



Condensation inside tubes: condensate film coefficient for horizontal tubes

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THE PREPARATION OF THIS DATA ITEM

The work on this particular Item was monitored and guided by the following Working Party on behalf of the Heat Transfer Steering Group:

Mr J.M. McNaught	— National Engineering Laboratory
Mr M.J.C. Moore	— John Brown Engineers and Constructors Ltd
Prof. J.W. Rose	— Queen Mary and Westfield College, London.

The Heat Transfer Steering Group first met in 1966 and now has the following membership:

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Dr A. C. Pauls*	— Independent
Prof. R.H. Sabersky*	— California Institute of Technology, USA
Mr E.A.D. Saunders	— Independent
Dr J. Taborek*	— Independent
Eur. Ing. M.A. Taylor	— Independent
Mr G.H. Walter	— Independent
Eur. Ing. D.A. Winkworth	— Foster Wheeler Energy Ltd.

* Corresponding Member

The technical work involved in the assessment of the available information and the construction and subsequent development of the Data Item was undertaken by

Mr G.H. Walter	— Independent.
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The person with overall responsibility for the work in this subject area is Mr S. J. Pugh, Head of Process Engineering Technology.

CONDENSATION INSIDE TUBES: CONDENSATE FILM COEFFICIENT FOR HORIZONTAL TUBES

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CONDENSATION INSIDE TUBES: condensate film coefficient for horizontal tubes

1. NOTATION AND UNITS

		Units	
		SI	British
c_p	specific heat capacity	J/kg K	Btu/lb °F
D	internal diameter of tube	m	ft
g	acceleration due to gravity	m/s ² (9.81)	ft/h ² (4.17 × 10 ⁸)
h_{fg}	specific enthalpy of vaporisation (latent heat)	J/kg	Btu/lb
j_g^*	gas phase dimensionless superficial velocity, defined in Equation (3.2)	—	—
$\dot{M}$	total mass flow rate per tube = ($\dot{M}_g + \dot{M}_l$)	kg/s	lb/h
Pr	Prandtl number = $\eta c_p / \lambda$	—	—
$Re_{f,x}$	condensate film Reynolds number at vapour mass fraction x , defined in Equation (3.13)	—	—
$Re_{g,x}$	vapour Reynolds number at vapour mass fraction x , defined in Equation (3.7)	—	—
Re_l	enhanced liquid phase Reynolds number, defined in Equation (3.16)	—	—
T	temperature	K or °C	°F
U	overall heat transfer coefficient referred to inside diameter of tube	W/m ² K	Btu/ft ² h °F
U'	combined heat transfer coefficient for coolant, tube wall and any fouling referred to inside diameter of tube	W/m ² K	Btu/ft ² h °F
v_g	superficial vapour phase velocity = $4\dot{M}_g/(\pi\rho D^2)$	m/s	ft/h
X_{tt}	Martinelli parameter, defined in Equation (3.3)	—	—
x_g	vapour mass fraction = $4\dot{M}_g/(\dot{M}_g + \dot{M}_l)$	—	—
α	heat transfer coefficient	W/m ² K	Btu/ft ² h °F
ΔT	temperature difference across the condensate film = ($T_{sat} - T_w$)	K or °C	°F
ε_G	void fraction	—	—

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η	dynamic viscosity	kg/m s (Pa s)	lb/ft h
Λ	dimensionless group defined in Equation (3.12)	—	—
λ	thermal conductivity	W/m K	Btu/h ft °F
μ	relative error in prediction of condensate film coefficient	—	—
ρ	fluid density	kg/m ³	lb/ft ³
τ_i^*	dimensionless interfacial shear stress defined in Equations (3.10) and (3.11)	—	—
Ω	physical properties parameter defined in Equation (3.8)	W/(m ^{7/4} K ^{3/4})	Btu/(ft ^{7/4} h °F ^{3/4})

Additional Subscripts

c	condensate film
cs	cold stream
$expt$	experimentally measured value
g	gas or vapour phase
$init$	initial value
l	liquid phase
m	mean value
$pred$	predicted value
sat	at saturation
w	at vapour side surface of any dirt film present, as appropriate
x	at vapour mass fraction x_g

2. PURPOSE AND SCOPE OF THIS ITEM

This is the third^{*} of a series of Data Items on heat transfer in the condensation of **single component fluids** inside tubes of circular cross-section. The first, Data Item ESDU 89038⁵, deals with reflux condensation in vertical tubes, where the vapour flows up the tubes and the condensate drains downwards. The second, Data Item ESDU 91024⁸, deals with the estimation of the heat transfer coefficient of the condensate film in vertical tubes with co-current downflow of vapour and condensate. This, the third Item, deals similarly with flow through horizontal tubes. It presents the available data and provides correlations for predicting the local condensate film coefficient at a prescribed vapour mass fraction.

