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Determination and correlation of heat transfer coefficients in a falling film evaporator

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Abstract

The aim of the work was to determine the heat transfer parameters of a single effect evaporator under different operating conditions, in order to extrapolate them to a multiple effect unit. The falling film evaporator consisted of 12 stainless steel vertical tubes, 1″ OD and 3 m long, having an evaporation capacity of 240 kg/h. In this unit the conditions of each effect of a multiple effect evaporator were simulated, varying the feed concentration and the pressure, setting in this way the saturation temperature and the transfer regime. Obtained values were correlated by means of an equation that links the heat transfer coefficient with the fluid properties, geometric parameters and flow conditions. Comparison with existing correlations was carried out.
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1. Introduction

The concentration of a fruit juice is a widely used practice in the fruit juice manufacturing industry, and it has two main purposes: (1) to reduce the volume and weight of the product, with the subsequent lowering of storage, packaging and distribution costs, and (2) to increase the stability of the juice by reducing its water activity, which is a predominant factor in the majority of the mechanisms of deterioration.

Although other methods of concentration such as freezing concentration and reverse osmosis are used nowadays, evaporation is still the most popular due to operational and economic reasons.

Evaporation is a unit operation that eliminates water from a liquid food. If the liquid contains dissolved solids, the concentrated solution can become saturated or oversaturated, with solid crystals deposition.

As fruit juices contain many substances that can be damaged if submitted to high temperatures during relatively long periods, evaporation under vacuum seems to be the logical answer to this problem. When vacuum evaporation is carried out, the boiling point is lowered, and so thermal degradation is minimized.

A falling film evaporator is essentially a shell and tube heat exchanger. Steam condensing on the shell side provides the latent heat that allows the evaporation of a mass of water from the solution flowing in the tube side. Water vapor and concentrated juice, in thermodynamic equilibrium, are then separated. This process can be accomplished in one evaporation body, so the boiling concentrated solution is withdrawn from the unit for further processing and the vapor is condensed in a separated condenser. Such equipment is named "single effect evaporator".

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Nomenclature

A	heat transfer area, m^2		x	fractional weight of sucrose, kg sucrose/kg solution
C	condensed steam mass flow rate, kg/s			
F	cold stream mass flow rate, kg/s		ΔT	temperature difference between hot and cold fluid, °C
g	acceleration of gravity, m/s^2			
$\hat{h}$	enthalpy per unit mass, J/kg		ρ	density, kg/m^3
h	film coefficient, W/(m^2 °C)		μ	viscosity, Pa s
h^+	dimensionless heat transfer coefficient, defined by Eq. (8)		*Subscripts*	
k	thermal conductivity, W/(m °C)		C	condensate
L	liquid mass flow rate, kg/s		F	feed
Pr	Prandtl number, dimensionless		i	inner surface
Re	Reynolds number, dimensionless		L	liquid
Q	heat exchanged, W		m	logarithmic mean
r	evaporator tube radius, m		o	outer surface
R_F	fouling resistance, m^2 °C/W		S	steam
S	steam mass flow rate, kg/s		V	vapor
U	overall heat transfer coefficient, W/(m^2 °C)		w	wall
V	vapor mass flow rate, kg/s			

However, if a high degree of concentration is needed, it is advisable to use more than one smaller unit in series instead of a large one. Under these conditions, the vapor and the solution leaving the first unit are, respectively, the heating medium and the process stream for the second one. For a suitable driving force to exist in the second unit, the solution boiling point has to be reduced, and this is accomplished by reducing the pressure in the evaporation chamber. In this way, a train of evaporators of decreasing pressure in the direction of the heating vapor is obtained.

This type of arrangement is defined as "multiple effect evaporator", and equipments of 3, 4 and 5 effects in series are common in the food industry.

The equations governing the processes occurring at each evaporator are the well known mass and energy balances, and heat transfer rate equation from the hot to the cold stream. Referring to Fig. 1, they can be stated as follows:

Mass balances, steady state:

Cold stream (process, or juice side), overall:

$$F = V + L \tag{1}$$

Cold stream (process, or juice side), dissolved solids:

$$x_F F = x_V V + x_L L \tag{2}$$

Thermodynamic equilibrium relationships state that $x_V \rightarrow 0$.

