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SHELL-SIDE HEAT TRANSFER IN BAFFLED CYLINDRICAL SHELL- AND TUBE EXCHANGERS—AN ELECTROCHEMICAL MASS-TRANSFER MODELLING TECHNIQUE

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Abstract—This paper describes the application of an electrochemical mass-transfer modelling technique to the determination of local shell-side heat-transfer coefficients in a model of a baffled cylindrical shell-and-tube heat exchanger. The validity and accuracy of the electrochemical method are demonstrated by comparison with heat-transfer measurements and with mass-transfer data obtained using the mercury evaporation technique. An error analysis is also presented.

In order to summarise the data of this (and subsequent) work, a number of correlations have been tested and equations of the form

$$j_M = ARe^{-1.0} + BRe^{-0.5} + C$$

are shown to give the best fit.

NOMENCLATURE

A, constant in j-factor correlation;
A_M, minimum area for flow at centreline of baffle compartment [m^2];
B, constant in j-factor correlation;
C, constant in j-factor correlation;
C_P, fluid heat capacity at mean temperature [$J kg^{-1} K^{-1}$];
D_V, mass diffusivity [$m^2 s^{-1}$];
D, constant in j-factor correlation;
d, tube diameter [m];
G, mass flowrate [$kg s^{-1}$];
h, heat-transfer coefficient [$J m^{-2} s^{-1} K^{-1}$];
k, fluid thermal conductivity at mean temperature [$J m^{-1} s^{-1} K^{-1}$];
k_c, mass-transfer coefficient [$m s^{-1}$];
u, fluid velocity, $= \dfrac{G}{\rho A_M}$ [$m s^{-1}$].

Greek symbols

μ, viscosity at mean fluid temperature [$kg m^{-1} s^{-1}$];
ρ, fluid density at mean temperature [$kg m^{-3}$].

Dimensionless numbers

j_H, j-factor for heat transfer $\left(= \dfrac{h}{\rho C_P u} Pr^{2/3}\right)$;
j_M, j-factor for mass transfer $\left(= \dfrac{k_c}{u} Sc^{2/3}\right)$;
Pr, Prandtl number $\left(= \dfrac{C_P \mu}{k}\right)$;
Re, Reynolds number $\left(= \dfrac{du\rho}{\mu} = \dfrac{dG}{A_M \mu}\right)$;
Sc, Schmidt number $\left(= \dfrac{\mu}{\rho D_V}\right)$.

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INTRODUCTION

THE DESIGN of shell and tube heat exchangers requires a knowledge of the average shell-side heat-transfer coefficient. This parameter is difficult to predict and, in particular, the effect of variations in shell-side geometry is not well understood. In this and subsequent papers, we seek to elucidate the relationship between the geometrical factors and the shell-side heat-transfer coefficient. This paper deals largely with the validation of a mass-transfer analogue technique for the determination of local heat-transfer coefficients. Various forms of correlation have also been tested for use in summarising the data of this and subsequent work, and for further analysis of the effects of variation in shell-side geometry on the transfer coefficient.

Early design techniques used an effective mass velocity based on various geometrical factors to predict the average shell-side coefficient [1–5]. More recently, variations of the stream analysis approach due to Tinker [4, 5] have been presented by Parker and Mok [6], and Devore [7]. Palen and Taborek [8] have reported a more rigorous computer-based approach and Grant [9] has also developed a prediction method, based on a stream analysis, in the form of a computer program.

These prediction techniques require a more detailed knowledge of the shell-side heat-transfer characteristics than has been previously available. Such information is provided by measurements of local transfer coefficients obtained for short lengths of individual tubes in the bundle, both by direct means (electrical heating) and by use of a mass-transfer analogy. The effects of the various design parameters can thus be studied and this approach gives some insight into the flow patterns on the shell-side and allows the calculation of bundle and zonal average transfer coefficients as required.

Among the various mass-transfer methods the electrochemical method is a well established technique, from which, by use of the Chilton–Colburn analogy

[10], the corresponding heat-transfer coefficients can be found. This paper compares data obtained by this method with those obtained both by direct heat-transfer experiments [11–15] and by a mercury evaporation technique [16, 17] in bundles of identical geometry.