The nature of the condensation process is outlined in the introductory section (Section 2) of ESDU 91024⁸, and the factors influencing the choice of in-tube condensation rather than shell-side condensation are stated in Section 4 of that Item.

3. CONDENSATE FILM COEFFICIENT

3.1 Flow Regimes in Horizontal Tube-side Condensation

The hydrodynamics of the condensate film must be taken into consideration and the flow patterns known in order to select the appropriate correlation for calculating the condensate film coefficient. Different flow regimes require different forms of correlations. In the case of horizontal tube orientation there are two basic flow patterns: annular flow where shear forces predominate and stratified flow where gravity forces predominate. The onset of condensation can occur over a wide range of vapour velocities.

Annular flow, where the liquid flows on the wall of the tube as a film and the vapour flows in the centre, occurs at high vapour flow rates. At the highest vapour rates the vapour phase may carry a significant proportion of the liquid phase in the form of entrained droplets. Entrainment becomes less significant as condensation proceeds and the film thickens. The condensate film coefficient is then controlled by vapour shear at the vapour-liquid interface.

At lower vapour velocities vapour shear becomes less important and gravity more important in defining the flow pattern until stratified flow is attained.

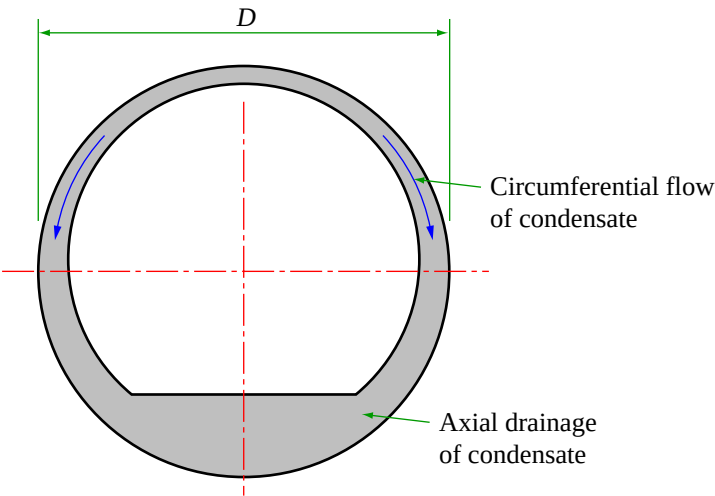
In the limit, at very low vapour velocities, flow in the upper part of the tube is governed by gravity. The flow in the condensate film as it runs down the side walls is predominately circumferential as shown in Sketch 3.1 and the film thickens from a small (but finite) value at the top of the tube until it reaches the pool at the bottom. In stratified flow, vapour shear can still be significant but becomes progressively less important as the vapour velocity tends to zero.

The condensate that collects at the bottom of the tube drains in the axial direction as a result either of the pressure gradient (vapour shear) or the hydraulic gradient as shown in Sketch 3.2.

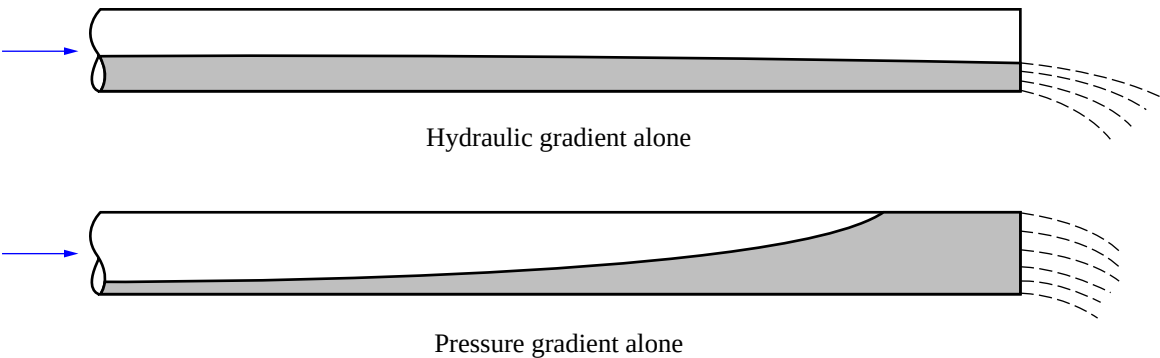
In practice drainage flow profiles will usually be some combination of those due to both the hydraulic and pressure gradients, but the relative magnitudes of these gradients, and hence the specific interface profile in the stratified flow zone, cannot be properly ascertained.

A typical vapour-condensate interface profile is shown in Sketch 3.3a. Occasionally the inlet flow rate and liquid loading (amount of condensate formed) is so high that stratified flow does not occur. As condensation proceeds the transition will then be to slug and plug flow. This situation is illustrated in Sketch 3.3b.

