A SHORT INTRODUCTION TO TWO-PHASE FLOWS

Condensation and boiling heat transfer

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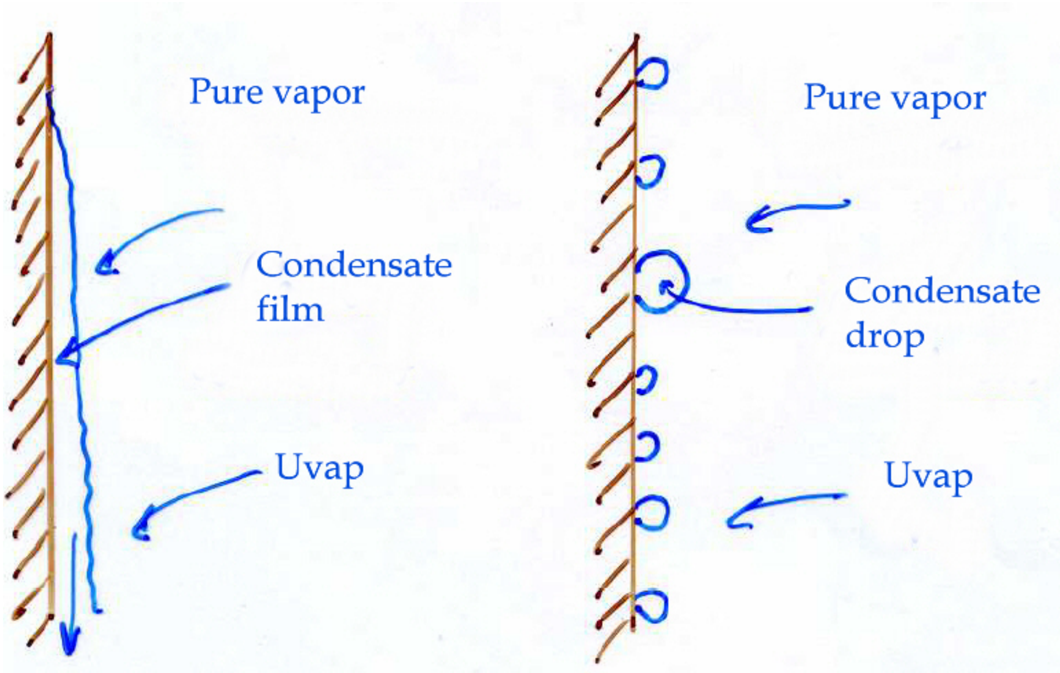
HEAT TRANSFER MECHANISMS

- Condensation heat transfer:
  - drop condensation
  - film condensation
- Boiling heat transfer:
  - Pool boiling, natural convection, *ébullition en vase*
  - Convective boiling, forced convection,
- Only for pure fluids. For mixtures see specific studies. Usually in a mixture, $h \leqslant \sum x_i h_i$ and possibly $\ll h_i$.
- Many definitions of heat transfer coefficient,

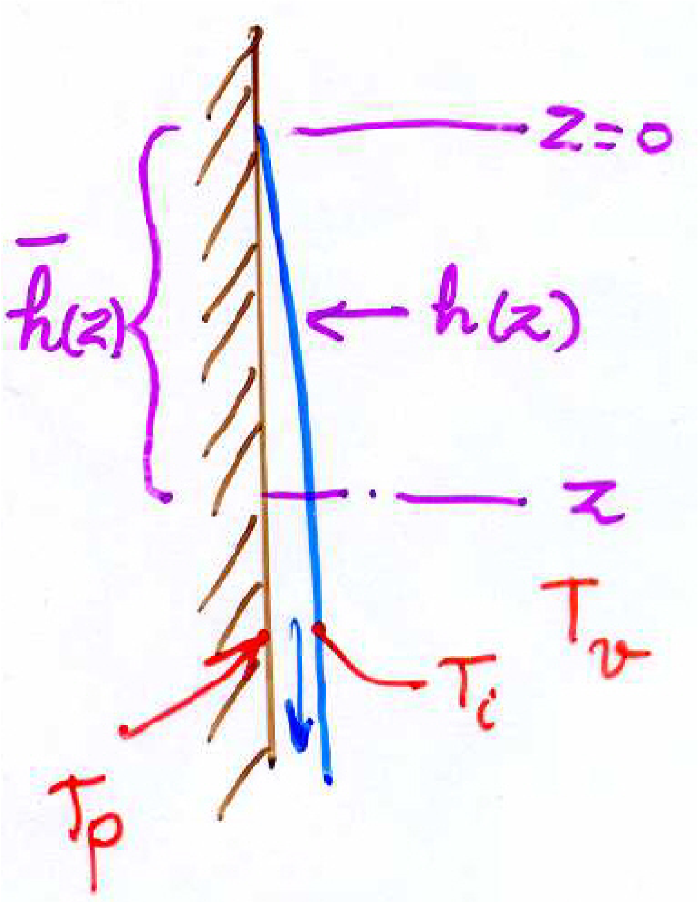
$$h[\mathrm{W/m^2/K}] = \frac{q}{\Delta T}, \quad \mathrm{Nu} = \frac{hL}{k}, \quad k(T?)$$

CONDENSATION OF PURE VAPOR



- Flow patterns
  - Liquid film flowing.
  - Drops, static, hydrophobic wall ($\theta \approx \pi$). Clean wall, better htc.
- Fluid mixture non-condensible gases:
  - Incondensible accumulation at cold places.
  - Diffusion resistance.
  - Heat transfer deteriorates.
  - Traces may alter significantly h

FILM CONDENSATION



- Thermodynamic equilibrium at the interface,

$$T_i = T_{\text{sat}}(p_\infty)$$

- Local heat transfer coefficient,

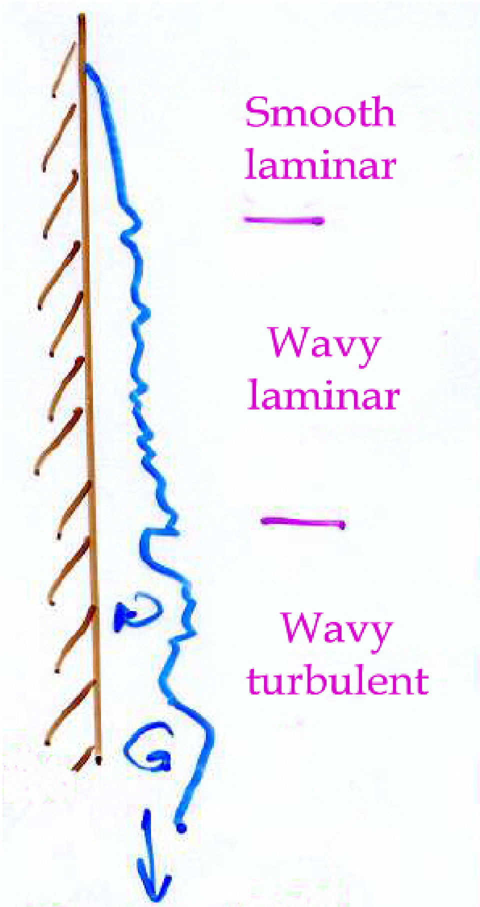
$$h(z) \triangleq \frac{q}{T_i - T_p} = \frac{q}{T_{\text{sat}} - T_p}$$

- Averaged heat transfer coefficient,

$$\overline{h}(L) \triangleq \frac{1}{L}\int_0^L h(z)\mathrm{d}z$$

- NB: Binary mixtures $T_i(x_\alpha, p)$ and $p_\alpha(x_\alpha, p)$. Approximate equilibrium conditions,
  - For non condensible gases in vapor, $p_V = xP_{\text{sat}}(T_i)$, Raoult relation
  - For dissolved gases in water, $p_G = Hx_G$, Henry's relation

CONTROLLING MECHANISMS



- Slow film, little convective effect, conduction through the film (main thermal resistance)
- Heat transfer controlled by film characteristics, thickness, waves, turbulence.
- Heat transfer regimes,

$$\Gamma \triangleq \frac{M_L}{\mathcal{P}}, \quad \mathrm{Re}_F \triangleq \frac{4\Gamma}{\mu_L}$$

  - Smooth, laminar, $\mathrm{Re}_F < 30$,
  - Wavy laminar, $30 < \mathrm{Re}_F < 1600$
  - Wavy turbulent, $\mathrm{Re}_F > 1600$

CONDENSATION OF SATURATED STEAM

- Simplest situation, only a single heat source: interface, stagnant vapor,
- Laminar film (Nusselt, 1916, Rohsenow, 1956), correction 10 to 15%,

$$h(z) = \left(\frac{k_L^3 \rho_L g(\rho_L - \rho_V)(h_{LV} + 0,68 C_{PL}[T_{\text{sat}} - T_P])}{4\mu_L (T_{\text{sat}} - T_P) z} \right)^{\frac{1}{4}}$$

- Averaged heat transfer coefficient ($T_W = \text{cst}$) : $h(z) \propto z^{-\frac{1}{4}}$, $\overline{h}(L) = \frac{4}{3} h(L)$
- Condensate film flow rate, energy balance at the interface,

$$\Gamma(L) = \frac{\overline{h}(L)(T_{\text{sat}} - T_P)L}{h_{LV}}$$

- Heat transfer coefficient-flow rate relation,

$$\frac{\overline{h}(L)}{k_L} \left(\frac{\mu_L^2}{\rho_L(\rho_L - \rho_V)} \right)^{\frac{1}{3}} = 1,47\,\text{Re}_F^{-\frac{1}{3}}$$

- h_{LV} and ρ_V at saturation. k_L, ρ_L at the *film temperature* $T_F \triangleq \frac{1}{2}(T_W + T_i)$,
- $\mu = \frac{1}{4}(3\mu_L(T_P) + \mu_L(T_i))$, exact when $1/\mu_L$ linear with T.

SUPERHEATED VAPOR

- Two heat sources: vapor ($T_V > T_i$) and interface.
- Increase of heat transfer wrt to saturated conditions, empirical correction,

$$\bar{h}_S(L) = \bar{h}(L) \left(\frac{1 + C_{PV}(T_V - T_{\text{sat}})}{h_{LV}} \right)^{\frac{1}{4}}$$

- Energy balance at the interface, film flow rate,

$$\Gamma(L) = \frac{\bar{h}_S(L)(T_W - T_{\text{sat}})L}{h_{LV} + C_{PV}(T_V - T_{\text{sat}})}$$

FILM FLOW RATE-HEAT TRANSFER COEFFICIENT

- Laminar,

$$\frac{\bar{h}(L)}{k_L}\left(\frac{\mu_L^2}{\rho_L(\rho_L-\rho_V)}\right)^{\frac{1}{3}} = 1,47\,\mathrm{Re}_F^{-\frac{1}{3}}$$

- Wavy laminar and previous regime (Kutateladze, 1963), $h(z) \propto \mathrm{Re}_F^{-0,22}$),

$$\frac{\bar{h}(L)}{k_L}\left(\frac{\mu_L^2}{\rho_L(\rho_L-\rho_V)}\right)^{\frac{1}{3}} = \frac{\mathrm{Re}_F}{1,08\mathrm{Re}_F^{1,22}-5,2}$$

- Turbulent and previous regimes (Labuntsov, 1975), $h(z) \propto \mathrm{Re}_F^{0,25}$,

$$\frac{\bar{h}(L)}{k_L}\left(\frac{\mu_L^2}{\rho_L(\rho_L-\rho_V)}\right)^{\frac{1}{3}} = \frac{\mathrm{Re}_F}{8750+58\mathrm{Pr}_F^{-0,5}(\mathrm{Re}_F^{0,75}-253)}$$

- NB: Implicit relation, Re_F depends on $\overline{h}(L)$ through Γ.