PREVIOUS WORK

Flow in a baffled cylindrical tube bundle is complicated for two reasons, the first is the approximately sinusoidal overall flow pattern as the liquid flows through the bundle, and the second is the influence of the various leakage flows introduced by the necessity of providing manufacturing clearances between the tubes and baffles and between the baffles and the exchanger shell. Correspondingly a complex heat-transfer situation exists with a large number of geometrical parameters affecting both the flow and the heat transfer. Those relevant to exchangers both with and without leakage are shell diameter; baffle cut down, baffle spacing; and tube diameter, pitch and arrangement. For the leakage case the various leakage flows are determined by baffle thickness, and the size of the clearance between the individual tubes and the baffles, and between the baffles and the shell.

The major experimental study of the average shell-side heat-transfer coefficient was carried out by Bergelin and co-workers at the University of Delaware [11, 12], following earlier work on rectangular tube banks [18–21]. The entire Delaware investigation has been summarised in a report [13] and by Bell [14]. The work covered both shell-side heat transfer with normal clearances and with these clearances reduced or eliminated.

Local coefficients have been measured by replacing a section of a tube by an electrically-heated probe [22–25]. Because of experimental problems the exchangers did not represent industrial practice. Gay and co-workers [16, 17] used a mercury evaporation technique to model heat transfer from a short section of an individual tube. Their exchanger was substantially identical to a cylindrical segmentally baffled exchanger used in the work of Bergelin [11] and clearance flows were eliminated by gasketting. Their results agreed well with those of Bergelin. The distribution of individual tube coefficients [17] gave some insight into the flow patterns on the shell-side. While restricted to the non-leakage case and to $54 < Re < 1625$, and thus not representative of industrial practice, this work together with that of Bergelin *et al.* provides a basis against which the electrochemical modelling technique (a ferri-ferrocyanide redox couple) can be validated.

EXPERIMENTAL

The heat exchanger models used in this work are, save for some small details, identical to that used by Gay and Williams [16], which resembled closely that of Bergelin *et al.* [11]. Thus the individual tube coefficients measured in this work could be compared with those of Gay and Williams and both compared with the bundle average values of Bergelin *et al.* As in these previous studies, leakage was eliminated between the baffle and shell and between tube and baffle hole. Initially this was done with the aid of gaskets as for the earlier work but gaskets tended to tear during bundle assembly and so produce unwanted leakage flows. Consequently plastic baffles were used which were a tight fit on the tubes and in the shell. By-passing between bundle and shell was also reduced by placing the tubes as near the wall of the shell as possible and by careful location of the baffle spacer rods. Work on two model exchangers is reported here, the first was the model used at the University of Aston in the main validation work, and some further results for this purpose were obtained at AEE Winfrith, on a second model. Only model 1 is described here, model 2 in all essential points was identical to model 1. The tube bundle is shown in Fig. 1, and the tube layout with tube reference numbers in Fig. 2. Figure 1 also shows the various electrodes and probes used. The tube bundle had eighty 0.375 in (9.5 mm) O.D. perspex tubes about 18 in (455 mm) long arranged on a staggered square pattern with a 1.25 pitch to diameter ratio. Baffles of 31.0 and 43.7% diametrical cut were used, at 4 in ($\sim$102 mm) centres with their faces 3.89 in ($\sim$99 mm) apart. The ends of the test bundle are closed off by two dummy tube plates, whose faces are 16 in ($\sim$406 mm) apart, in line with the outer edges of the inlet and exit ports. These cut downs were among those used by Gay and Williams and by Bergelin *et al.* The model, then, could be expected to produce results comparable to those obtained before.

The complete exchanger model was mounted vertically, in order to prevent gas being trapped in the system. Design details can be found in [16, 26] and [27].

The fluid circulated through the model was an electrolyte of the composition and properties shown in Table 1. The mass transfer coefficient for a length of single tube equal to the baffle spacing was obtained by measuring the limiting current through a nickel rod cathode, as shown in Fig. 1. The anode necessary to complete the electrolysis cell was a set of nickel plates lining the upper, i.e. the exit, port of the exchanger model.

