat transfer coefficient formula reads:

$$k = \frac{1}{\frac{1}{\alpha_{in}} + \left(\frac{\delta\, d_{in}}{\lambda\, d_{out}}\right) + \frac{d_{in}}{d_{out}}\,\frac{1}{\alpha_{out}}}\ . \tag{19}$$

It is impossible to work out an effective design solution using only the recognised formulas given above for the evaporator design; therefore, the estimation of those coefficients on the basis of a real operating evaporator or 3D simulation [17] is a very important factor for evaporator designer.

3 Industrial scale studies

The investigations of operational parameters presented in this paper took place from 2004 to 2014 year. Four complete falling film evaporators with different capacities and designs were the subject of investigation. Investigations took place immediately after the installation of the evaporators to ensure that they were in a brand new condition.

Figure 2: One of the investigated evaporator Opole Lubelskie 20 t/h Manufacturer Zakład Remontowy Instalacji Przemysłowych Zenon Łagan.

One heat and mass transfer effect of the multieffect evaporator is comprised of a heat exchanger and a steam/liquid separator. The heat exchanger is heated by the steam from the previous effect and the juice concentrate is fed concurrently on the heated side of the exchanger. The steam condenses on the outer side of vertical tubes of the heat exchanger. The evaporation takes place inside the tubes where juice concentrate is supplied using a special distribution system assuring equal feed for all tubes and covering all of the tube inlet circles with a uniform layer of film.

The pressure was measured using analogue manometers with a 1% level of accuracy. The density and mass flow at the inlet and outlet of the evaporator were measured using accurate class 0.5 mass flow meters. The sugar concentration of juice entering the evaporator, after each effect and at the output of the final product were measured using the optical refractometer with a 0.1 °Bx level of accuracy. All test results were additionally balanced using the total energy balance for the evaporator during steady operation. The total amount of evaporated steam was checked against the measured evaporator capacity with a measuring accuracy of 1%.

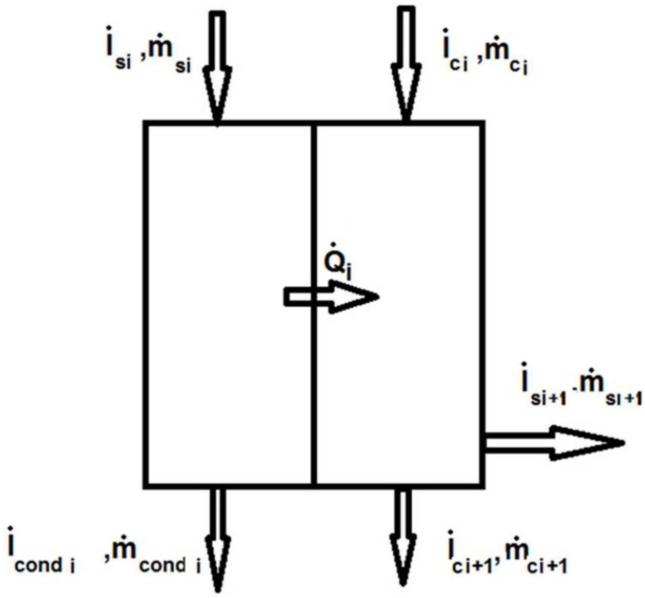

Figure 3: Schematic drawing of the ith case heat exchanger for energy balance.