OTHER MISCELLANEOUS EFFECTS

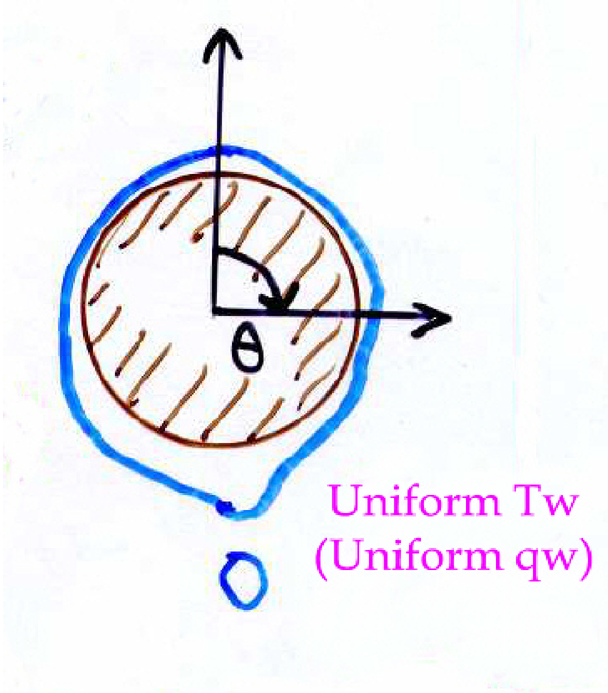
- Steam velocity, v_V, when dominant effect,
- V_v descending flow, vapor shear added to gravity,
- Decreases fil thickness,
- Delays transition to turbulence turbulence,

$$h \propto \tau_i^{\frac{1}{2}}$$

- See for example Delhaye (2008, Ch. 9, p. 370)
- When 2 effects are comparable, h_1 stagnant, h_2 with dominant shear ,

$$h = (h_1^2 + h_2^2)^{\frac{1}{2}}$$

CONDENSATION ON HORIZONTAL TUBES



- Heat transfer coefficient definition,

$$\bar{h} = \frac{1}{\pi}\int_0^{\pi} h(u)\mathrm{d}u$$

- Stagnant vapor conditions, laminar film, Nusselt (1916)

$$\bar{h} = \begin{matrix} 0.728 \\ (0.70) \end{matrix} \left(\frac{k_L^3 \rho_L(\rho_L - \rho_V) g h_{LV}}{\mu_L (T_{\mathrm{sat}} - T_p) D}\right)^{\frac{1}{4}}$$

- 0.728, imposed temperature, 0.70, imposed heat flux.
- Γ, film flow rate per unit length of tube.

- Film flow rate- heat transfer coefficient, energy balance,

$$\frac{\bar{h}}{k_L}\left(\frac{\mu_L^2}{\rho_L(\rho_L-\rho_V)}\right)^{\frac{1}{3}} = \begin{matrix} 1.51 \\ (1.47) \end{matrix} \ \mathrm{Re}_F^{-\frac{1}{3}}$$

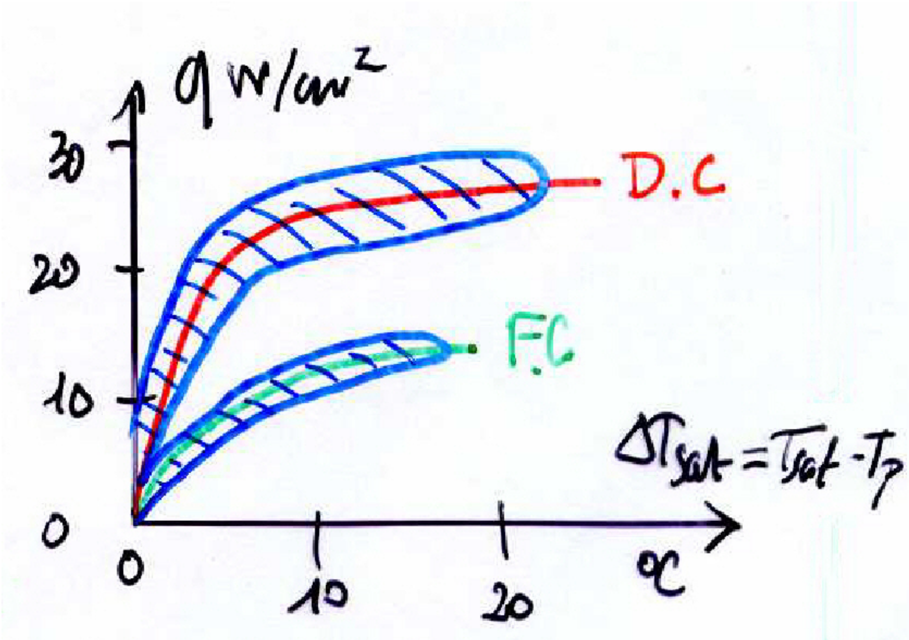
- Vapor superheat and transport proprieties, same as vertical wall
- Effect of steam velocity (Fujii),

$$\frac{\bar{h}}{h_0} = 1.4\left(\frac{u_V^2(T_{\mathrm{sat}}-T_P)k_L}{gDh_{LV}\mu_L}\right)^{0.05} \quad 1 < \frac{\bar{h}}{h_0} < 1.7,$$

- Tube number effect in bundles, (Kern, 1958),

$$\frac{\overline{h(1,N)}}{h_1} = N^{-1/6}$$

DROP CONDENSATION



- Mechanisms,
  - Nucleation at the wall,
  - Drop growth,
  - Coalescence,
  - Dripping down (non wetting wall)
- Technological perspective,
  - Wall doping or coating
  - Clean walls required, fragile
  - Surface energy gradient walls. Self-draining

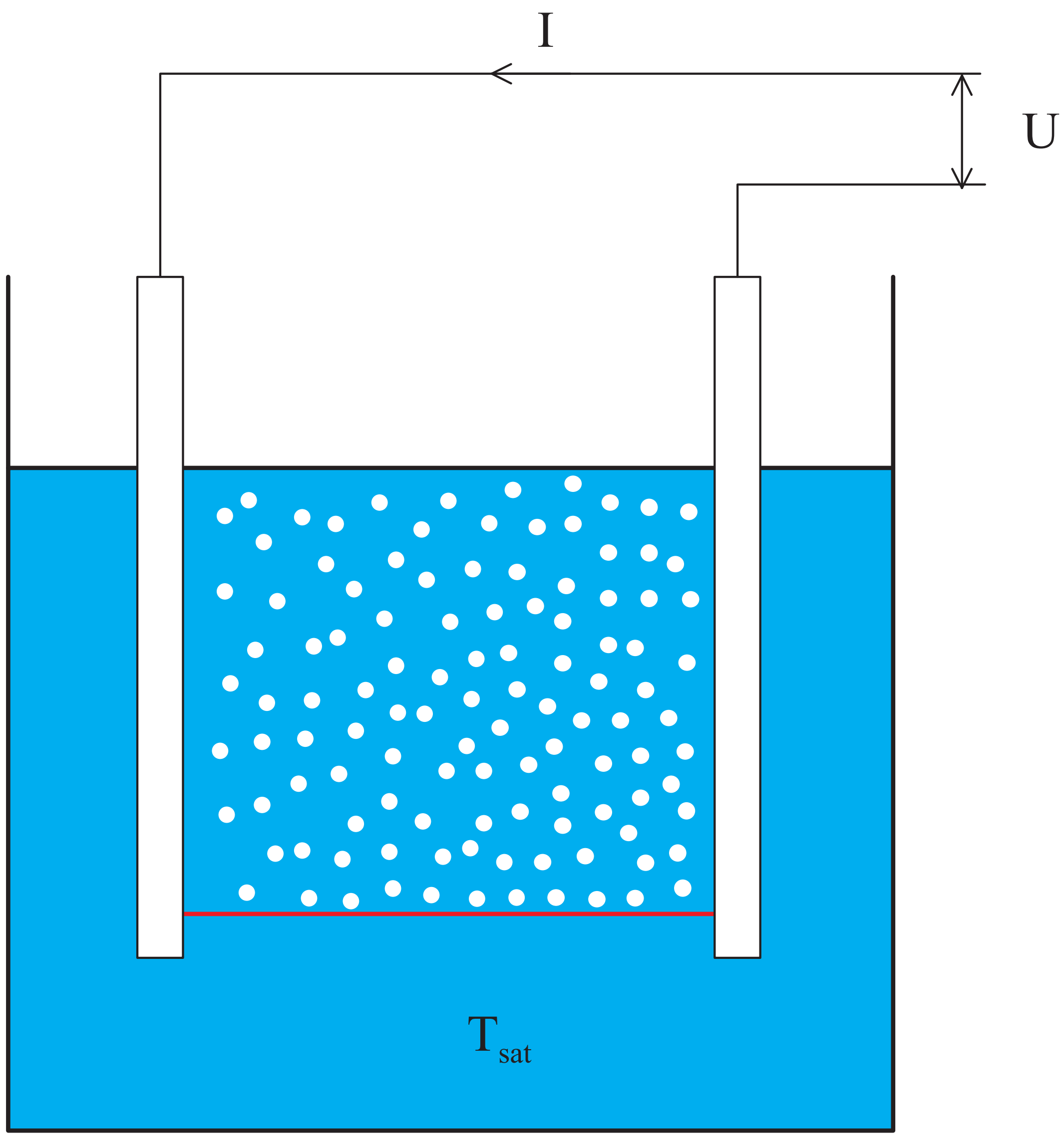
- heat transfer coefficient,

$$\frac{1}{h} = \frac{1}{h_G} + \frac{1}{h_d} + \frac{1}{h_i} + \frac{1}{h_{co}}$$

- G : non-condensible gas, d : drop, i : phase change, co coating thickness.
- Non-condensible gases effect, $\omega_i \approx 0,02 \Rightarrow h \rightarrow h/5$
- Example, steam on copper, $T_{\text{sat}} > 22^{\circ}\text{C}$, h in $\text{W/cm}^2/^{\circ}\text{C}$,

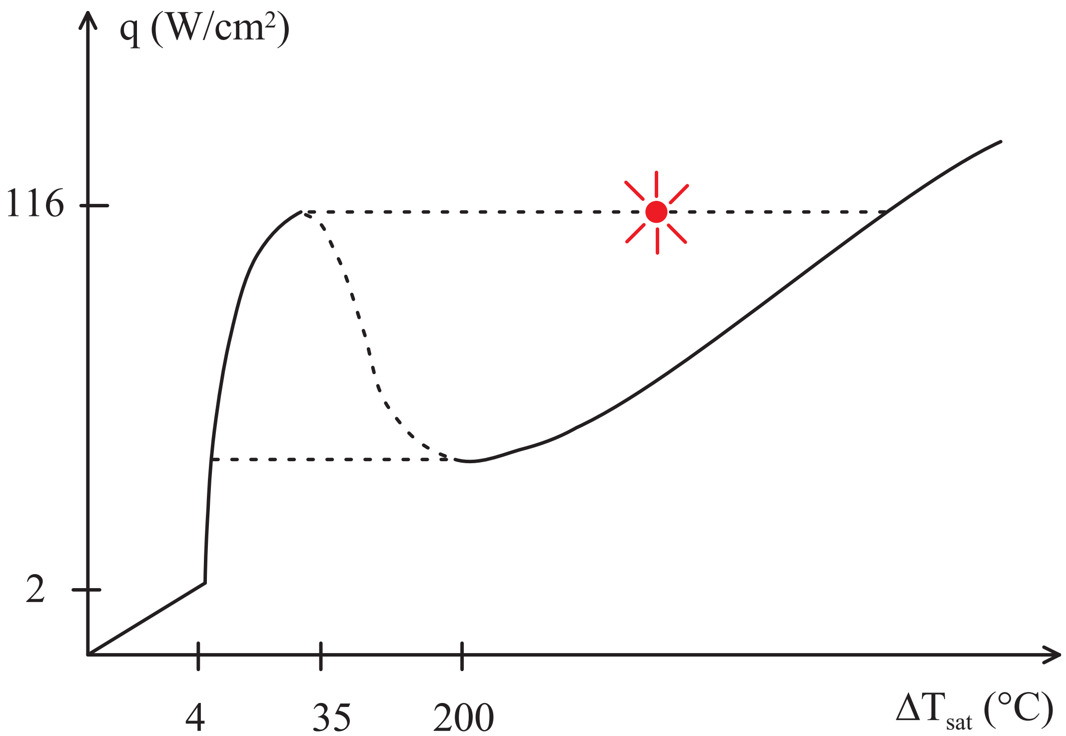
$$h_d = \min(0,5 + 0,2 T_{\text{sat}}, 25)$$

POOL BOILING



- Nukiyama (1934)
- Only one heat sink, stagnant saturated water,
- Wire NiCr and Pt,
  - Diameter: $\approx 50\mu$m,
  - Length: l
  - Imposed power heating: P