The general technique has been described elsewhere by numerous authors (see references in [27, 28]), and all the usual precautions were taken—among them were the exclusion of light and the use of de-oxygenated solutions. However, since the experimental situation in this work was more complex than any previously reported, preliminary proving experiments were carried out. These are not reported in detail here but included obtaining satisfactory plateau shaped curves over single cathodes and groups of cathodes, and an investigation into the effect of different locations of the anode. The alternatives examined included the anode as a shell-liner, as a lined exit port (as finally used) and as groups of tubes around the cathode being investigated. In all cases satisfactory plateaux giving the same coefficients for identical locations and flows were obtained. Full details are given by Mackley [27]. Before each run, the nickel cathode was activated by evolving hydrogen on its surface using the base electrolyte containing sodium

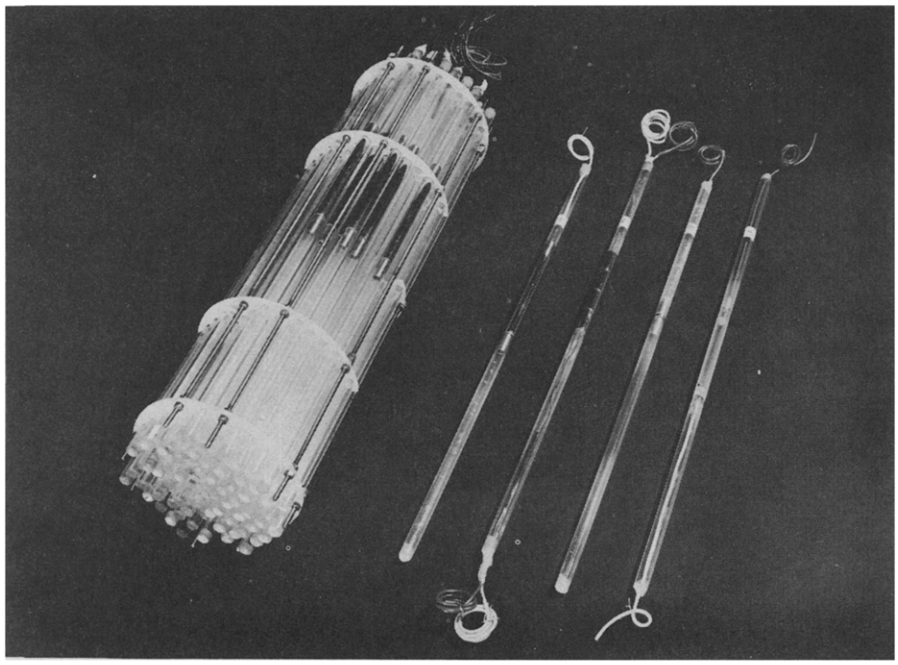

FIG. 1. Tube bundle with various probes: 1, baffle-spacing length electrode; 2, composite electrode; 3, micro-electrode; 4, pressure measuring tube.

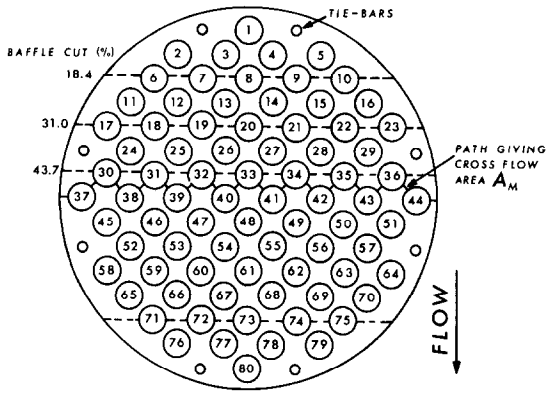

FIG. 2. Diagram showing the tube arrangement. The locations of the baffle edges for the various cutdowns are also shown.

hydroxide only. Temperature effects were corrected for both by considering the effect on the physical properties involved and also by a series of measurements of limiting current on a single cathode at constant flow but at different temperatures. There was close agreement ($\pm 1\%$ discrepancy at worst) and the experimental correction was used in practice.