Hot stream (condensing steam): $S = C$ (3)

Energy balances, steady state:

Cold stream (process, or juice side):

$$F\hat{h}_F + Q = V\hat{h}_V + L\hat{h}_L \tag{4}$$

Hot stream (condensing steam): $S\hat{h}_S = Q + C\hat{h}_C$ (5)

Transfer equation:

$$Q = UA\Delta T \tag{6}$$

The inverse of the overall heat transfer coefficient can be written as the addition of all the resistances to heat transfer posed by both fluids boundary layers (convective), and the tube wall and fouling (conductive):

$$\frac{1}{U} = \left(\frac{1}{h_i} + R_{F,i}\right)\frac{A}{A_i} + \frac{\Delta r}{k_w}\frac{A}{A_m} + R_{F,o} + \frac{1}{h_o} \tag{7}$$

Since resistance due to the wall and fouling is considerably lower than that imposed by the liquid films, and of these, the film heat transfer coefficient of the condensing steam is much higher than the other one, the controlling resistance in this system is that imposed by the inner liquid film. Hence, the design of a unit is strongly dependent on the predicted value of this variable, as the calculated area is directly proportional to the overall resistance, and this is controlled by the inner heat transfer coefficient.

The aim of this work was to obtain experimental values of the film coefficient for the evaporating stream under different conditions, and to fit these values to an equation. It is expected that, if a good correlation is
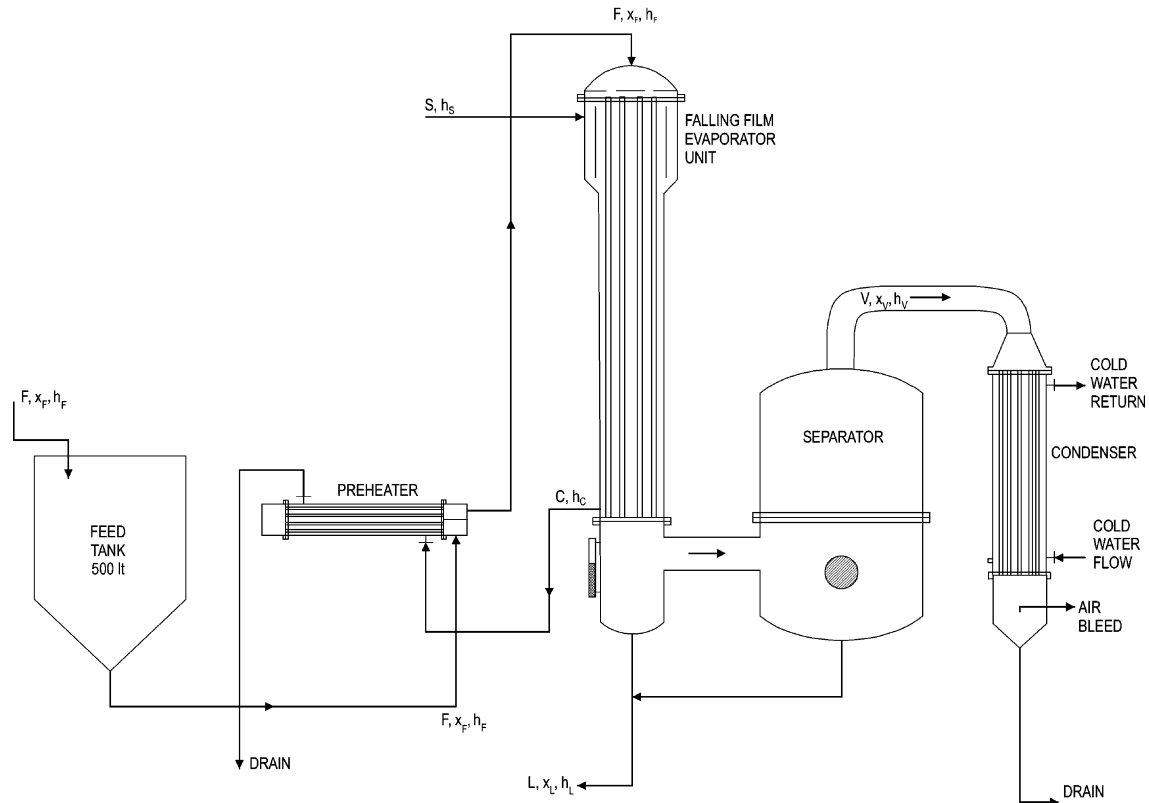
Fig. 1. Schematic representation of an evaporator.

obtained, then rating of existing units and design of new ones could be carried out with improved accuracy.

