Numerical investigation of cryogen re-gasification in a plate heat exchanger

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HR EXCELLENCE IN RESEARCH

Wrocław University of Science and Technology

GENERAL MOTIVATION

- Efficient re-gasification of cryogen is a crucial process in many cryogenic installations
- LNG evaporators for stationary and mobile applications (marine and land transport)
- Nitrogen or argon can be obtained at highest purity after re-gasification
- Advantages of Plate heat exchangers (PHE):
  - compact size (20-40% less space than equivalent tube HE)
  - very high heat transfer coefficients (high turbulence BUT low velocities)
  - lower costs (reduced flow rates, smaller pumps, multiple duties in single unit)

- Fast and reliable numerical model is crucial for PHE optimization:
  - geometrical optimization (heat transfer area)
  - heat transfer optimization (solidification problem)
  - maldistribution of a flow in PHE channels

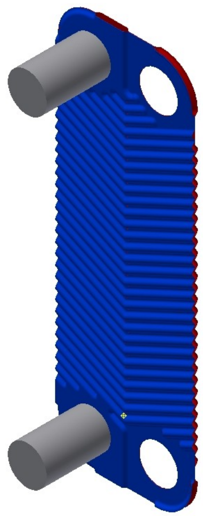
Poland – Taiwan scientific cooperation:

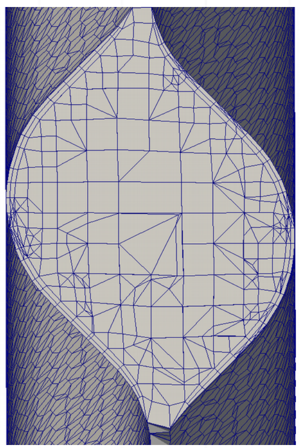
Development of plate heat exchangers for liquid inert gas vaporization and the modeling of the two-phase flow in heat exchangers (PHEVAP)

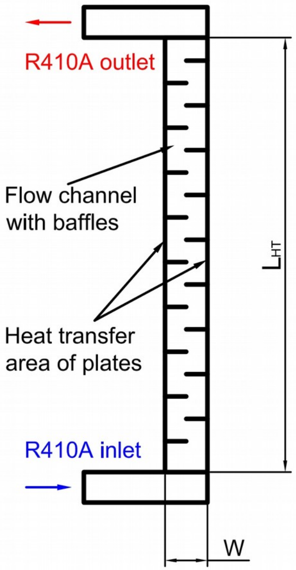
financed by The National Center for Research and Development, Poland

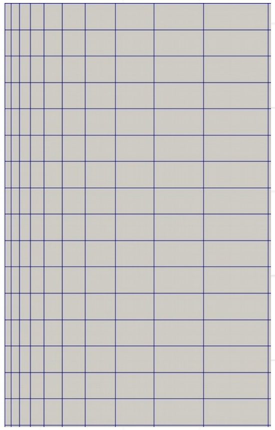
MOTIVATION FOR SIMPLIFIED NUMERICAL MODEL

- Phase change modeling requires very fine numerical mesh and time steps
- 3D single plate model requires **O(10^6)** computational cells and **O(10^{-6})** time step
- PHE has typically **O(100)** plates: **O(10^8)** computational nodes !









3D real geometry:
- chevron shape – complex channels
- 4.5×10^6 of computational cells
- 1 second ~ O(days)

2D simplified geometry:
- 5×10^4 of computational cells
- 1 second ~ O(hours)
- mixing baffles to enhance turbulence

PHASE CHANGE MATHEMATICAL MODEL

- **Lee** two-phase model for evaporation and condensation
- Phase change occurs in constant pressure and a quasi-thermo equilibrium state
- Mass transfer mainly depends on saturation temperature
- Interfacial temperature between phases is equal to saturation temperature
- Rates of mass exchange between liquid and vapor added to continuity equations

$$\frac{\partial \rho_l \alpha_l}{\partial t} + \nabla \cdot (\rho_l \boldsymbol{u_l} \alpha_l) = R_{cond} - R_{evap}$$

$$\frac{\partial \rho_v \alpha_v}{\partial t} + \nabla \cdot (\rho_v \boldsymbol{u_v} \alpha_v) = R_{evap} - R_{cond}$$

$$R_{cond} = C_c \rho_v \alpha_v (T_{sat} - T) \quad R_{evap} = 0 \quad T < T_{sat}$$

$$R_{evap} = C_e \rho_l \alpha_l (T - T_{sat}) \quad R_{cond} = 0 \quad T > T_{sat}$$