BOILING CURVE



- Imposed heat flux,

$$P = q\pi D l = UI$$

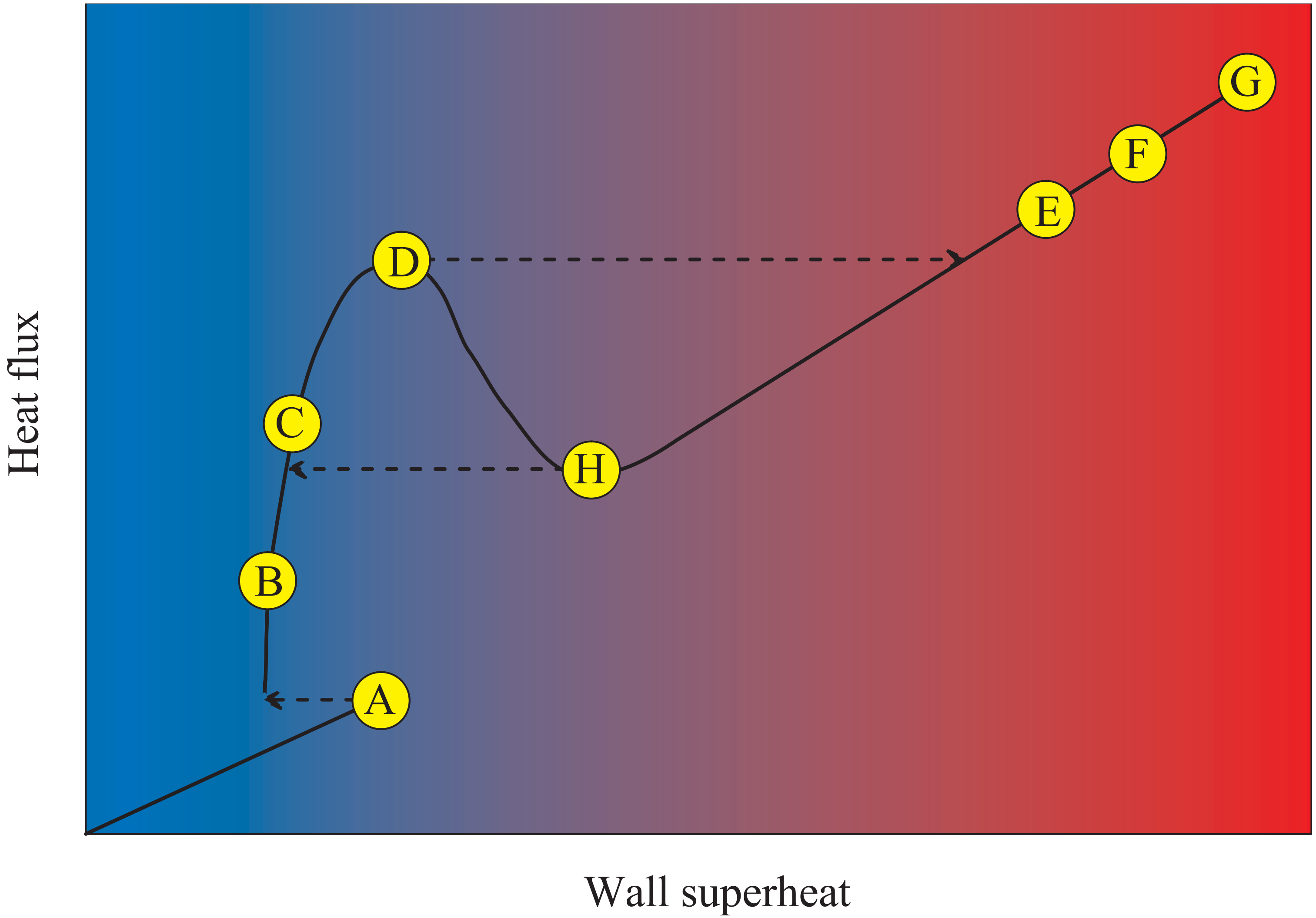
- Wall and wire temperature are equal, $D \to 0$

$$R(T) = \frac{U}{I}, \quad \langle T \rangle_3 \approx T_W$$

- Wall super-heat: $\Delta T = T_W - T_{\text{sat}}$
- Heat transfer coefficient,

$$h \triangleq \frac{q}{T_W - T_{\text{sat}}}$$

BOILING CURVE



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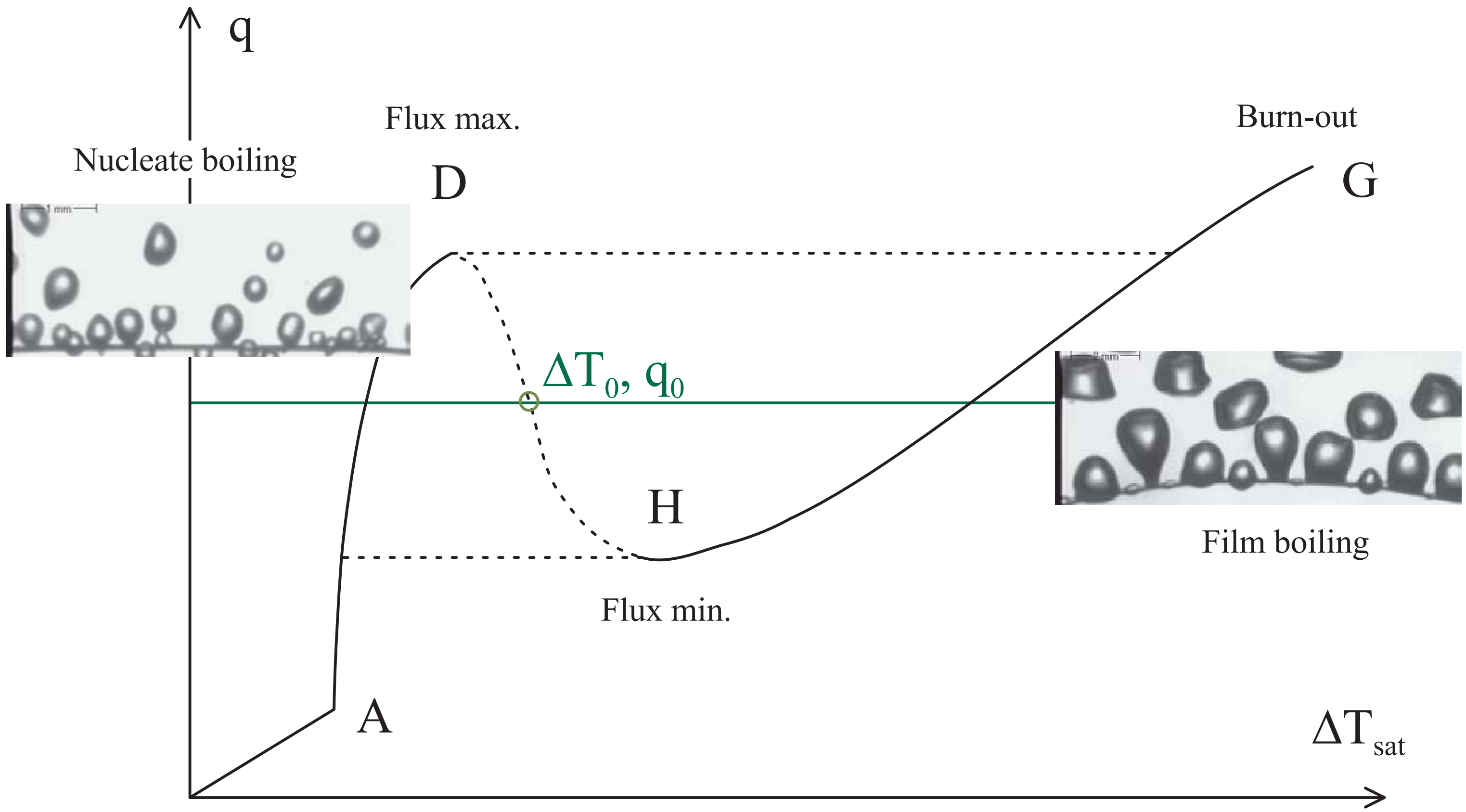
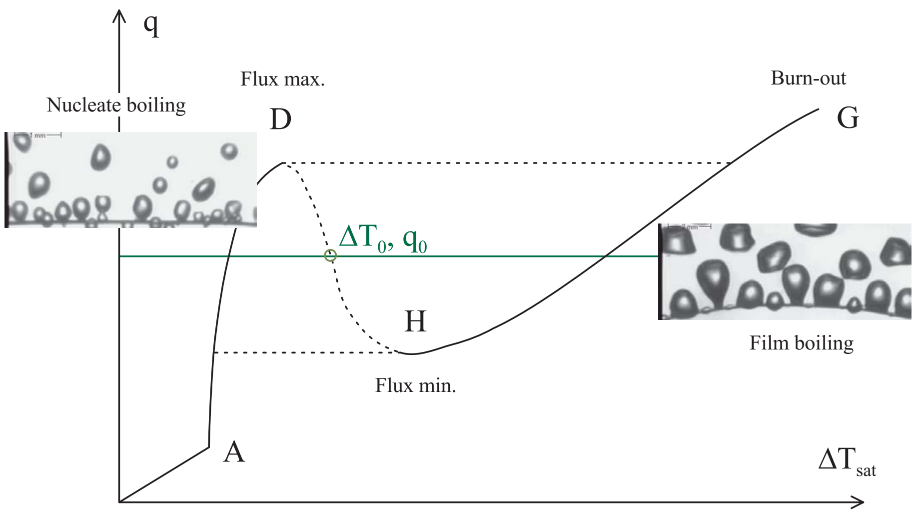
HEAT TRANSFER REGIMES



- OA: Natural convection
- AD: Nucleate boiling
- DH: Transition boiling
- HG: Film boiling

TRANSITION BOILING STABILITY



- Wire energy balance,

$$MC_v \frac{\mathrm{d}T}{\mathrm{d}t} = P - qS$$

- Linearize at $\Delta T_0,\ q_0,\ T = T_0 + T_1,$

$$MC_v \frac{\mathrm{d}T_1}{\mathrm{d}t} = \underbrace{P - q_0 S}_{=0} - S\frac{\partial q}{\partial \Delta T} T_1$$

- Solution, linear ODE,

$$T_1 = T_{10} \exp(-\alpha t), \quad \alpha = \frac{S}{MC_v}\left(\frac{\partial q}{\partial \Delta T}\right)_{T_0}$$

- 2 stable solutions, one unstable (DH),

$$\frac{\partial q}{\partial \Delta T} < 0$$

- Transition boiling, imposed temperature experiments (Drew et Müller, 1937).

NATURAL CONVECTION

- Wire diameter D, natural convection

$$q = h(T_F - T_{\text{sat}}), \quad \text{Nu} = \frac{hD}{k}$$

$$\text{Pr} = \frac{\nu_L}{\alpha_L}, \quad \text{Ra} = \frac{g\beta(T_F - T_{\text{sat}})D^3}{\nu_L \alpha_L}$$

- Nusselt number is the non-dimensional heat transfer coefficient (h).
- k_L, α_L, ν_L at the film temperature $\frac{1}{2}(T_F + T_{\text{sat}})$, β à T_{sat}.
- Churchill & Chu (1975), $10^{-5} < \text{Ra} < 10^{12}$,

$$\text{Nu} = \left(0,60 + \frac{0,387\,\text{Ra}^{1/6}}{\left[1 + \left(\frac{0,559}{\text{Pr}}\right)^{9/16}\right]^{8/27}} \right)^2$$

NATURAL CONVECTION ON A FLAT PLATE

- Scales A, P, plate area and perimeter. Length scale, $L = \frac{A}{P}$.