In addition to the mass-transfer measurements, pressure measurements similar to those of Bergelin *et al.* [11] were made within the bundle. A single perspex pressure measuring tube was used having two 0.0625 in (1.59 mm) tappings. The separation of the tappings was

Table 1. Properties of the electrolyte at 25°C

Composition:	$K_3Fe(CN)_6$ 0.01 M
	$K_4Fe(CN)_6$ 0.01 M
	NaOH 1.0 M
Density, ρ:	$1043\,kg\,m^{-3}$
Viscosity, μ:	$1.11 \times 10^{-3}\,kg\,m^{-1}\,s^{-1}$
Diffusivity, D_v:	$6.62 \times 10^{-10}\,m^2\,s^{-1}$

such that when the tube was inserted in the bundle, the tappings were located midway between the baffles in two adjacent baffle compartments. The purpose of these measurements was to act as an additional check that the conditions of the work of Bergelin *et al.* were being reproduced rather than being a fundamental contribution to pressure drop work.

As in the work of Gay and Williams, it was found that only 44 of the 80 tube positions needed to be examined as symmetry existed across the central plane of the bundle. The baffle cut-downs used were 31.0 and 43.7% of the shell diameter. The tests were confined to non-leakage cases, except for a few with leakage between the shell and baffle. The Reynolds number, *Re*, range was 24–1500 for the 43.7% baffle cut-down and 300–1500 for the 31.0% cut-down. In every case, the baffle compartment investigated was the third (of four) from the exchanger inlet.

Additional work was carried out at AEE Winfrith on the model 2 exchanger at an 18.4% baffle cut and a baffle spacing of 1.91 in (48.5 mm) over the range

$300 < Re < 25000$, and this provided an additional comparison with the work of Bergelin *et al.* [19]. In this bundle there was a nickel rod electrode at each of the 80 tube positions and a data-logging system was used to record current measurements, etc.

It should be emphasised that the transfer coefficients reported in this work are not true point values but represent an average value for any particular tube over a length equal to the baffle spacing.

was for the non-leakage 43.7% baffle cut-down case, the one case studied by Bergelin *et al.* (the work of Tomkins [29], in [15] and Brown [11]), Gay and Williams [16], and this work (at Aston). The entire data available from all four workers is plotted in Fig. 3.

However, the following points must be noted. The data of Gay and Williams and ourselves are essentially isothermal and those of Tomkins and Brown, necessarily, non-isothermal. Discrepancies exist for runs

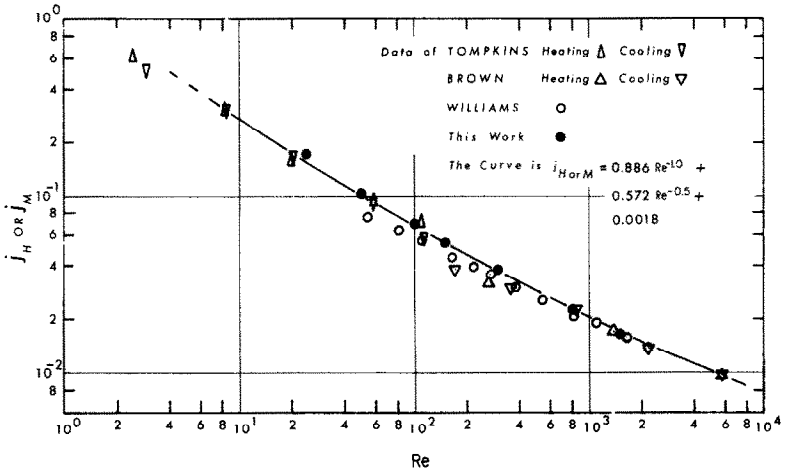

FIG. 3. Plot of j-factor against Reynolds number Re, for 43.7% cutdown case (Aston data).