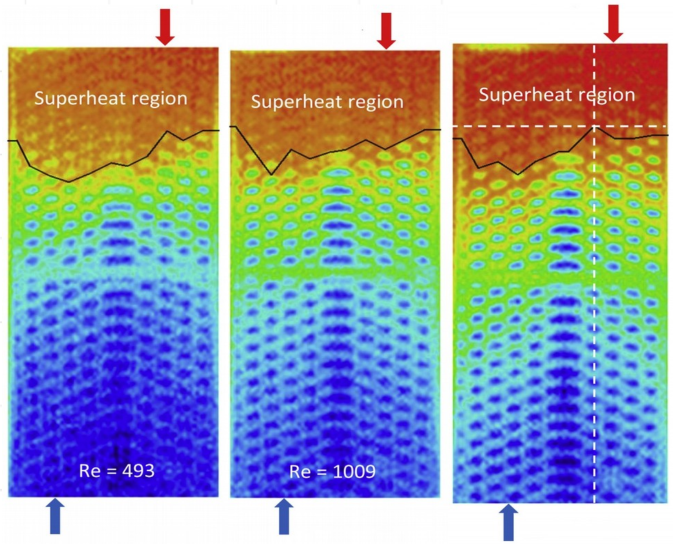
R_{cond}, R_{evap} – rates of mass exchange

Implemented as *interCondensatingEvaporatingFoam* in **OpenFOAM** software

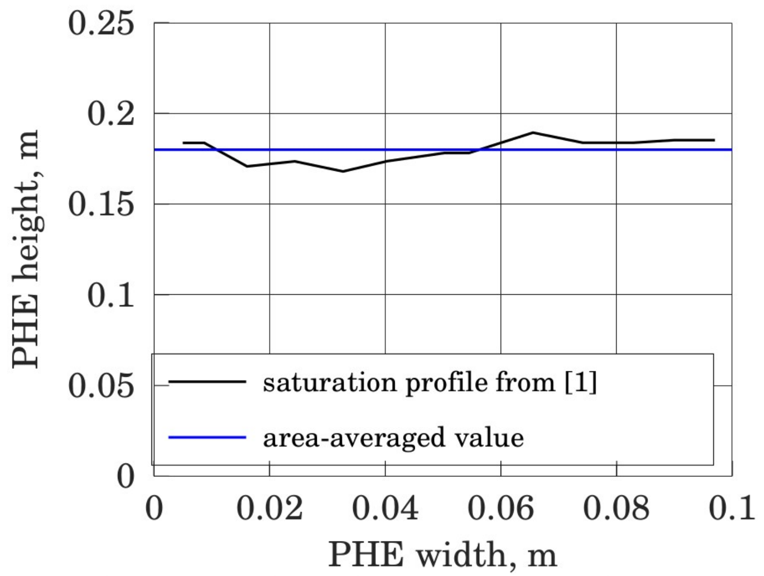
- No interface modeling – crucial savings of computational time

EXPERIMENT OF **Lin, Li and Yang 2014**

Re-gasification of R410A for different Re numbers (R410A mass flows)



Re = 500, 1000, 1500



Onset of superheat region: Re independent

Outlet superheat: $\Delta T = 10, \quad T_{out} = 11.1 C$

Working conditions:

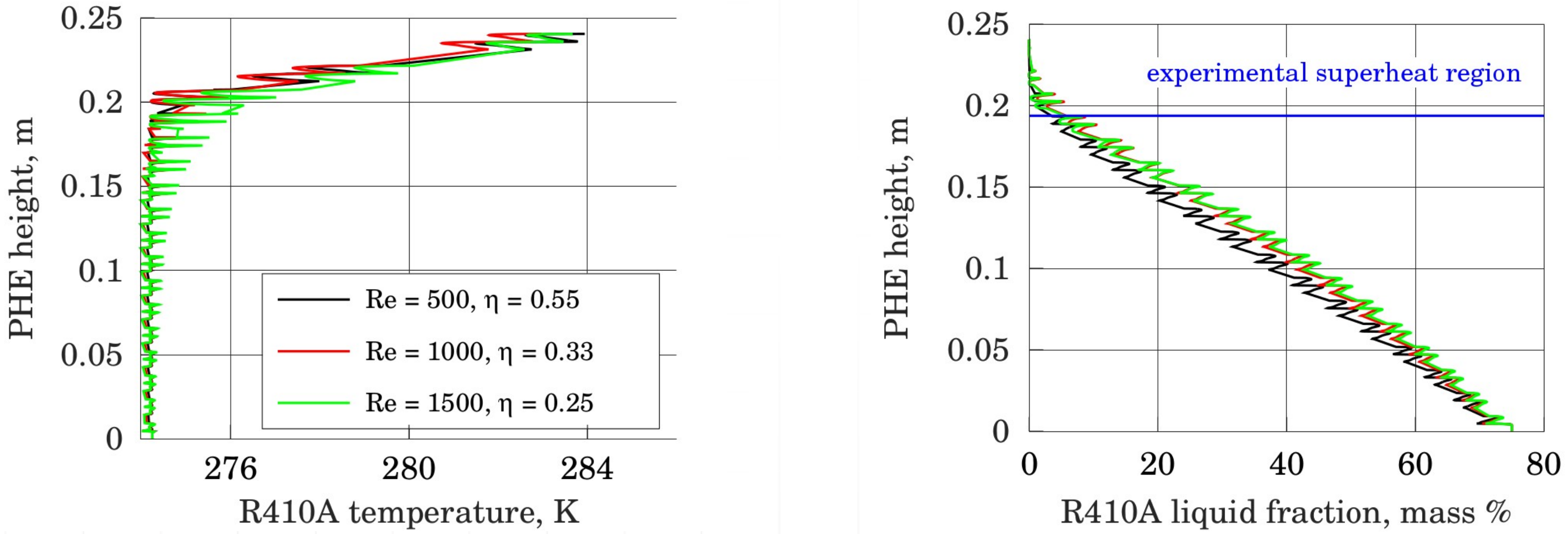
- $p = 8.2$ bar
- $T_{sat} = 1.1\ C$
- heating fluid mass flow adjusted for every Re