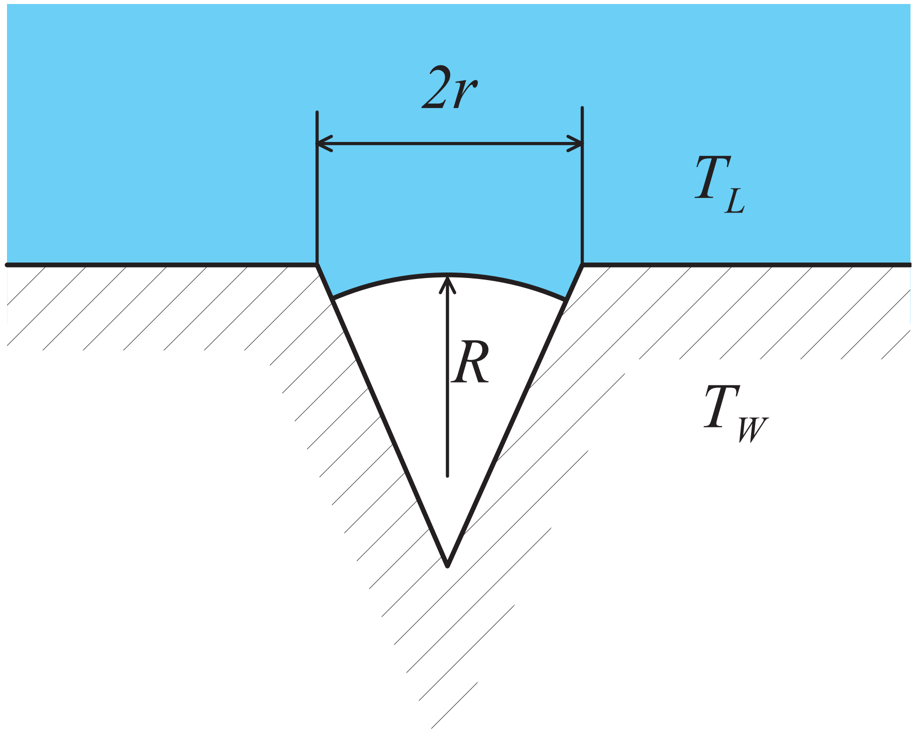
$$\mathrm{Nu} = \frac{hL}{k} = \frac{qL}{k_L(T_P - T_\infty)}, \quad \mathrm{Ra} = \frac{g\beta(T_P - T_\infty)L^3}{\nu_L \alpha_L}$$

- Two regimes,

$$\mathrm{Nu} = \begin{cases} \dfrac{0,560\,\mathrm{Ra}^{1/4}}{\left(1 + (0,492\mathrm{Pr})^{9/16}\right)^{4/9}} & 1 < \mathrm{Ra} < 10^7 \\ 0,14\,\mathrm{Ra}^{1/3}\left(\dfrac{1 + 0,0107\mathrm{Pr}}{1 + 0,01\mathrm{Pr}}\right) & 0,024 \leqslant \mathrm{Pr} \leqslant 2000, \quad \mathrm{Ra} < 2\,10^{11} \end{cases}$$

- Thermodynamic and transport properties Raithby & Hollands (1998). For liquids: all at $T_F = \frac{1}{2}(T_P + T_\infty)$

ONSET OF NUCLEATE BOILING



- Control parameters: p_L et $T_W = T_{L\infty}$
- Super-heated wall: $T_{L\infty} = T_{\text{sat}}(p_L) + \Delta T$
- Site distribution: r, $R = R(r, \theta)$
- Mechanical balance: $p_V = p_L + \frac{2\sigma}{R}$
- Thermodynamic equilibrium:

  $p_V = p_{\text{sat}}(T_{Li}) \Rightarrow T_{Li} = T_{\text{sat}}(p_V)$

  $$T_{Li} = T_{\text{sat}}(p_L + \frac{2\sigma}{R}) \approx (T_{L\infty} - \Delta T) + \frac{2\sigma}{R}\frac{\mathrm{d}T}{\mathrm{d}p}_{\text{sat}}$$

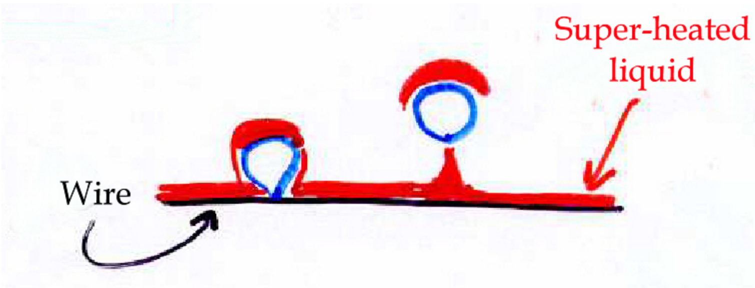
- Heat flux *to* interface: $q > 0$, $\dot{R} > 0$

  $$q = h(T_{L\infty} - T_{Li}) = h\left(\Delta T - \frac{2\sigma}{R}\frac{\mathrm{d}T}{\mathrm{d}p}_{\text{sat}}\right)$$

$$\Delta T > \Delta T_{\text{eq}} = \frac{2\sigma}{R}\frac{\mathrm{d}T}{\mathrm{d}p}_{\text{sat}}, \quad R > R_{\text{eq}} = \frac{2\sigma}{\Delta T}\frac{\mathrm{d}T}{\mathrm{d}p}_{\text{sat}}$$

1 bar, $\Delta T = 3^{\circ}C$, $R_{\text{eq}} = 5,2\ \mu\text{m}$, 155 bar, $\Delta T = 3^{\circ}C$, $R_{\text{eq}} = 0,08\ \mu\text{m}$

NUCLEATE BOILING MECHANISMS



- Super-heated liquid transport, Yagumata et al. (1955)

$$q \propto (T_P - T_{\text{sat}})^{1.2} n^{0.33}$$

- n: active sites number density,

$$n \propto \Delta T_{\text{sat}}^{5 \div 6} \Rightarrow q \propto \Delta T_{\text{sat}}^{3}$$

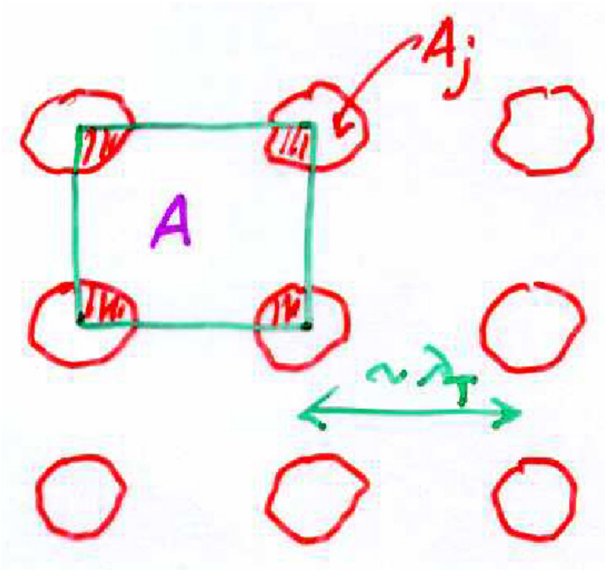
- Very hight heat transfer, precision unnecessary.

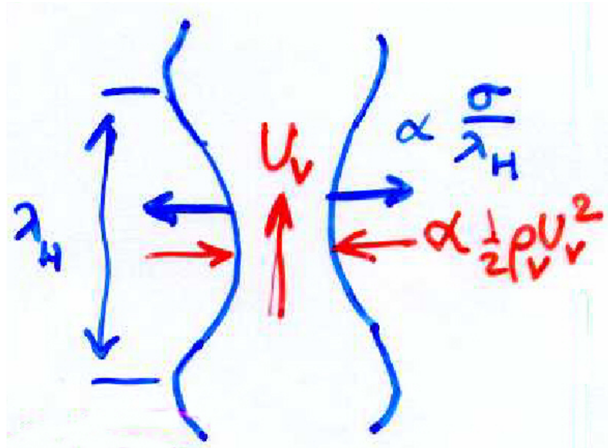
- Rohsenow (1952), analogy with convective h. t.: $\text{Nu} = C\text{Re}^a\text{Pr}^b$,
- Scales : $\text{Re} = \dfrac{\rho_L V L}{\mu_L}$,
  - Length: detachment diameter, capillary length: $L \approx \sqrt{\frac{\sigma}{g(\rho_L - \rho_V)}}$
  - Liquid velocity: energy balance, $q = \dot{m} h_{LV}$, $V \approx \frac{q}{\rho_L h_{LV}}$

$$\text{Ja} \triangleq \frac{C_{pL}(T_P - T_{\text{sat}})}{h_{LV}} = C_{sf}\text{Re}^{0.33}\text{Pr}_L^s$$

- $C_{sf} \approx 0.013$, $s = 1$ water, $s = 1.7$ other fluids.

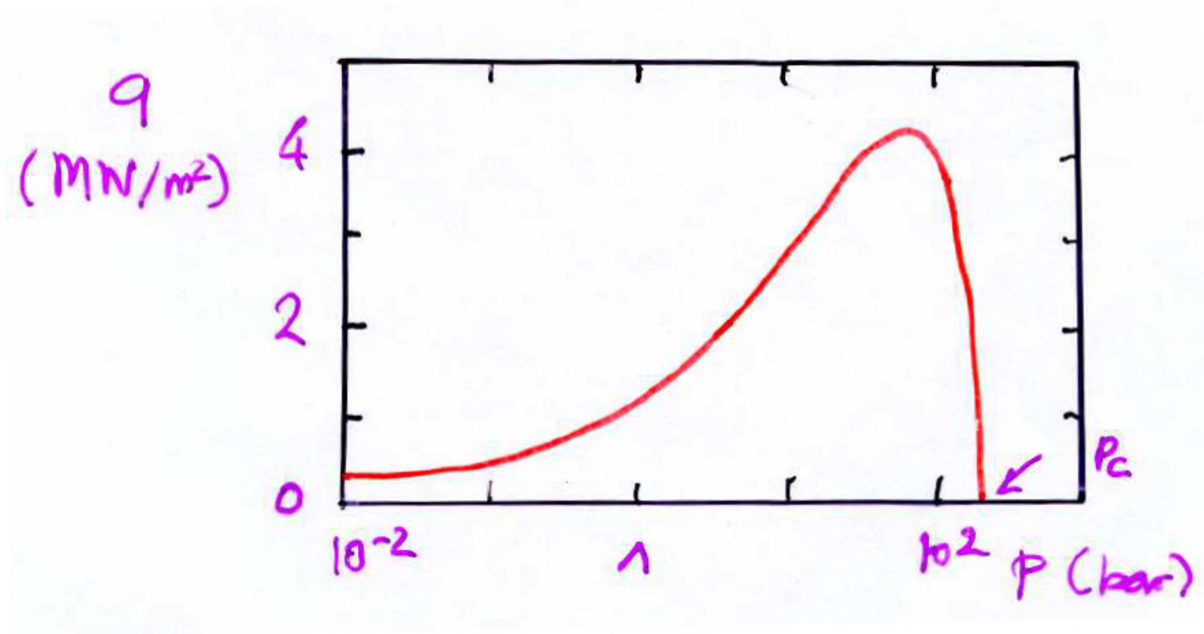
BOILING CRISIS, CRITICAL HEAT FLUX





- Flow pattern close to CHF: *critical heat flux*), Rayleigh-Taylor instability,
- Stability of the vapor column: Kelvin-Helmholtz,
- Energy balance over A,

$$\lambda_T = 2\pi\sqrt{3}\sqrt{\frac{\sigma}{g(\rho_L - \rho_V)}}, \quad \frac{1}{2}\rho_V U_V^2 < \pi\frac{\sigma}{\lambda_H}, \quad qA = \rho_V U_V A_J h_{LV}$$



- Zuber (1958), jet radius $R_J = \frac{1}{4}\lambda_T$, $\lambda_H = 2\pi R_J$, marginal stability,

$$q_{CHF} = 0.12\rho_V^{1/2} h_{LV} \sqrt[4]{\sigma g(\rho_V - \rho_L)}$$

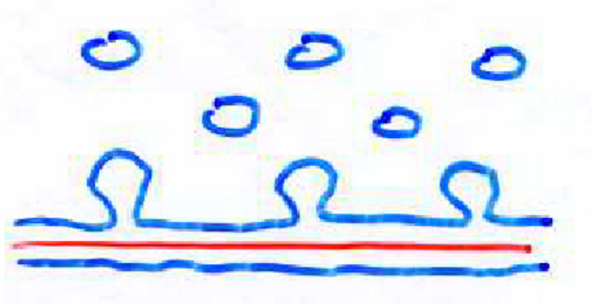
- Lienhard & Dhir (1973), jet radius $R_J = \frac{1}{4}\lambda_T$, $\lambda_H = \lambda_T$,

$$q_{CHF} = 0.15\rho_V^{1/2} h_{LV} \sqrt[4]{\sigma g(\rho_V - \rho_L)}$$

- Kutateladze (1948), dimensional analysis and experiments,

$$q_{CHF} = 0.13\rho_V^{1/2} h_{LV} \sqrt[4]{\sigma g(\rho_V - \rho_L)}$$

FILM BOILING



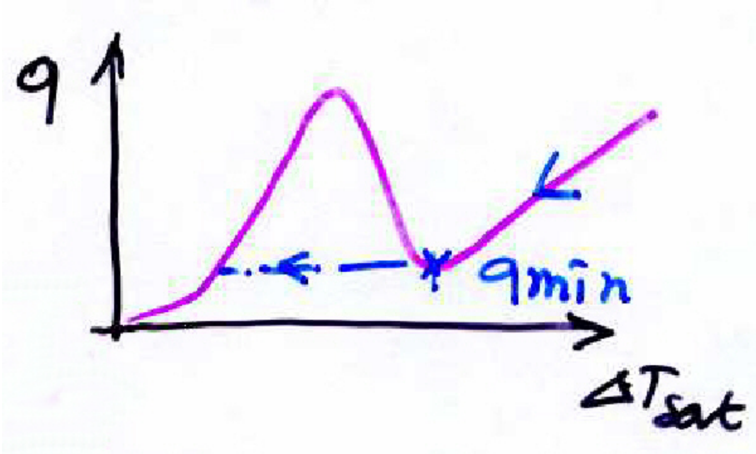
- Analogy with condensation (Nusselt, Rohsenow), Bromley (1950), $V \leftrightarrows L$