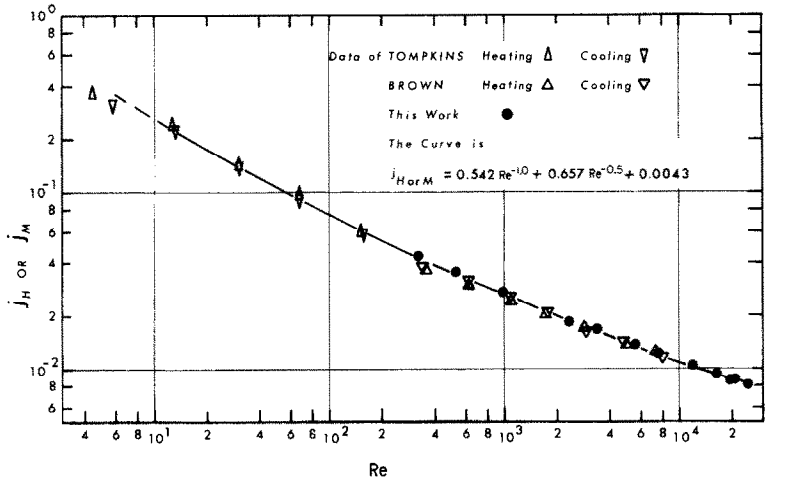

FIG. 4. Plot of j-factor against Reynolds number Re, for 18.4% cutdown case (Winfrith data).

DISCUSSION OF RESULTS

The experimental results were expressed in terms of the j-factor of Chilton and Colburn [10] and the Reynolds number. The fluid velocity used in the calculations is based on the minimum flow area at the centre row of tubes, following Gay and Williams [16] and Bergelin *et al.* [11], see Fig. 2.

Bundle-average coefficients could be calculated for each of the configurations studied from a simple arithmetic average over all the tubes, since mass transfer with constant bulk concentration is analogous to isothermal heat transfer. The initial comparison possible

carried out at the lower flow-rates where the temperature rises through the Delaware exchanger are greatest (of the order of 25°F) and negligible for the high flow-rate runs where the temperature rises are small (of the order of 5°F). But these low flow-rate runs are also the ones with the greater uncertainty in the heat balances [11]. In addition the data of Tompkins and Brown are overall figures for a complete exchanger, whereas the remaining data were obtained in a single baffle compartment in the body of the exchanger.

A further comparison is provided by an additional case studied at AEE Winfrith. Figure 4 shows the data

of Tompkins, of Brown, and of Mackley [27] for the no-leakage 18.4% baffle cut-down, 1.91 in (48.5 mm) baffle spacing case.

The agreement is in general excellent, but once again there is a tendency for the low flow-rate high temperature rise cases of both Tompkins and Brown to fall below the isothermal line.

ERROR ANALYSIS

In order to estimate the maximum error bounds all the uncertainties in the experimental quantities were regarded as random and the usual statistical error analysis carried out on the present data. This gave 95% confidence limits of $\pm 3\%$ being due to those quantities which varied from run to run, the largest contribution being due to a $\pm 1\%$ uncertainty in the flowrate. A reproducibility test was carried out at Winfrith when a particular configuration was rebuilt and retested after a gap of three months. The data from both runs fell within $\pm 3\%$ of a single smooth curve drawn through both sets.

No detailed error analysis appears to have been carried out by Bergelin and co-workers who merely state that their flowrates were known to $\pm 1\%$ and that their heat balances were correct to within 3%, except at the lowest flow rate due to experimental difficulties in establishing the temperature of the exit oil stream. Hence a least possible error band is indicated of about $\pm 6\%$. Williams [30] gave no estimate of experimental error but had to apply a correction to each data point of about 30%.

VALIDITY OF THE METHOD

The agreement between the data obtained by direct measurement and that from mass-transfer measurements using the redox method of this work shown in Figs. 3 and 4 is such as to demonstrate both the validity of the Chilton–Colburn analogy for this situation and also the degree of accuracy obtained with the redox technique. This conclusion is strengthened by the results of the error analysis.