On each effect, the energy balance for evaporation heat transfer can be formulated as follows (Fig. 3):

$$\dot{Q}_i = \dot{I}_{si} - \dot{I}_{condi} = \dot{I}_{ci+1} + \dot{I}_{si+1} - \dot{I}_{ci} = A_i\, k\, (t_{si} - t_{si+1}) \ . \tag{20}$$

The basis of the overall heat transfer calculations for the evaporator effect shown in Eq. (20) is the mass of evaporated H_2O m_{si+1}. This is accurately measured using optical refractometer after each evaporator effect. The overall heat transfer coefficient was calculated using the real temperature difference measured at each effect. The boiling point elevation (BPE) due to the sugar concentration was considered. The formula (20) with the consideration of t_{BPE} is transformed into

$$k = \frac{\dot{I}_{ci+1} + \dot{I}_{si+1} - \dot{I}_{ci}}{A_i\ [(t_{si} - t_{si+1}) - t_{BPE}]} \cong \frac{\dot{m}_{si+1}\, r_i}{A_i\ [(t_{si} - t_{si+1}) - t_{BPE}]}\ . \tag{21}$$

The results from a number of tests have been averaged for all investigated evaporators. In each case, a steady state was reached. The results shown in Tab. 1 are averages where the deviation from the averaged value is ±18%. In Tab. 1, comparisons between differently estimated values for the overall wall heat transfer coefficients for each effect of the evaporator is shown.

Table 1: The overall heat transfer coefficients according to different sources and industrial experimentation.

Evaporator effect	Overall wall heat transfer coefficient k [W m^{-2} K^{-1}]				
	Verein Deutscher Ingenieure [12] (calculated)	cheresources.com [16] (calculated)	Prost, Gonzalez & Urbicain [13] (calculated)	Chun & Seban, [14]	Industrial experiment ±18%
I	1915	2056	2532	1474	1926
II	1860	1700	2582	1351	1566
III	1689	1471	2288	1188	1351
IV	1551	1364	2312	1030	1108
V	1025	704	2001	503	717

Nevertheless, it is not possible to ensure the accuracy of formula by more than ±25%. Therefore, for the evaporator designer, simple equations may be worked out on the basis of industrial tests.

The averaged results for investigated evaporators are presented in Fig. 5. Each point represents average for several measurements in one case of the tested evaporator. The inlet conditions were within the range from 90 ^{o}C up to 100 °C and sugar content from 8.5 °Bx up to 12.5 °Bx. Final product after 5th effect with the temperature range from 55 °C up to 60 °C and 70 °Bx. The results are related to the averaged sugar content in one evaporator effect.

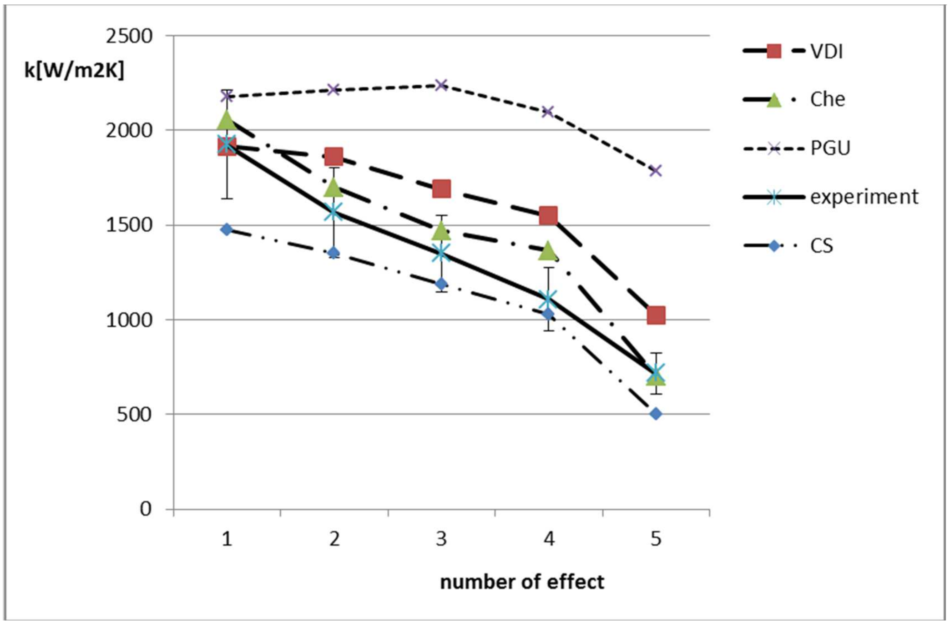

Figure 4: Comparison of the overall wall heat transfer coefficients given in Tab. 5 with the industrial tests VDI – [12], Che – [16], PGU – [13], CS – [14].