2. Experimental equipment

The experimental evaporator was of the falling film type. In these units, the liquid is fed to the upper end and flows downwards as a film on the inner surface of the tubes. As evaporation proceeds, vapor in equilibrium with the boiling solution is dragged downwards and both leave the unit by the bottom, where the phases separation takes place.

This type of evaporator is particularly useful for fruit juice concentration, as it allows a short contact time with the heating surface, reducing the chances of thermal damage. It is of easy cleaning and startup, and it ensures a minimum loss of product.

The evaporator used was an ALVAL made. It consists of 12 1″ OD vertical tubes of 3 m length, having an evaporation capacity of 240 kg/h. It is mounted on a skid which contains all the ancillary equipment (separator, preheater, condenser, pumps) and instrumentation.

Since the available equipment is a single effect evaporator, the initial operating conditions were set for the typical conditions found in the first effect of a multiple

effect evaporator. After a number of runs, the conditions were changed to meet those prevailing at the second effect, and then changed once again to reproduce the conditions at the third effect. For doing so, the inlet concentration, flow rate and pressure were varied. The values of these parameters were previously determined by means of a simulation package developed at PLAP-IQUI (Ugrin & Urbicain, 1999; González, Ugrin, & Urbicain, 2001).

The model fluid studied was a solution of sucrose in water, as it can be considered similar to a fruit juice from the point of view of its thermo-hydraulic behavior. Juice higher susceptibility to bacterial and fungal contamination and also costs involved, were other considerations taken into account to make the solution model the best choice.

The operating conditions are listed in Table 1.

3. Experimental results

The measured variables have been: flow rate and pressure of saturated steam (heating stream), pressure, inlet temperature, inlet concentration and outlet temperature of sucrose solution (process stream).

The heat and mass balances Eqs. (1)–(6) led to the calculation of the overall heat transfer coefficient and,

Table 1
Experimental conditions for a three effect evaporator

	Cold stream				Hot stream	
	Temperature [°C]	Pressure [bar abs]	Concentration [kg sugar %]	Flow rate [m³/h]	Pressure [bar abs]	Temperature [°C]
First effect	17–28	0.55	9.3–12.8	0.240–0.340	2.00	120.2
Second effect	15–26	0.30	17.3–18.8	0.205–0.320	1.42	109.7
Third effect	11–24	0.10	26.7–36.2	0.165–0.220	1.00	99.6