CORRELATION OF THE RESULTS

The data discussed in this paper were obtained from essentially only two configurations of a small heat exchanger. They cannot therefore form a basis for any generalised correlation intended for design purposes. But a correlation may usefully serve another purpose, namely that of summarising the data obtained from each of a number of configurations in a form suitable for use in subsequent analysis. Results from such an analysis will be published in a future paper. The data of Fig. 4 cover a wide range of Re and of property number, Pr or Sc, see Table 3. As such they are useful as a data-set to evaluate the various possible correlations. Only a limited number of analytical relationships have been presented for flow across tube banks [31, 33, 34]. Those in which Re is raised to a single power obviously cannot correlate the data shown in Fig. 4. Whitaker [33] has presented a generalised correlation applicable to numerous situations which include cross-flow in tube banks and flow in packed beds. It is of similar form to an earlier one for cross-flow over single tubes due to Richardson [40]. Table 2 gives the form of his correlation when expressed in terms of j_H rather than Nu, equation (a). More complex correlations have been proposed for this situation. Several of these were converted to their j_H or j_M form as summarised in Table 2 [equations (e), (f) and (g)] and were tested for their applicability. The correlation of White and Churchill [32] has already been shown by Gay and Williams not to correlate their data below $Re = 200$. Additional models have been obtained by relaxing the powers in Whitaker's model, to give equations (b), (c), (d) and (h). Each equation was fitted to the data by an unconstrained Nelder and Mead [35] LSQ fitting routine, starting from more than one initial parameter set and using the same convergence criterion for each case. The low Re results of both Tompkins and Brown which may not be typical of isothermal operation were not used in the fit (a total of six points was discarded).

Table 2. Correlations tested

Equation	Empirical formula	Source
(a)	j_H or $j_M = ARe^{-0.5} + BRe^{-0.33}$	Whitaker [33]
(b)	j_H or $j_M = ARe^{-C} + BRe^{-0.3}$	Richardson [40]
(c)	j_H or $j_M = ARe^{-C} + BRe^{-0.33}$	Relaxed forms of above
(d)	j_H or $j_M = ARe^{-C} + BRe^{-0.35}$	
(e)	j_H or $j_M = ARe^{-1.0} + BRe^{-0.5} + C$	Van der Hegge Zijnen [37]
(f)	j_H or $j_M = ARe^{-1.0} + BRe^{C-1}$	Van der Hegge Zijnen [37]
(g)	j_H or $j_M = ARe^{-1.0} + BRe^{-0.5} + CRe^{D-1}$	Fand and Keswani [36]
(h)	j_H or $j_M = ARe^{-0.8} + BRe^{-0.33}$	Modified Whitaker

Table 3. Use of the two-thirds power on the property number

Source	Technique	Property number Pr	Property number Sc
Tompkins [29]	Heat transfer	1150	
Brown [11]	Heat transfer	29.2	
Williams [16]	Mercury evaporation		1.22
Mackley [27]	Redox couple		1570
Lucas* [38, 39]	Heat transfer	0.7	
	Naphthalene evaporation		2.46
	Redox couple		1600

*Rapid billet heating furnace.

All the three-parameter equations fitted the data well. Equation (a) does not give as good a fit, but if the powers of Re are set at 0.8 and 0.33, as in equation (h), then the fit is nearly as good as the three-parameter equations (b), (c), and (d). Varying the 0.33 power in the three-parameter models does not improve the fit—a four-parameter equation has too many degrees of freedom. This supposition is confirmed by the behaviour of the four-parameter equation (g). The two starting points produced two different sets of parameters which minimised the objective function to almost same value —in addition parameter D moved to a value of 0.52, so that Re appeared twice varied to almost the same power.

As already stated, all the three-parameter equations give a good representation of the data, equation (e) being marginally superior to the others. Since this also has fixed powers on Re this equation has been chosen to fit both the data of this work and of subsequent work. The preferred correlation therefore is

$$j_{H \text{ or } M} = ARe^{-1.0} + BRe^{-0.5} + C.$$

For the data of Fig. 4, $A = 0.542$, $B = 0.657$ and $C = 0.0043$. When the data of Fig. 3 were fitted by this equation the coefficients $A = 0.866$, $B = 0.572$ and $C = 0.0018$ were obtained.

EFFECT OF PROPERTY NUMBER

The Chilton–Colburn analogy assumes a two-thirds power on the property number (Pr or Sc) in the j_H and j_M factors. The data in Figs. 3 and 4 cover a wide range of property number (see Table 3) and the good agreement shown confirms the validity of this power. Table 3 also includes the property number range of Lucas [38, 39] who reached a similar conclusion in a study of rapid heating billet furnaces.