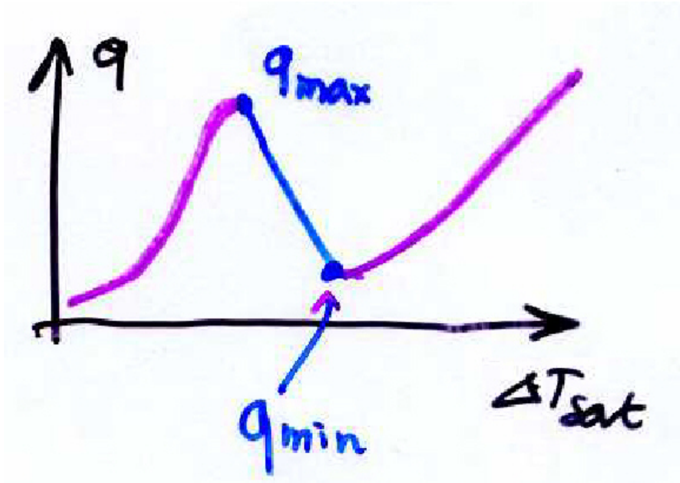
$$\mathrm{Nu}_L = 0.62 \left(\frac{\rho_V g(\rho_L - \rho_V) h'_{LV} D^3}{\mu_V k_V (T_W - T_{\mathrm{sat}})} \right)^{\frac{1}{4}}, \quad h'_{LV} = h_{LV} \left(1 + 0.34 \frac{C_{PV}(T_W - T_{\mathrm{sat}})}{h_{LV}} \right)$$

- Transport and thermodynamical properties:
  - Liquid at saturation T_{sat},
  - Vapor at the film temperature, $T_F = \frac{1}{2}(T_{\mathrm{sat}} + T_W)$.
- Radiation correction: $T_W > 300°\mathrm{C}$, ϵ : emissivity, $\sigma = 5,67\,10^{-8}$ W/m^2/K^4

$$h = h(T < 300°C) + \frac{\epsilon\sigma(T_W^4 - T_{\mathrm{sat}}^4)}{T_W - T_{\mathrm{sat}}}$$

TRANSITION BOILING





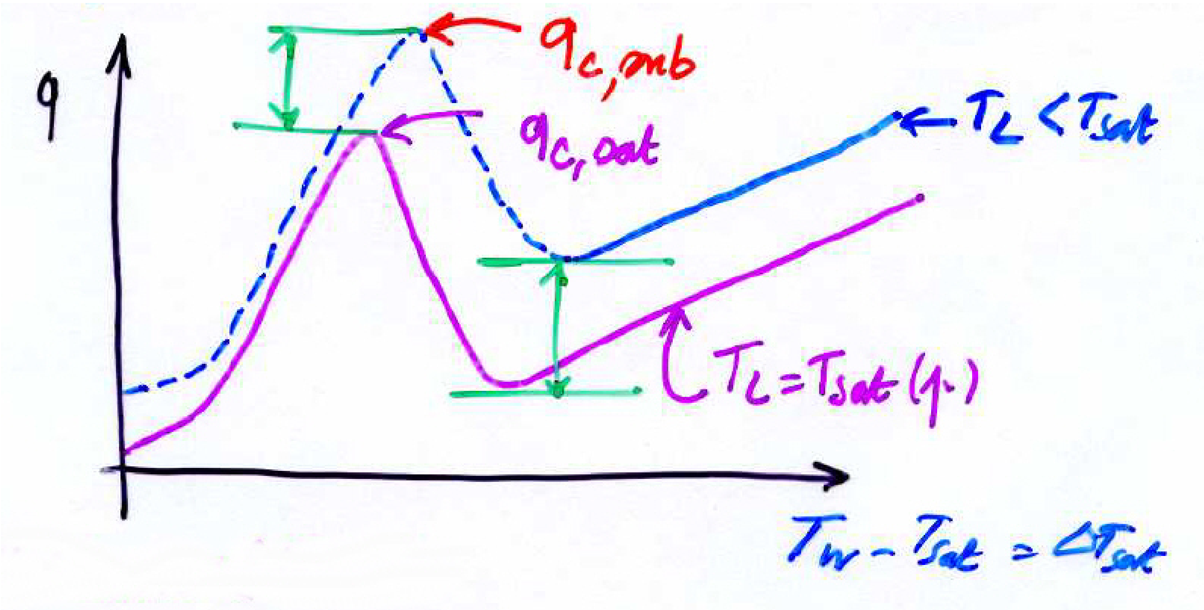
- Minimum flux,

$$q_{\min} = C h_{LV} \sqrt[4]{\frac{\sigma g(\rho_L - \rho_V)}{(\rho_L + \rho_V)^2}}$$

  - Zuber (1959), $C = 0.13$, stability of film boiling,
  - Berenson (1960), $C = 0,09$, rewetting, Liendenfrost temperature.

- Scarce data in transition boiling,
- Quick fix, $\Delta T_{\min}$ and $\Delta T_{\max}$, from each neighboring regime (NB and FB),
- Linear evolution in between (log-log plot!).

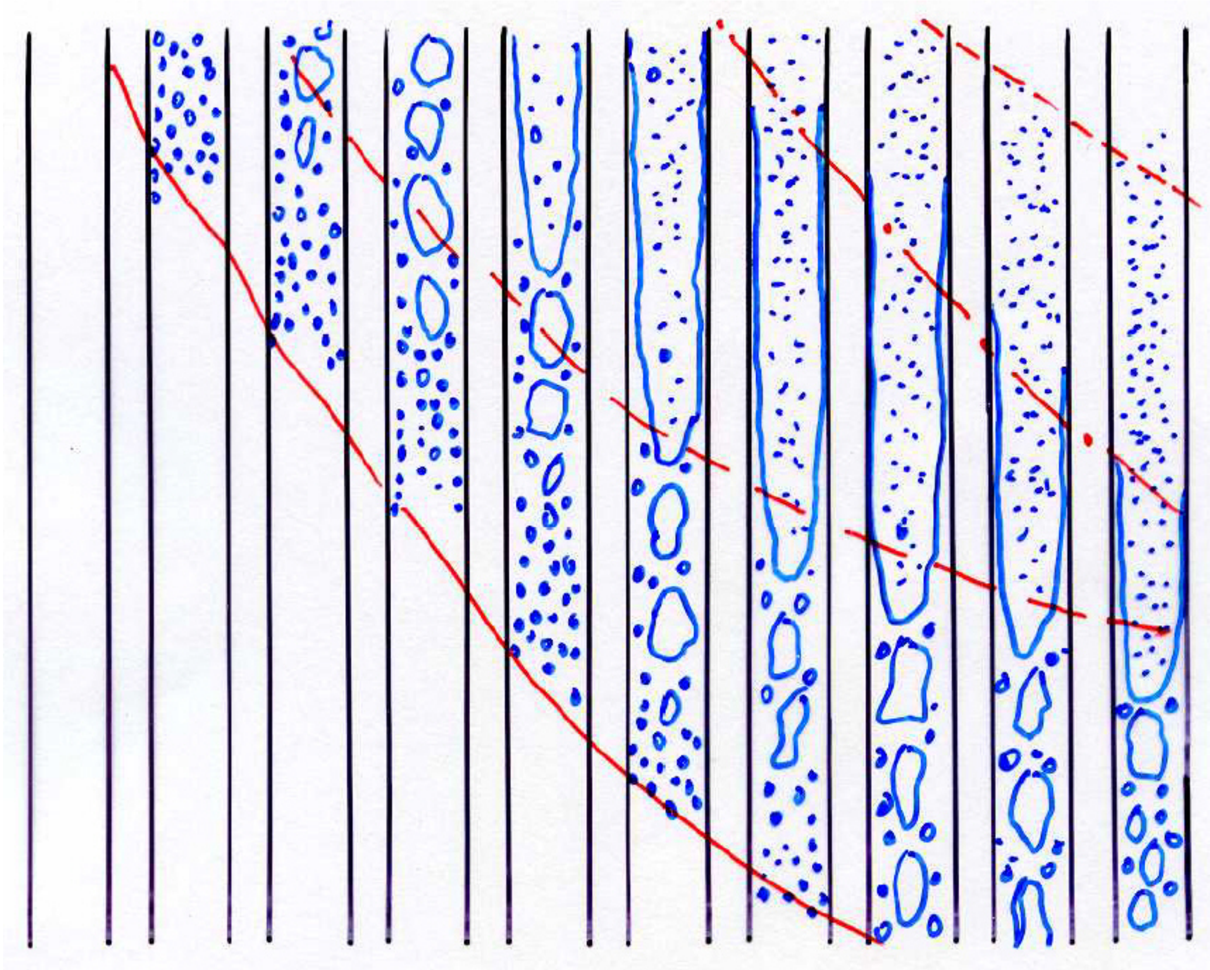
SUB-COOLING EFFECT



- Liquid sub-cooling, $T_L < T_{\text{sat}}$, $\Delta T_{\text{sub}} \triangleq T_{\text{sat}} - T_L$
- Ivey & Morris (1961)

$$q_{C,\text{sub}} = q_{C,\text{sat}} \left(1 + 0,1 \left(\frac{\rho_L}{\rho_V} \right)^{3/4} \frac{C_{PL} \Delta T_{\text{sub}}}{h_{LV}} \right)$$

CONVECTIVE BOILING REGIMES



→ Increasing heat flux, constant flow rate →

1. Onset of nucleate boiling
2. Nucleate boiling suppression
3. Liquid film dry-out
4. Super-heated vapor

BACK TO THE EQUILIBRIUM (STEAM) QUALITY

- Regime boundaries depend very much on z. Change of variable, x_{eq}
- Equilibrium quality, non dimensional *mixture enthalpy*,

$$x_{\mathrm{eq}} \triangleq \frac{h - h_{L\mathrm{sat}}}{h_{LV}}$$

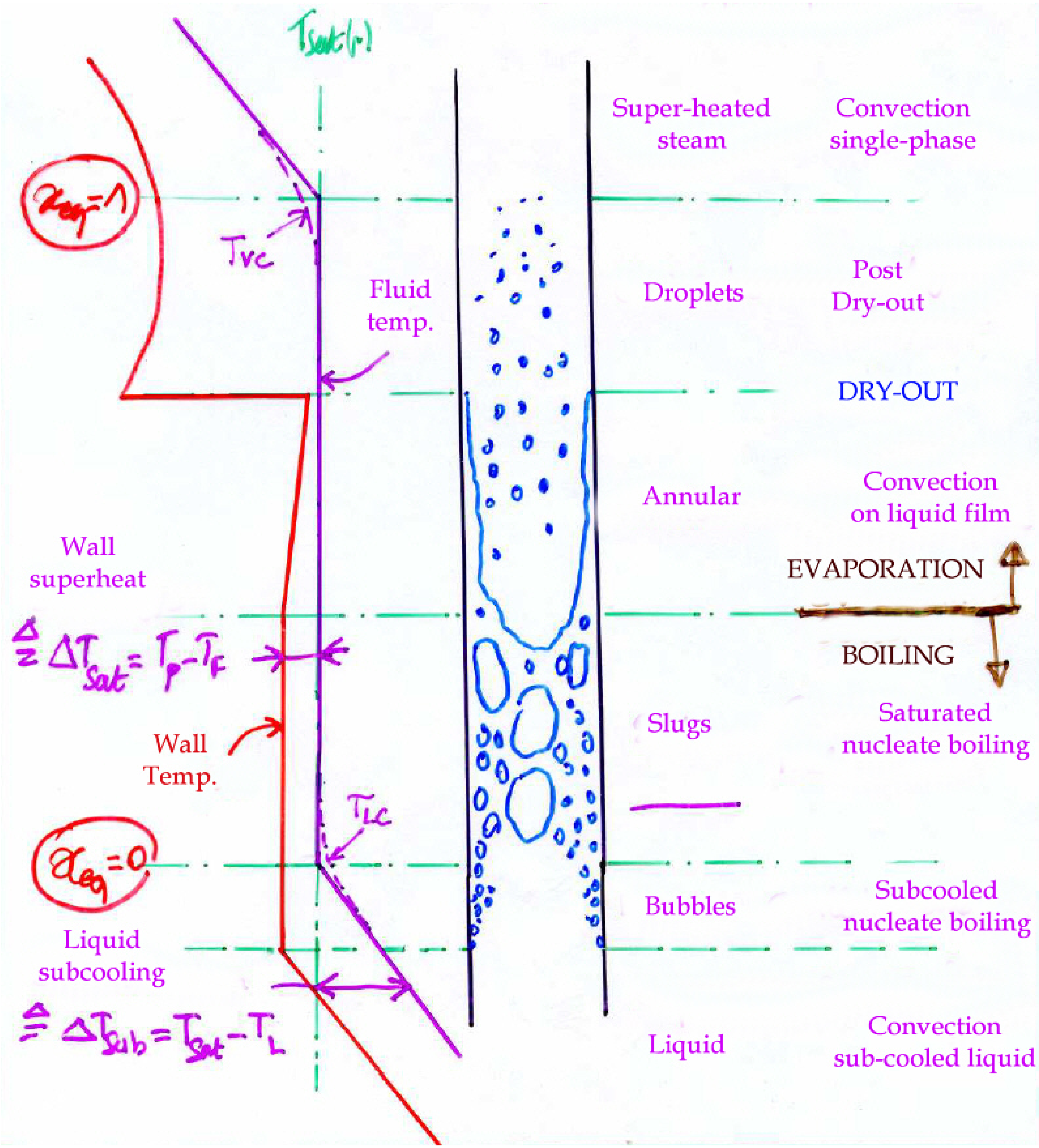
- Energy balance, low velocity, stationary flows,

$$M\frac{\mathrm{d}h}{\mathrm{d}z} = M h_{LV}\frac{\mathrm{d}x_{\mathrm{eq}}}{\mathrm{d}z} = q\mathcal{P}$$