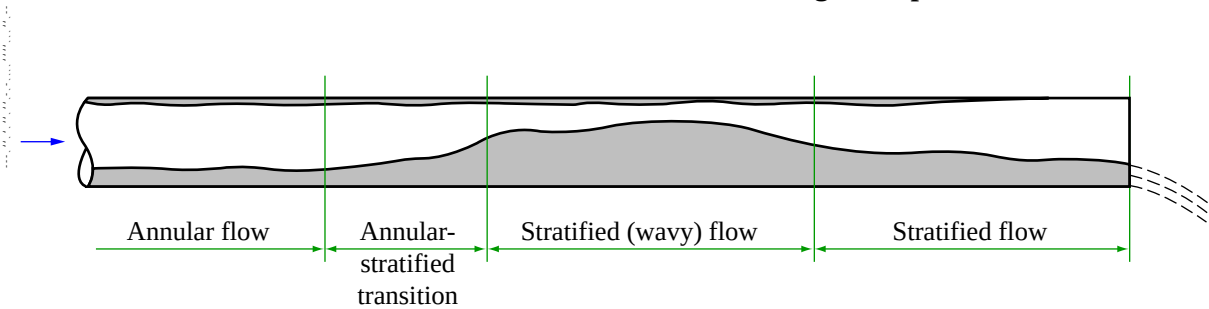
* Additionally there are three Data Items, ESDU 90024⁶, 91023⁷ and 93014⁹ concerned with the prediction of pressure drop in vertical downflow and horizontal tubes; the latter Data Item describes a computer program, ESDUpac A9314, for pressure drop prediction.



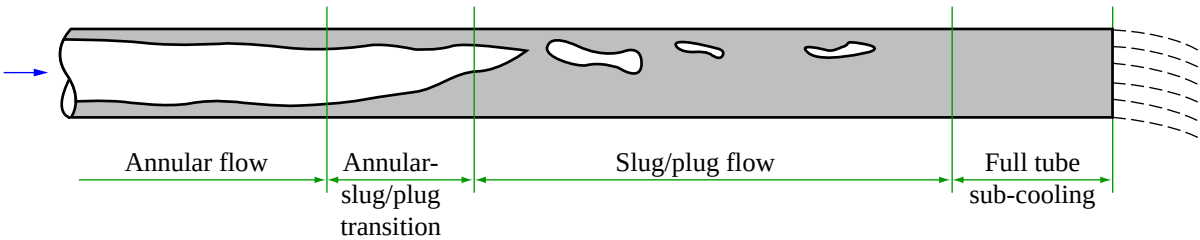
Sketch 3.1 Idealised cross-section of stratified flow at low vapour velocities