CONCLUSIONS

The electrochemical mass-transfer modelling technique adopted has been shown to give good results in the determination of transfer coefficients in shell-and-tube heat exchangers.

The error analysis shows it to be of higher accuracy than the alternative techniques, as well as being more convenient to use. In particular, since the coefficients are deduced from current measurements, the method lends itself to rapid data acquisition using modern data-logging techniques, a facility made use of on the second rig at AEE Winfrith.

A suitable correlating equation for average transfer coefficients is

$$j_{H \text{ or } M} = ARe^{-1.0} + BRe^{-0.5} + C,$$

similar in form to the equation of Van der Hegge Zijnen for single cylinders in cross-flow. The effect of fluid property variations is adequately represented by an 0.67 power on Pr or Sc in the calculation of j_H or j_M.

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TRANSFERT DE CHALEUR SUR L'ENVELOPPE DANS LES ECHANGEURS TUBULAIRES AVEC BAFFLES ET A ENVELOPPE CYLINDRIQUE—UNE METHODE DE MODELISATION ELECTROCHIMIQUE PAR TRANSFERT DE MASSE

Résumé—L'article décrit une application d'une méthode de modélisation électrochimique, par transfert de masse, à la détermination des coefficients de transfert thermique locaux sur l'enveloppe, dans un type d'échangeur de chaleur tubulaire avec baffles et à enveloppe cylindrique. La validité et la précision de la méthode électrochimique sont mises en évidence par comparaison avec des mesures de transfert thermique et avec des données de transfert massique obtenues à l'aide d'une technique d'évaporation du mercure. On présente également un calcul d'erreur.

Afin de regrouper les données obtenues dans ces travaux (et les travaux suivants), un certain nombre de formules ont été testées et les équations telles que:

$$j_M = ARe^{-1,0} + BRe^{-0,5} + C$$

ont fourni la meilleure approximation.

DER MANTELSEITIGE WÄRMEÜBERGANG IM ZYLINDRISCHEN ROHRBÜNDEL—EIN ELEKTROCHEMISCHES STOFFÜBERGANGS-MODELLVERFAHREN

Zusammenfassung—Die Arbeit beschreibt die Anwendung eines elektrochemischen Stoffübergangs-Modellverfahrens zur Bestimmung der örtlichen, mantelseitigen Wärmeübergangskoeffizienten im Modell eines zylindrischen Rohrbündelwärmeübertragers mit Umlenkblechen. Die Gültigkeit und die Genauigkeit der elektrochemischen Methode wird anhand eines Vergleichs mit Wärmeübergangsmessungen und mit Stoffübergangsdaten, die mit der Quecksilber-Verdampfungstechnik erhalten worden sind, aufgezeigt. Eine Fehlerrechnung wird ebenfalls durchgeführt.

Um die Ergebnisse dieser (und nachfolgender) Arbeit zusammenfassen zu können, wurde eine Reihe von Korrelationsgleichungen auf ihre Eignung überprüft. Am geeignetsten erwies sich dabei die Beziehung

$$j_M = ARe^{-1,0} + BRe^{-0,5} + C.$$

ТЕПЛООБМЕН В МЕЖТРУБНОМ ПРОСТРАНСТВЕ ЦИЛИНДРИЧЕСКИХ КОЖУХНО-ТРУБНЫХ ТЕПЛООБМЕННИКОВ С ПЕРЕГОРОДКАМИ. ЭЛЕКТРОХИМИЧЕСКИЙ МЕТОД МОДЕЛИРОВАНИЯ МАССООБМЕНА.

Аннотация — Описывается применение электрохимического метода моделирования массообмена для определения локальных коэффициентов теплообмена в межтрубном пространстве цилиндрического кожухно-трубного теплообменника с перегородками. Путем сравнения измеренных коэффициентов теплообмена с результатами по массообмену, полученными при использовании метода испарения ртути, демонстрируются справедливость и точность электрохимического метода. Представлен также анализ погрешностей. В целях суммирования данных этой и последующей работ были выверены ряд соотношений и найдено, что наиболее подходящими являются уравнения вида

$$j_M = ARe^{-1,0} + BRe^{-0,5} + C.$$