- Uniform heat flux, x_{eq} linear in z. Close to equilibrium, $x_{\mathrm{eq}} \approx x$
- According to the assumptions of the HEM,

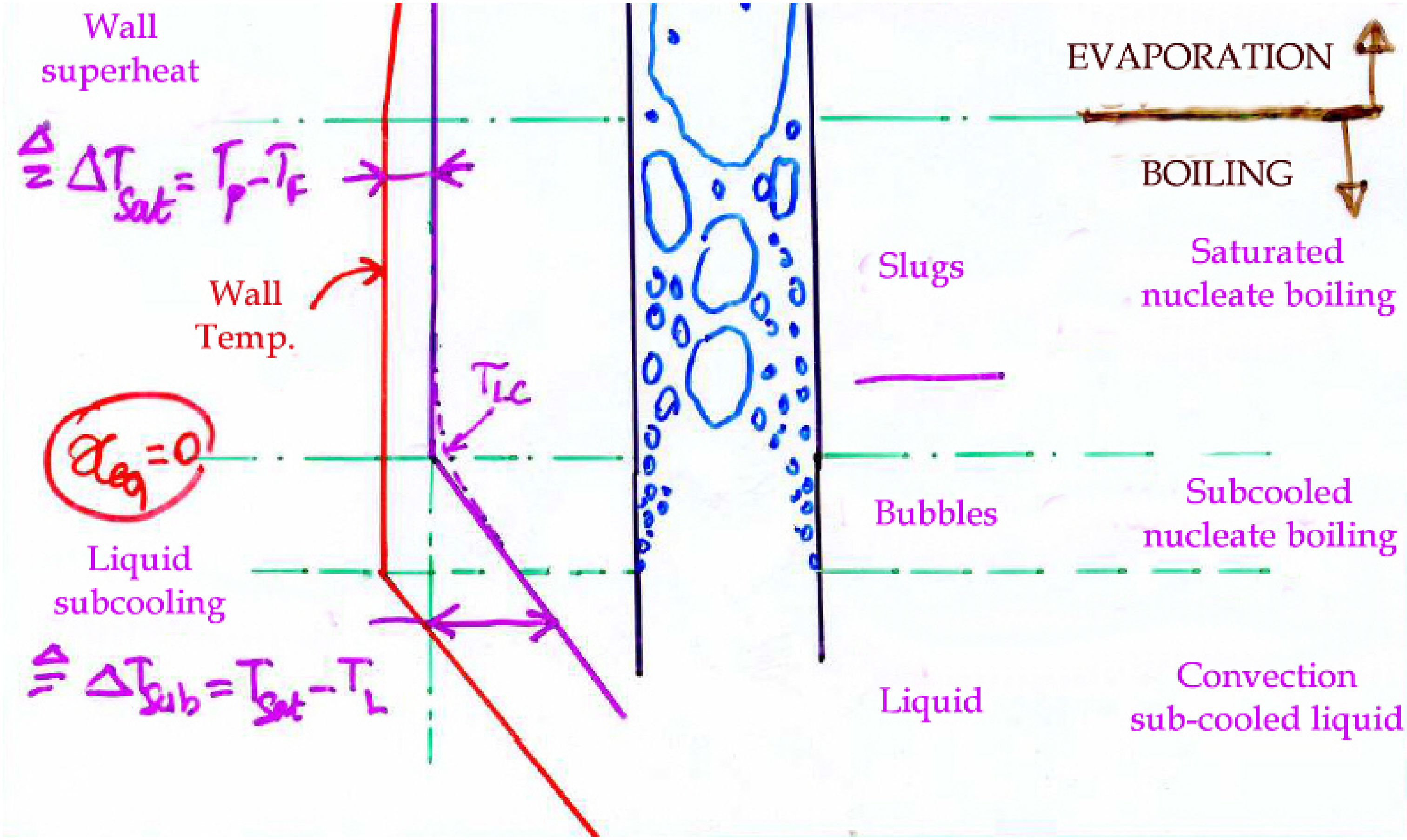
$0 > x_{\mathrm{eq}}$	single-phase liquid (sub-cooled)
$0 < x_{\mathrm{eq}} < 1$	two-phase, saturated
$1 < x_{\mathrm{eq}}$	single-phase vapor (super-heated)

CONVECTIVE HEAT TRANSFER IN VERTICAL FLOWS



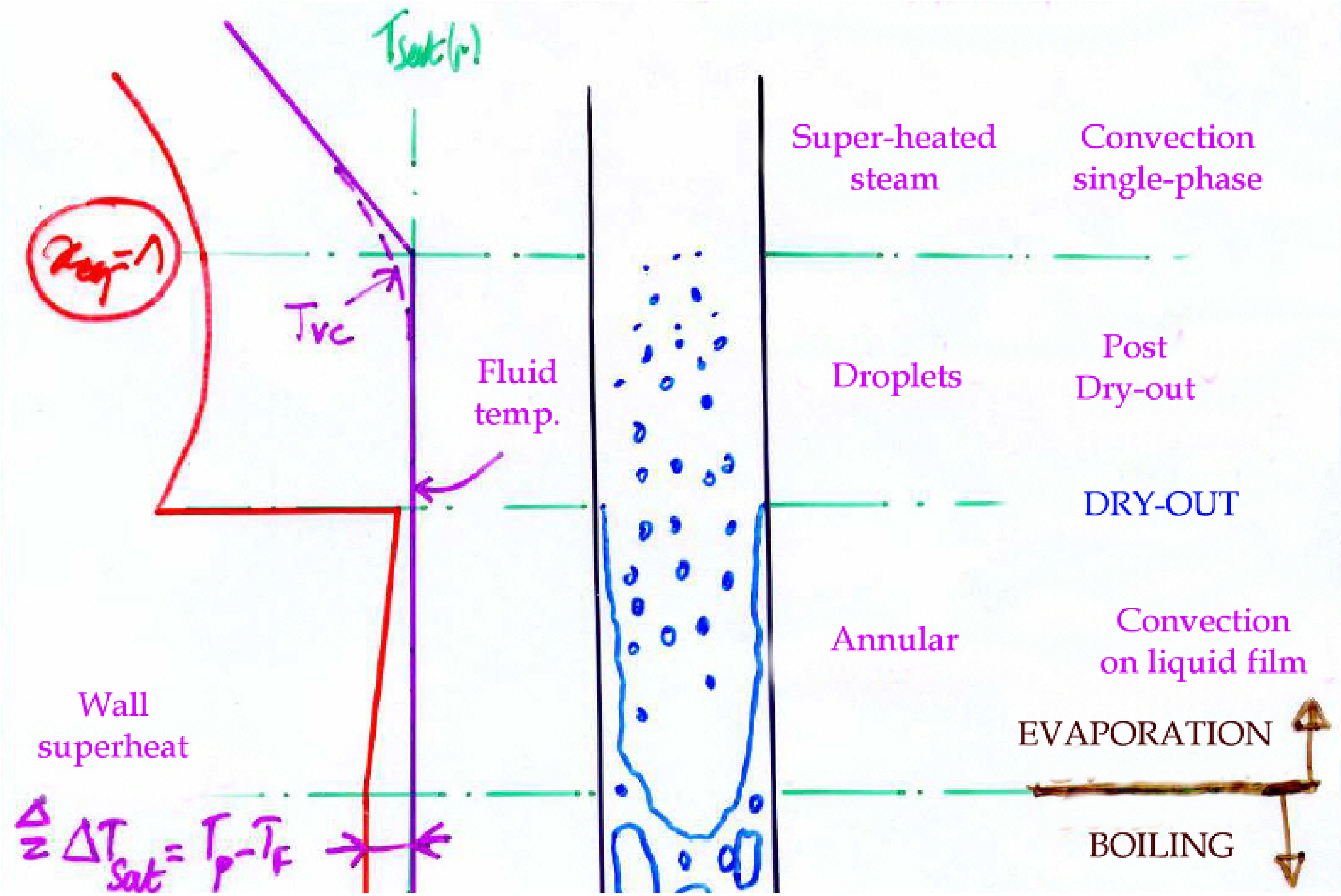
Boiling flow description

- Constant heat flux heating,
- Fluid temperature evolution, (T_{sat}),
- Wall temperature measurement,
- Flow regime,
- Heat transfer controlling mechanism.



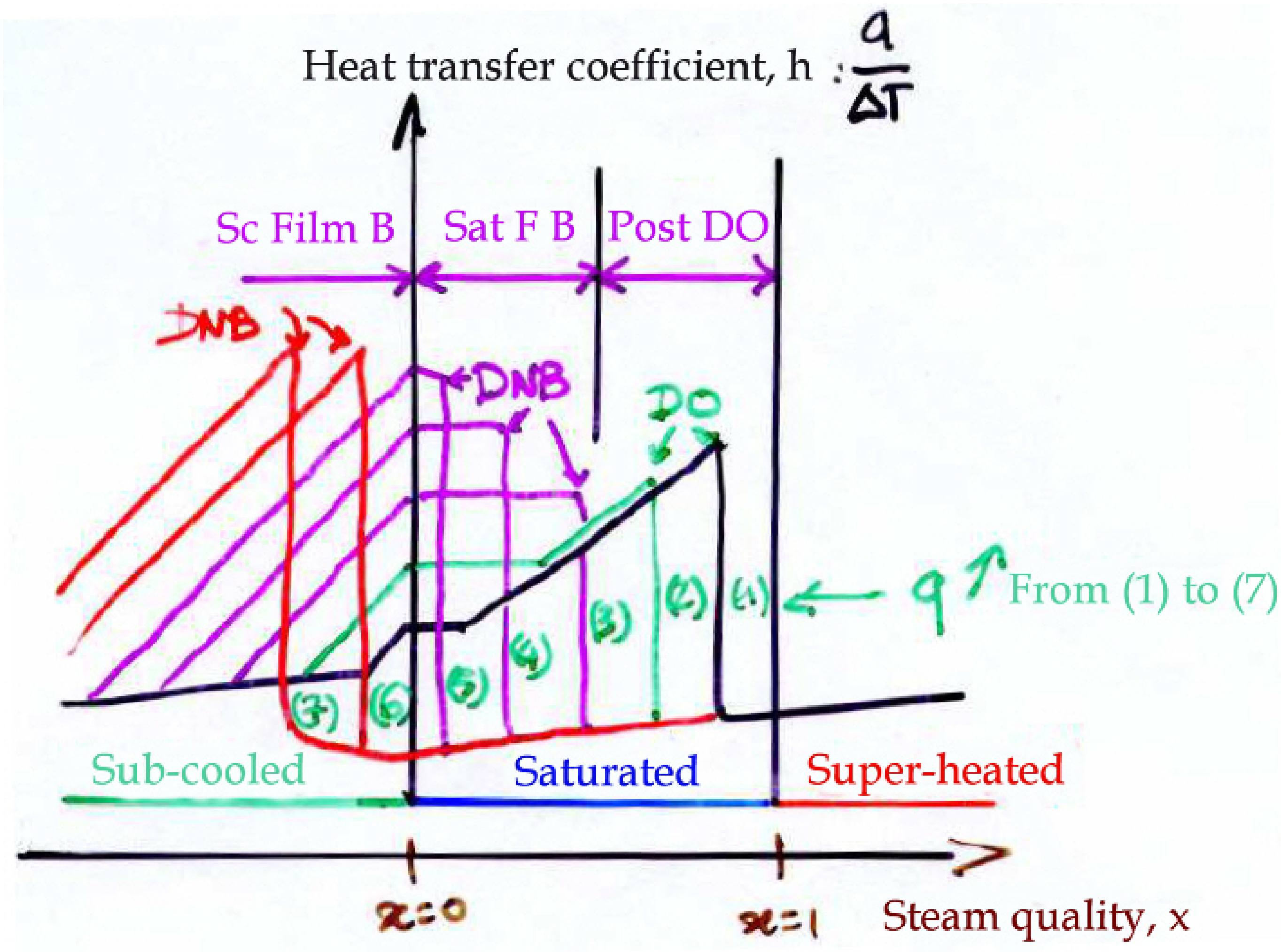
From the inlet, flow and heat transfer regimes,

- Single-phase convection
- Onset of nucleate boiling, ONB
- Onset of signifiant void, OSV
- Important points for pressure drop calculations, flow oscillations.