Table 2
Calculated values of heat transfer parameters

Run	w [m³/h]	Q [W]	U [W/m² °C]	h_o [W/m² °C]	Re_L	Pr_L	h_i [W/m² °C]	h^+
First effect								
1	0.310	64004.3	2080.4	6643.3	2399.2	3.54	4357.8	0.2334
2	0.325	65679.5	2186.9	6648.5	2612.7	3.45	4850.2	0.2549
3	0.300	65164.0	2169.8	6646.9	2204.7	3.58	4767.4	0.2585
4	0.280	63312.7	2108.1	6641.4	2048.5	3.59	4482.1	0.2434
5	0.260	65240.8	2172.3	6647.1	1744.8	3.68	4779.7	0.2647
6	0.300	71034.8	2365.3	6670.0	2098.3	3.69	5807.6	0.3224
7	0.340	67634.6	2252.0	6655.6	2702.0	3.51	5178.1	0.2759
8	0.310	67213.7	2238.0	6654.0	2168.3	3.85	5105.6	0.2939
9	0.290	65983.9	2144.8	6649.6	1894.2	3.98	4647.0	0.2739
10	0.240	65986.2	2144.8	6649.6	1247.1	4.41	4647.3	0.2979
11	0.320	69282.4	2306.9	6662.3	2238.2	3.86	5473.2	0.3154
12	0.300	68887.2	2293.8	6660.6	2029.7	3.88	5400.8	0.3128
Second effect								
13	0.250	66522.2	2162.3	6169.2	842.1	6.76	5007.3	0.4396
14	0.280	65525.0	2181.8	6163.9	1129.0	6.11	5116.9	0.4159
15	0.300	66164.6	2203.1	6167.3	1378.5	5.69	5233.2	0.4017
16	0.205	66769.1	2223.2	6170.5	411.4	9.15	5345.8	0.5991
17	0.230	66447.9	2159.9	6168.8	654.8	7.27	4994.6	0.4643
18	0.320	66270.4	2206.6	6167.8	1393.2	6.02	5252.7	0.4187
19	0.270	64098.9	2134.3	6156.7	1024.0	6.32	4867.7	0.4064
20	0.220	62111.4	2068.1	6147.3	638.4	7.31	4541.7	0.4250
21	0.290	65035.6	2165.5	6161.4	1163.8	6.11	5029.8	0.4089
22	0.260	63834.4	2125.5	6155.4	971.2	6.32	4822.9	0.4027
23	0.205	63887.7	2127.3	6155.6	494.9	7.99	4831.9	0.4849
24	0.230	70788.7	2357.1	6193.6	680.6	7.19	6168.0	0.5702
Third effect								
25	0.200	57418.8	1633.1	5626.6	140.2	29.4	2994.5	0.7801
26	0.200	54407.8	1547.4	5608.7	115.3	36.8	2722.8	0.8379
27	0.185	44629.0	1243.4	5570.2	136.3	31.5	1908.2	0.5222
28	0.190	47402.3	1376.9	5577.0	69.8	59.6	2240.6	0.9831
29	0.165	47014.2	1337.1	5575.8	15.6	199.5	2137.5	2.2350
30	0.190	47982.5	1393.7	5578.9	50.3	80.4	2285.3	1.2464
31	0.220	50824.8	1476.3	5590.3	155.3	34.0	2513.7	0.7298
32	0.210	47524.3	1380.4	5577.4	105.0	47.0	2250.0	0.8299
33	0.185	47197.3	1370.9	5576.4	49.8	78.5	2225.0	1.1931
34	0.180	45117.5	1310.5	5571.1	22.6	162.7	2070.8	1.8753
35	0.195	48473.1	1408.0	5580.6	48.5	86.3	2323.5	1.3344
36	0.205	47229.1	1371.8	5576.5	67.2	69.3	2227.4	1.0911

with known values of the resistances offered by the wall and fouling, and calculated for the steam, the values of the film coefficient for the process stream were obtained by means of Eq. (7).

The film coefficient was multiplied by an adequate parameter in order to make it dimensionless, h^+. It is a common practice in the literature to show results in this form, and it is convenient for comparison purposes with

other works. The dimensionless heat transfer coefficient was obtained by means of the following equation:

$$h^+ = h_i \left(\frac{\mu_L^2}{\rho_L^2 k_L^3 g} \right)^{1/3} \tag{8}$$

The experimental results and calculated values for some runs are shown in Table 2.

4. Correlation of experimental values

The values of the dimensionless heat transfer coefficient for evaporation were fitted to an equation of the type:

$$h^+ = aRe_L^b Pr_L^c \tag{9}$$

The regression was made by two different ways: the first one was to fit the data to a pair of correlations, each one covering a different range of flow regimes: laminar and transition; the second fitted the data to only one correlation for all values of Reynolds number available. Comparing both correlation parameters R^2, it was concluded that the best fit was achieved by means of a unique correlation, and it is valid for Reynolds from 15 to 3000, and Prandtl from 2.5 to 200.

The resulting equation was:

$$h^+ = 1.6636Re_L^{-0.2648} Pr_L^{0.1592} \quad 15 < Re_L < 3000;$$
$$2.5 < Pr_L < 200 \quad \text{with } R^2 = 0.988 \tag{10}$$

The experimental data and the obtained correlation are shown in Fig. 2.