Sketch 3.2 Idealised condensate drainage flow profiles

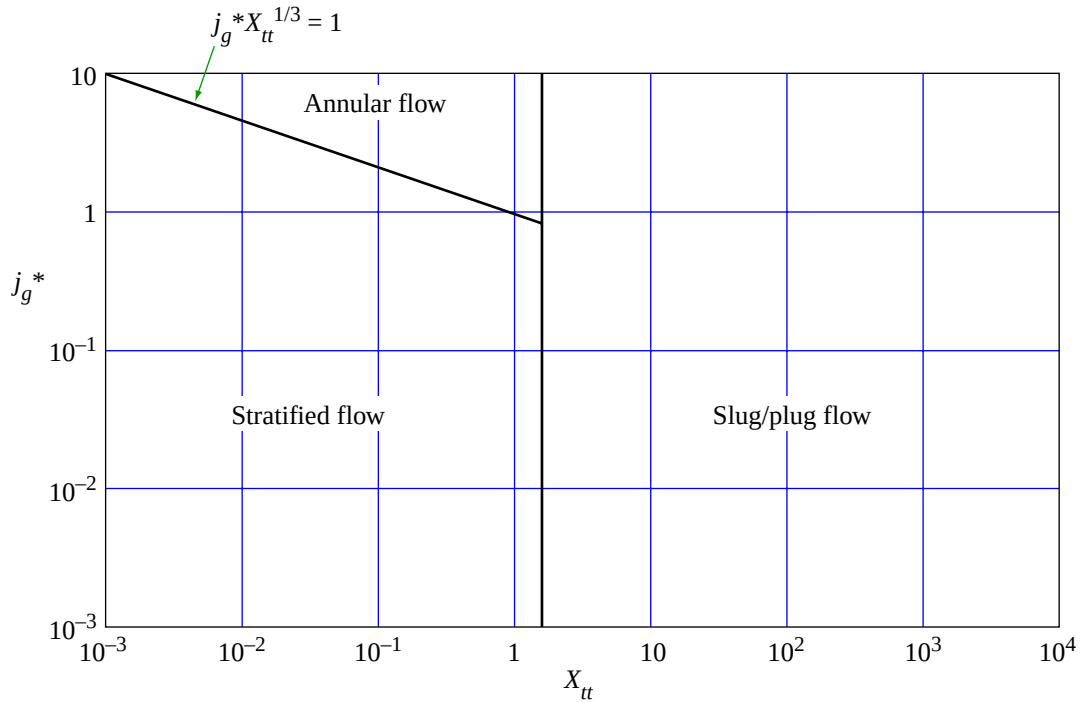


Sketch 3.3a Normal interface profile



Sketch 3.3b Interface profile for high liquid loadings

3.2 Flow Regime Transitions



Sketch 3.4 Modified flow regime map for horizontal two-phase flow.
After Taitel and Dukler² and Breber *et al.*³

3.2.1 Annular to stratified transition

The criterion adopted here for the annular to stratified flow transition is

$$j_g^* X_{tt}^{1/3} = 1 \quad (3.1)$$

* where

$$j_g^* = \frac{\dot{M}_g}{\frac{\pi}{4} D^2 [D \rho_g (\rho_l - \rho_g) g]^{1/2}}, \quad (3.2)$$

the gas phase dimensionless superficial velocity, and the Martinelli parameter is

$$X_{tt} = \left[\frac{\text{frictional pressure gradient assuming liquid flowing alone}}{\text{frictional pressure gradient assuming gas flowing alone}} \right]^{1/2} \\ \approx \left(\frac{1 - x_g}{x_g} \right)^{0.9} \left(\frac{\rho_g}{\rho_l} \right)^{0.5} \left(\frac{\eta_l}{\eta_g} \right)^{0.1}. \quad (3.3)$$

The validity of Equation (3.3) assumes both phases are in turbulent flow. On account of the data scatter this expression may also be used in Equation (3.1) even though one or both phases may be in laminar flow.

* The j_g^* parameter may be rearranged as a Froude number modified by the density ratio, as follows

$$j_g^* = \left(\frac{\rho_g}{\rho_l - \rho_g} \right)^{1/2} \frac{v_g}{(Dg)^{1/2}}.$$

Use of the parameter $j_g * X_{tt}^{1/2}$ is in accordance with the theoretical model developed by Taitel and Dukler² and represents a linearisation of their stratified/annular transition (see Sketch 3.4). The value of unity chosen for the constant on the right hand side of Equation (3.1) has been derived from the data, as outlined in Appendix A.

3.2.2 Transition to slug/plug flow

In the great majority of condensing situations the flow will proceed from annular to stratified. The available data for confirming the slug/plug boundary under condensing conditions are few. Consequently the criterion adopted here is

$$X_{tt} = 1.5, \quad (3.4)$$

as suggested by Breber *et al.*³, Equation (3.4) is simpler and more conservative than the Taitel and Dukler² demarcation.

3.3 Correlation for the Local Condensate Film Coefficient

3.3.1 Correlations for stratified and annular flows

The scope of the data bank used in deriving Equation (3.6), and confirming the validity of Equations (3.5) and (3.6), is given in Table 11.1. The ranges of the data are given in Section 4 and comparison of predictions with the data are given in Appendix A.

For $j_g * X_{tt}^{1/3} \leq 1$, $X_{tt} \leq 1.5$ and $Re_{g,x} \leq 7780$,

$$\alpha_{c,x} = \frac{0.54\Omega}{(D\Delta T)^{0.25}}. \quad (3.5)$$

For $j_g * X_{tt}^{1/3} \leq 1$, $X_{tt} \leq 1.5$ and $Re_{g,x} > 7780$,

$$\alpha_{c,x} = 0.09 Re_{g,x}^{0.2} \frac{\Omega}{(D\Delta T)^{0.25}} \quad (3.6)$$

where $Re_{g,x} = \frac{4\dot{M}_g}{\pi D \eta_g}$ (3.7)

and $\Omega = \left[\frac{\lambda_l^3 \rho_l (\rho_l - \rho_g) g h_{fg}}{\eta_l} \right]^{1/4}$. (3.8)

For $j_g * X_{tt}^{1/3} > 1.0$

$$\alpha_{c,x} = \lambda_l \left[\frac{\rho_l (\rho_l - \rho_g) g}{\eta_l^2} \right]^{1/3} \left[\frac{\rho_l^{1.5} \tau_i^*}{771.6} \right]^{1/2} \quad (3.9)$$

where $\tau_i^* = \Lambda \left(\frac{x_g}{1 - x_g} \right)^{1.4} Re_{f,x}^{1.8}$ for $x_g \leq 0.81$ (3.10)

and $\tau_i^* = 0.383 \Lambda \left(\frac{Re_{f,x}}{1 - x_g} \right)^{1.8}$ for $x_g > 0.81$. (3.11)