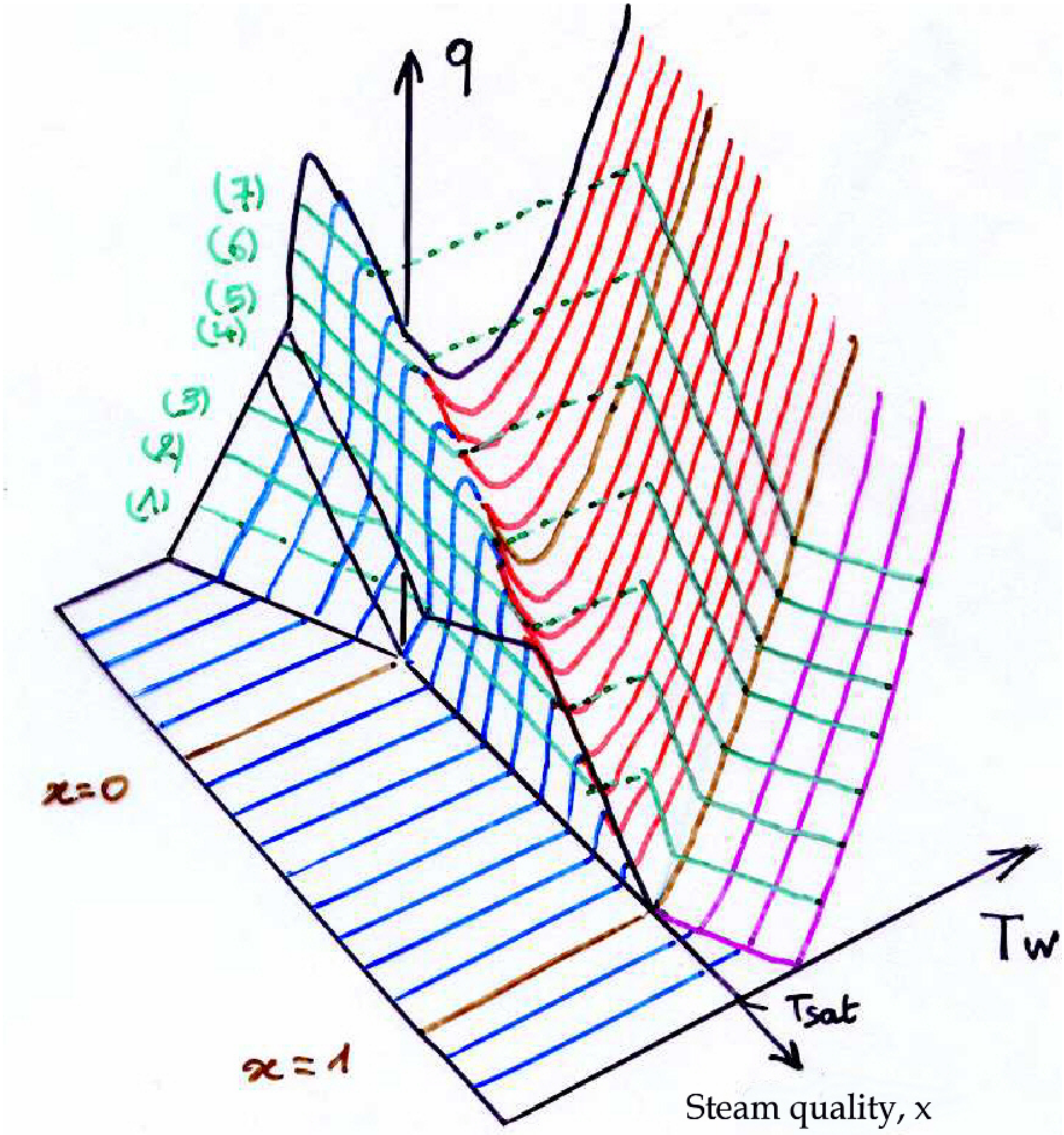
- Nucleate boiling suppression,
- Liquid film dry-out, boiling crisis (I),
- Single-phase vapor convection.

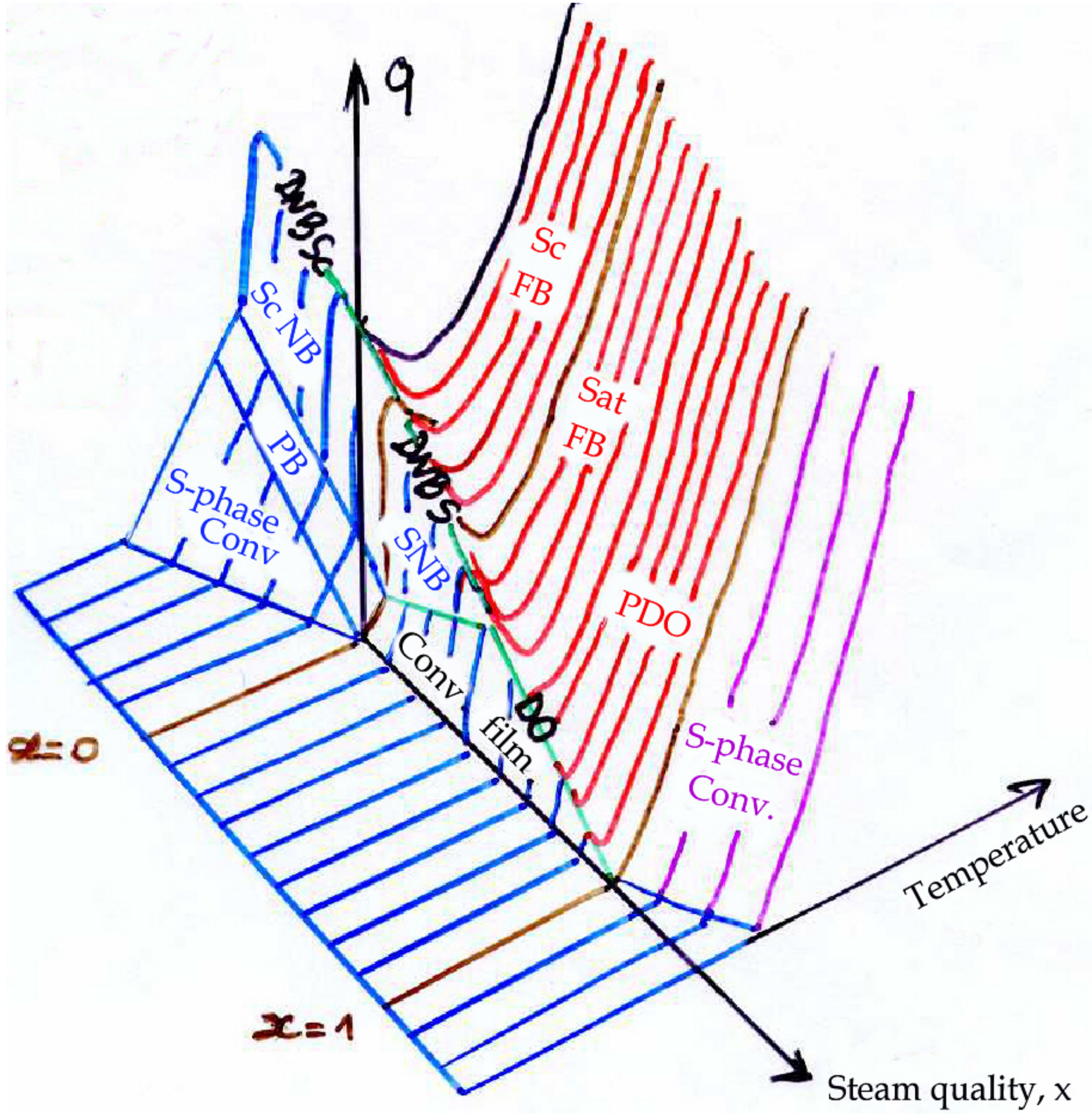
HEAT TRANSFER COEFFICIENT



DO: dry-out, DNB: departure from nucleate boiling (saturated, sub-cooled), PDO: post dry-out, sat FB: saturated film boiling, Sc Film B: sub-cooled film boiling

BOILING SURFACE





S-Phase conv: single-phase convection, PB: partial boiling, NB: nucleate boiling (S, saturated, Sc, subcooled), FB: film boiling, PDO: post dry-out, DO: dry-out, DNB: departure from nucleate boiling.

SINGLE-PHASE FORCED CONVECTION

- Forced convection (Dittus & Boelter, Colburn), $\text{Re} > 10^4$,

$$\text{Nu} \triangleq \frac{hD}{k_L} = 0,023\text{Re}^{0,8}\text{Pr}^{0,4}, \quad \text{Re} = \frac{GD}{\mu_L}, \quad \text{Pr}_L = \frac{\mu_L C_{PL}}{k_L}$$

- Fluid temperature, T_F, *mixing cup* temperature, that corresponding to the area-averaged mean enthalpy.
- Transport properties at T_{av}
  - Local heat transfer coefficient,

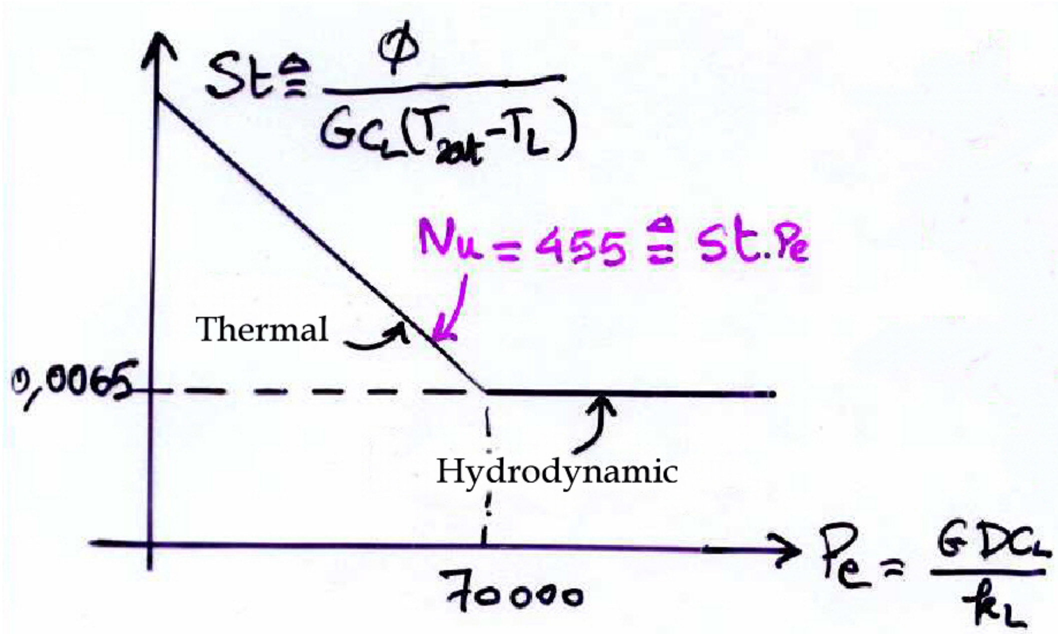
$$q \triangleq h(T_W - T_F), \quad T_{av} = \frac{1}{2}(T_W + T_F)$$

  - Averaged heat transfer coefficient (length L),

$$\bar{q} \triangleq \bar{h}(\bar{T}_W - \bar{T}_F), \quad \bar{T}_F = \frac{1}{2}(T_{Fin} + T_{Fout}), \qquad T_{av} = \frac{1}{2}(\bar{T}_W + \bar{T}_F)$$

- Always check the original papers...