This equation can be further simplified by the definition of a functional relationship between the Reynolds and Prandtl numbers. Strictly speaking, these numbers are not related each other, as they are independent dimensionless numbers that represent relationships between geometric and physical parameters of the heat transfer system. However, for a system like this, any change of any of them provokes a unique change on the other one. In fact, for a given temperature, the viscosity of a liquid defines its flowing behavior and consequently its residence time, and so the final concentration of the solution. On the other hand, given the concentration and temperature, the physical properties that define the Prandtl number are fixed. So, the conditions that determine a Reynolds number also determine only one value of the Prandtl number.
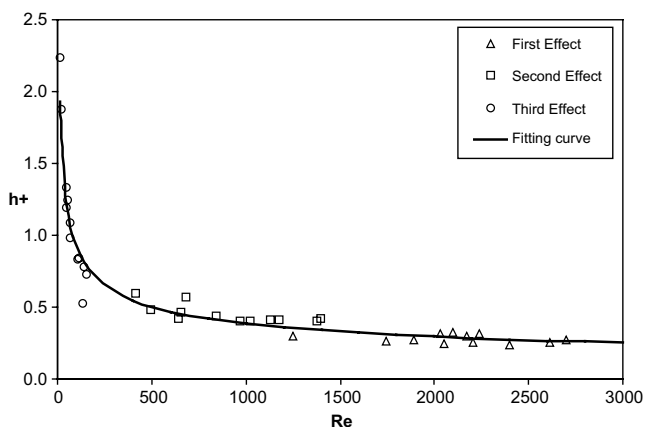
It was found that the relationship between both for the studied system was:

$$Pr_L = 1878Re_L^{-0.8204} \quad 15 < Re_L < 3000;$$
$$2.5 < Pr_L < 200 \quad \text{with } R^2 = 0.9862 \tag{11}$$

Making the substitution of Eq. (11) into Eq. (10) a simpler correlation of h^+ as a function of Re_L only is obtained:

$$h^+ = 5.5236Re_L^{-0.3854} \tag{12}$$

Naturally, the relationship $N_{Pr,L}$ versus $N_{Re,L}$ that leads to Eq. (12) is strictly valid only for the studied system. For other solutions or geometries, the generally valid correlation to be taken into account is Eq. (10).

5. Comparison with other correlations

In Table 3 different correlations for the calculation of h^+ are shown, and in Fig. 3 a graphical comparison is made of the values obtained by means of the present correlation, and those calculated by other authors, for an arbitrarily chosen Prandtl number = 4.

It can be seen that the results of this work agree with those of Mc Adams, Drew, and Bays (1940), Nusselt (1916), and Chung and Seban (1971) in the laminar and transition regions, in the fact that h^+ decreases with decreasing Re_L. However, although the shape of the curves is similar, it is evident that the correlation obtained in this work leads to values considerably higher than those calculated by means of those correlations.

Although the turbulent zone has not been covered by the present work, it may be expected that in that zone the tendency reverts, as shown by correlations of Mc Adams et al. (1940), Garwin and Kelly (1955), Wilke (1962), Ahmed and Kaparthi (1963), Herbert and Stern (1968), and Chung and Seban (1971).

The correlations of Ahmed and Kaparthi (1963), Wilke (1962) and Mc Adams et al. (1940) seem to maintain the same slope in both the laminar and transition zone, which is in contradiction with the results of this work.

On the other hand, a recent work from Uche, Artal, and Serra (2002) in which a review is made of numerous data of different authors, with Re_L ranging from 10 to 20 000, shows clearly that the coefficient decreases between the lower limit and $Re_L = 2500$ where it becomes a minimum, and then it increases with increasing Re_L (see Fig. 3).

The results are eloquent and show ratios that range from 3 to 11 times for the preceding works, and from 1 to 4.5 comparing them with the present work. The problem the engineer faces when he/she has to select a

Fig. 2. Experimental values and fitting curve.

Table 3
Correlations reported in the literature for the calculation of dimensionless heat transfer coefficient