The functional approximation of the averaged results returns a simple equation

$$k = 9491.5\, b^{-0.652} \; . \tag{22}$$

The regression coefficient is $R^2 = 0.9984$. This simple relation allows for estimation of tube numbers for normalised sieve plates and shell inner diameters for each effect, therefore estimate the main costs of the evaporator.

4 Conclusions

Recognised formulas for overall wall heat transfer coefficients for each evaporator effect give rather divergent results. For the evaporator designer, it is really difficult to guess which relationship is more trustworthy.

The analysis of the formulas, based on the author's experiments, shows that the 'VDI' and 'cheresources' values are close to the field experiments. They can be used in evaporator design process.

Since the very complex relationships for the overall wall heat transfer coefficient gives results within ±25%, a simple formula (22) may be used for

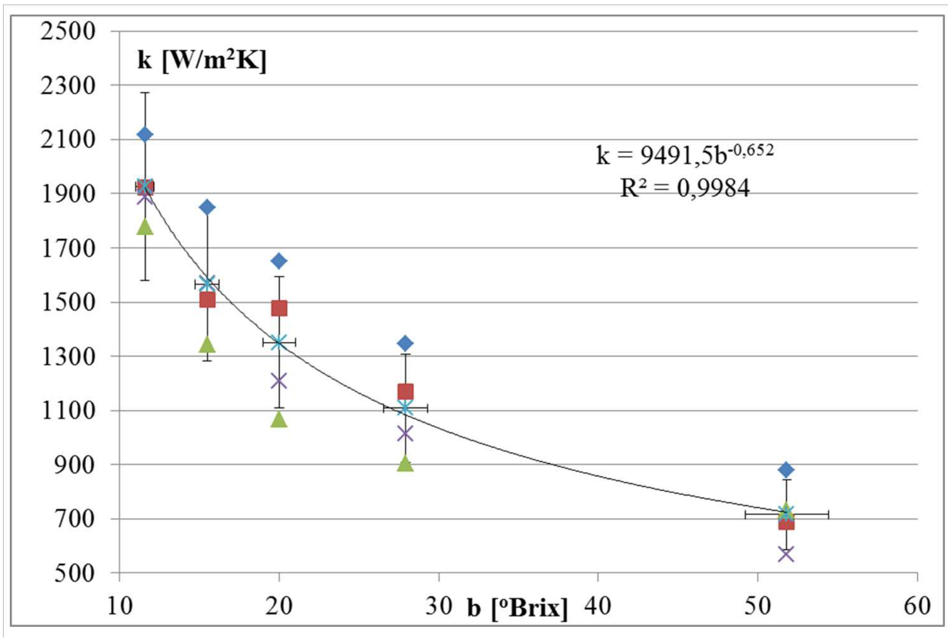

Figure 5: Averaged industrial results for overall wall heat transfer coefficient, different shapes shows different evaporator. For legend see previous page.

the design purpose without sacrificing the calculation accuracy. According to the author's long time experiments none of the laboratory formulas gave more accurate results for the case of investigated evaporators.

There are several reasons of discrepancies between the laboratory data and industrial measurements. There are also significant differences between the laboratory formulas obtained by different authors. There are several reasons for this issue, namely different used pipe size, length, sugar-water mixtures instead of fresh juice etc. Also the industrial conditions are usually more disturbed by different factors than the laboratory tests.

There were two main objectives of this investigation. First objective to find out which laboratory formula is the most accurate and suitable for an evaporator designer. Second objective was to estimate a relatively simple formula for estimation of the evaporator material costs prior to the exact design. Those results were above presented.

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