NUCLEATE BOILING & SIGNIFICANT VOID



- Onset and suppression of nucleate boiling, ONB, (Frost & Dzakowic, 1967),

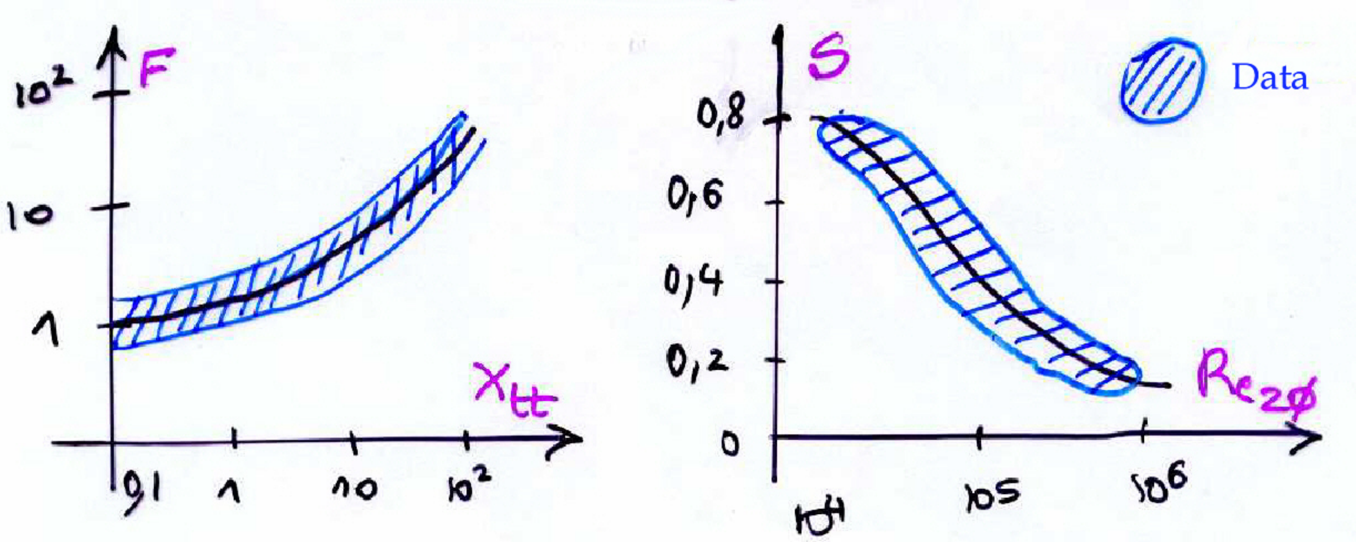
$$T_P - T_{\text{sat}} = \left(\frac{8 \sigma q T_{\text{sat}}}{k_L \rho_V h_{LV}} \right)^{0,5} \text{Pr}_L$$

- Onset of signifiant void, OSV, (Saha & Zuber, 1974)

$$\text{Nu} = \frac{qD}{k_L (T_{\text{sat}} - T_L)} = 455, \quad \text{Pé} < 7\,10^4, \text{ thermal regime}$$

$$\text{St} = \frac{q}{GC_{PL}(T_{\text{sat}} - T_L)} = 0,0065, \quad \text{Pé} > 7\,10^4, \text{ hydrodynamic regime}$$

DEVELOPPED BOILING AND CONVECTION



- Weighting of two mechanisms, $x_{\text{eq}} > 0$ (Chen, 1966)
  - Nucleate boiling(Forster & Zuber, 1955), S, suppression factor,same model for pool boiling,
  - Forced convection, Dittus Boelter, F, amplification factor,

$$h = h_{\text{FZ}} S + h_{\text{DB}} A$$

$$\frac{1}{S} = 1 + 2.53\,10^{-6} (\text{Re} F^{1.25})^{1.17}, \quad F = \begin{cases} 1 & 1/X \leqslant 0.1 \\ 2.35(1/X + 0.213)^{0.736} & 1/X > 0.1 \end{cases}$$

CHEN CORRELATION (CT'D)

- Nucleate boiling,

$$h_{FZ} = 0.00122 \frac{k_L^{0.79} \, C_{pL}^{0.45} \, \rho_L^{0.49}}{\sigma \mu_L^{0.29} \, h_{LV}^{0.24} \, \rho_V^{0.24}} (T_W - T_{\text{sat}})^{0.24} \Delta p_{\text{sat}}^{0.75}$$

- Forced convection

$$h_{DB} = 0.023 \, \frac{k_L}{D} \text{Re}^{0.8} \text{Pr}_L^{0.4}$$

- From Clapeyron relation, slope of saturation line,

$$\Delta p_{\text{sat}} = \frac{h_{LV}(T_W - T_{\text{sat}})}{T_{\text{sat}}(v_V - v_L)}$$

- Non dimensional numbers definitions,

$$\text{Re} = \frac{GD(1 - x_{\text{eq}})}{\mu_L}, \quad X = \left(\frac{1 - x_{\text{eq}}}{x_{\text{eq}}}\right)^{0.9} \left(\frac{\rho_V}{\rho_L}\right)^{0.5} \left(\frac{\mu_L}{\mu_V}\right)^{0.1}, \quad \text{Pr}_L = \frac{\mu_L C_{pL}}{k_L}$$

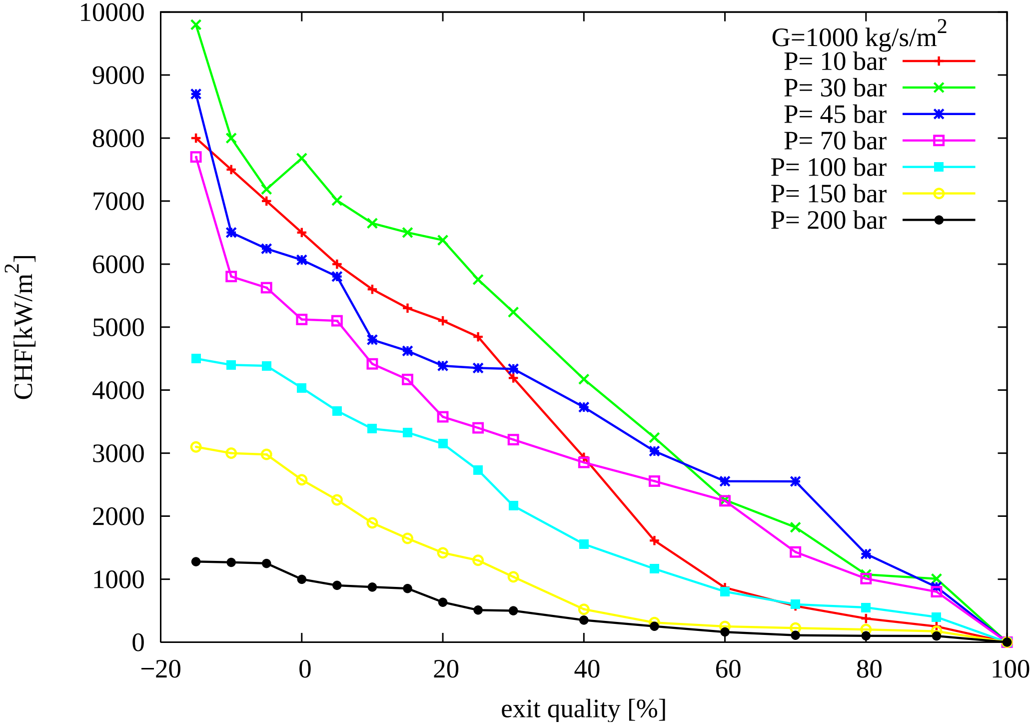
- NB: implicit in $(T_W - T_{\text{sat}})$.

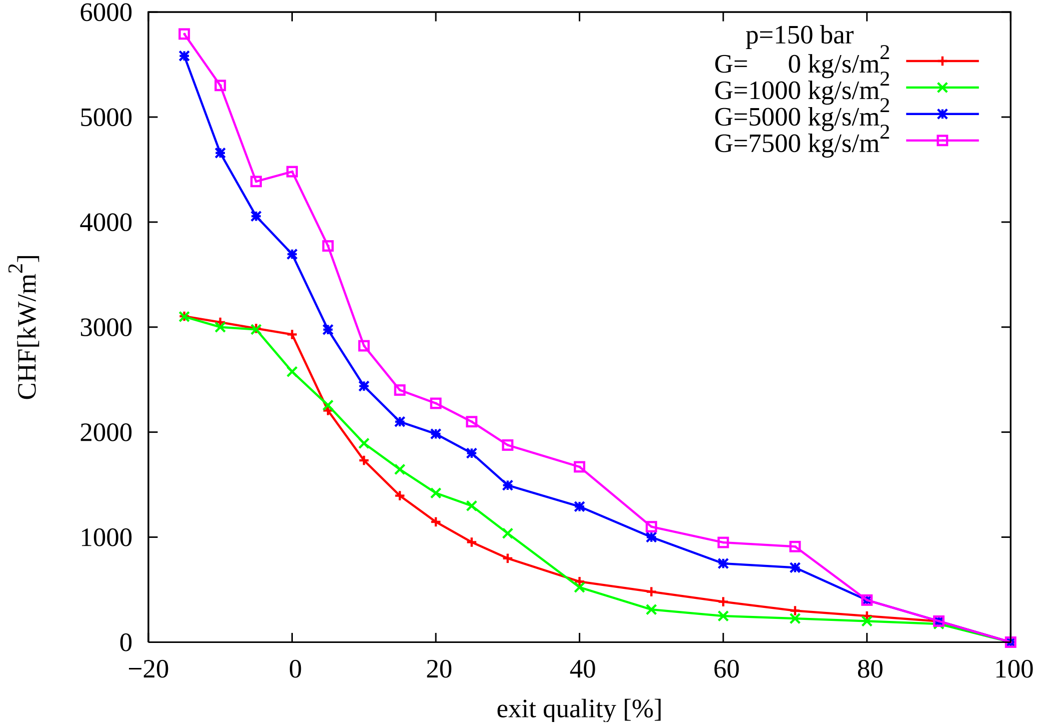
CRITICAL HEAT FLUX

- No general model.
  - Dry-out, multi-field modeling
  - DNB, correlations or experiment in real bundles
- Very sensitive to geometry, mixing grids,
- Recourse to experiment is compulsory,
- In general, $q_{\text{CHF}}(p, G, L, \Delta H_i, ...)$, artificial reduction of dispersion.
- For tubes and uniform heating, no length effect, $q_{\text{CHF}}(p, G, x_{\text{eq}})$
  - Tables by Groenveld,
  - Bowring (1972) correlation, best for water in tubes
  - Correlation by Katto & Ohno (1984), non dimensional, many fluids, regime identification.

MAIN PARAMETERS EFFECT ON CHF

After Groeneveld & Snoek (1986), tube diameter, $D = 8$ mm.





- Generally decreases with the increase of the exit quality. $q_{\mathrm{CHF}} \to 0$, $x_{\mathrm{eq}} \to 1$.
- Generally increases with the increase of the mass flux,
- CHF is non monotonic with pressure.

MORE ON HEAT TRANSFER

- Boiling and condensation,
  - Delhaye (1990)
  - Delhaye (2008)
  - Roshenow *et al.* (1998)
  - Collier & Thome (1994)
  - Groeneveld & Snoek (1986)
- Single-phase,
  - Bird *et al.* (2007)
  - Bejan (1993)

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Bird, R. B., Stewart, W. E., & Lightfoot, E. N. 2007. *Transport phenomena*. Revised second edn. John Wiley & Sons.

Collier, J. G., & Thome, J. R. 1994. *Convective boiling and condensation*. third edn. Oxford: Clarendon Press.

Delhaye, J. M. 1990. Transferts de chaleur : ebullition ou condensation des corps purs. *Techniques de l'ingénieur*.

Delhaye, J.-M. 2008. *Thermohydraulique des réacteurs nucléaires*. Collection génie atomique. EDP Sciences.

Groeneveld, D. C., & Snoek, C. V. 1986. *Multiphase Science and Technology*. Vol. 2. Hemisphere. G. F. Hewitt, J.-M. Delhaye, N. Zuber, Eds. Chap. 3: a comprehensive examination of heat transfer correlations suitable for reactor safety analysis, pages 181–274.

Raithby, G. D., & Hollands, K. G. 1998. *Handbook of heat transfer*. 3rd edn. McGraw-Hill. W. M. Roshenow, J. P. Hartnett and Y. I Cho, Eds. Chap. 4-Natural convection, pages 4.1–4.99.

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