| No	Equation	Comments	Author
1.	$h^+ = 0.01\,(Re_L Pr_L)^{1/3}$	Turbulent flow of falling water inside copper tubes, for Re_L between 1600 and 50000, and a mean water film temperature of 88 °C	Mc Adams et al. (1940)
2.	$h^+ = 0.02007 Re_L^{1/3}(sen\,\theta)^{0.2}$	Turbulent flow of a liquid falling over a flat plate 1 m long with varying slope and Re_L between 2900 and 12800, and a mean liquid film temperature of 34 °C	Garwin and Kelly (1955)
3.	$h^+ = 8.7 \times 10^{-3} Re_L^{0.4} Pr_L^{0.344}$	Turbulent flow for heating of a falling film of water, and water–ethylenglycol mixtures, over the outside of a metallic rod 2.4 m long and diameter of 4.2 cm internally heated by means of hot water	Wilke (1962)
4.	$h^+ = 6.92 \times 10^{-3} Re_L^{0.345} Pr_L^{0.4}$	Flow of falling water and aqueous solutions of glycerol in the inner surface of a copper tube of 3.015 cm I.D. with Re between 3 and 10250 and Pr between 3.6 and 950	Ahmed and Kaparthi (1963)
5.	$h^+ = 8.54 \times 10^{-4} Re_L^{0.65}$	Water flowing inside a copper tube of 2.408 cm ID, and Re_L ranking from 3000 to 20000, with a mean film temperature of 71 °C	Herbert and Stern (1968)
6.	$h^+ = 3.8 \times 10^{-3} Re_L^{0.4} Pr_L^{0.65}$	Water evaporating as a falling film over an electrically heated vertical tube. Re_L between 320 and 21000, and saturation temperature between 28 and 100 °C	Chung and Seban (1971)
7.	$h^+ = 0.89\delta^{1/3}/\{5 + [\tan^{-1}(2.73\sqrt{Pr_L}) - \tan^{-1}(0.455\sqrt{Pr_L})]/0.091\sqrt{Pr_L} + \ln(\delta/30)/(0.36 Pr_L)\}$ with $\ln\delta = 0.786 + 0.103\ln Re_L + 0.041(\ln Re_L)^2$	Theoretical expression for predicting heat transfer coefficients, based on some hydrodynamic considerations on falling liquid films for Pr_L greater than 1 and film thickness $\delta > 30$	Narayana Murthy and Sarma (1977)
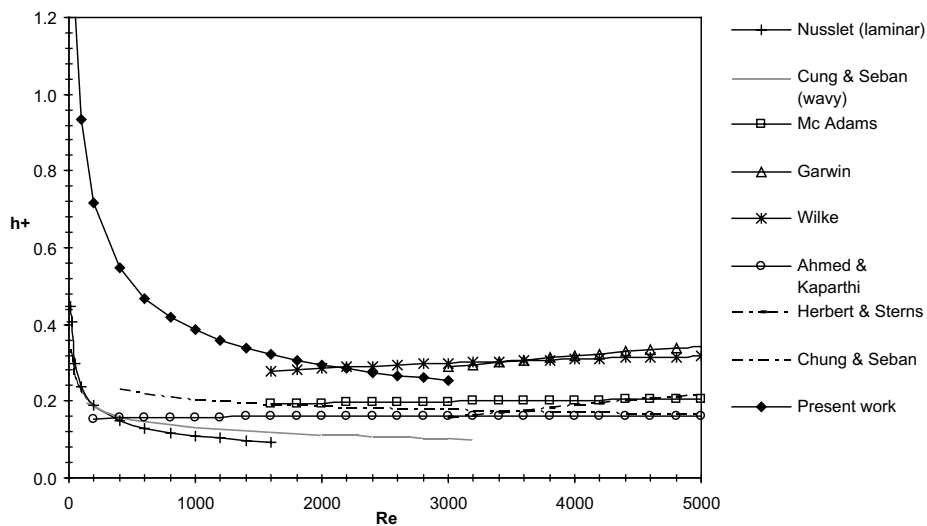
Fig. 3. Graphical comparison of present work with other authors'.

proper equation for the evaporating heat transfer coefficient is apparent.

6. Conclusions

The purpose of this work has been the collection of experimental data in a single effect vertical falling film evaporator in order to obtain a general correlation for the calculation of the liquid side heat transfer coefficient, suitable for the conditions found in any effect of a multiple effect unit as those usually found in the fruit concentrate juice industry.

A correlation of h^+ as a function of the film Reynolds and Prandtl numbers was found. A correlation in terms of Reynolds number only was also obtained for this particular system.

Although the equipment characteristics made it possible the experimentation in laminar and transition zones only, the results apply since those are the conditions

found at a juice production plant. This is particularly true from the second effect on, in which the increasing concentration and decreasing temperature make the viscosity to increase, lowering the prevailing Reynolds number.

Acknowledgements

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