COMPARISON WITH EXPERIMENT – correction parameter

- *InterCondensatingEvaporatingFoam* model failed to compare with experiment
- Solution: introduction of a numerical correction parameter η :

$$h_{v\ num} = \eta(Re)\, h_{v\ real}$$

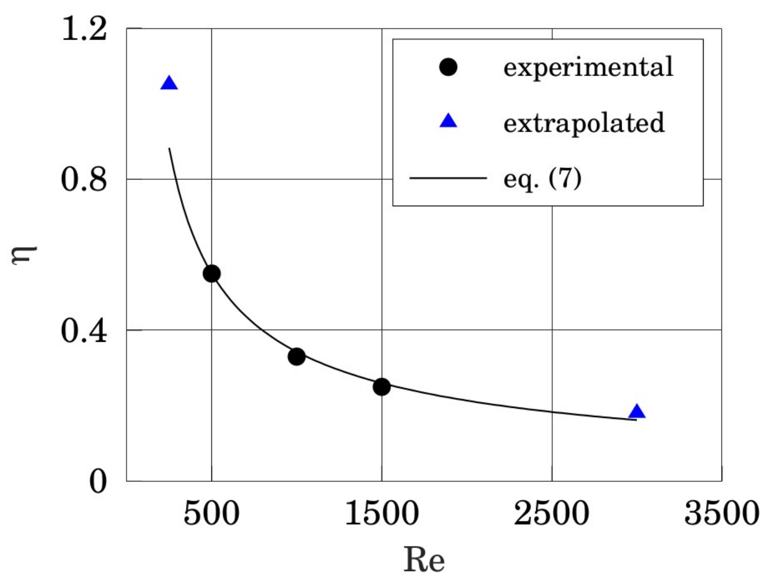
$h_{v\ num/real}$	numerical/real enthalpy of vaporization
$\eta(Re)$	numerical correction parameter



Outlet superheat: $\Delta T = T_{out} - T_{inlet} \sim 10^{o}$

SUCCESSFUL COMPARISON WITH EXPERIMENT

NUMERICAL CORRECTION PARAMETER



Re	η	T_{out}, C
250	1.05	11.31
500	0.55	12.01
1000	0.33	11.02
1500	0.25	11.56
3000	0.18	11.11

$$\eta(Re) = 38.237\, Re^{-0.683}$$

$$R^2 = 0.953$$

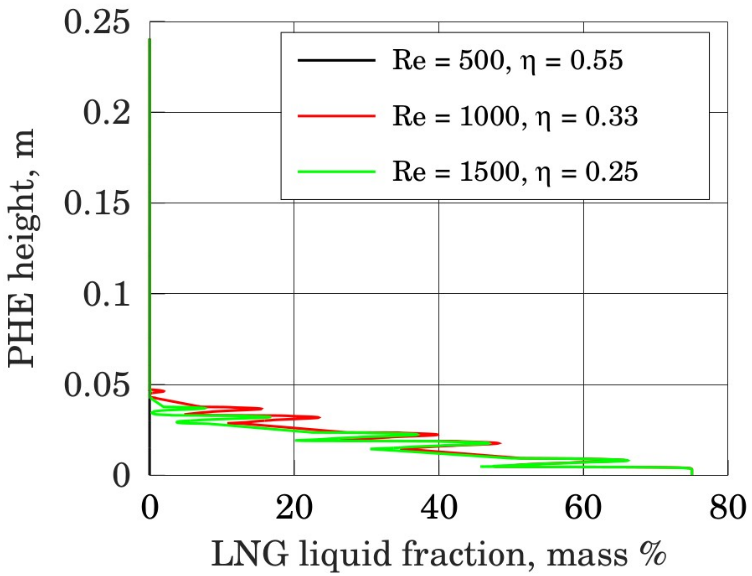
η less sensitive for larger Re

$\eta \approx 1$ for smaller Re

Mixing baffles:

- good for smaller Re – 2D flow has similar pattern as real 3D flow
- for higher Re flow gets 3D, additional/artificial turbulence enhancement is necessary

APPLICATION TO LNG RE-GASIFICATION



Re	η	T_{out}, C
500	0.55	18
1000	0.33	17
1500	0.25	16

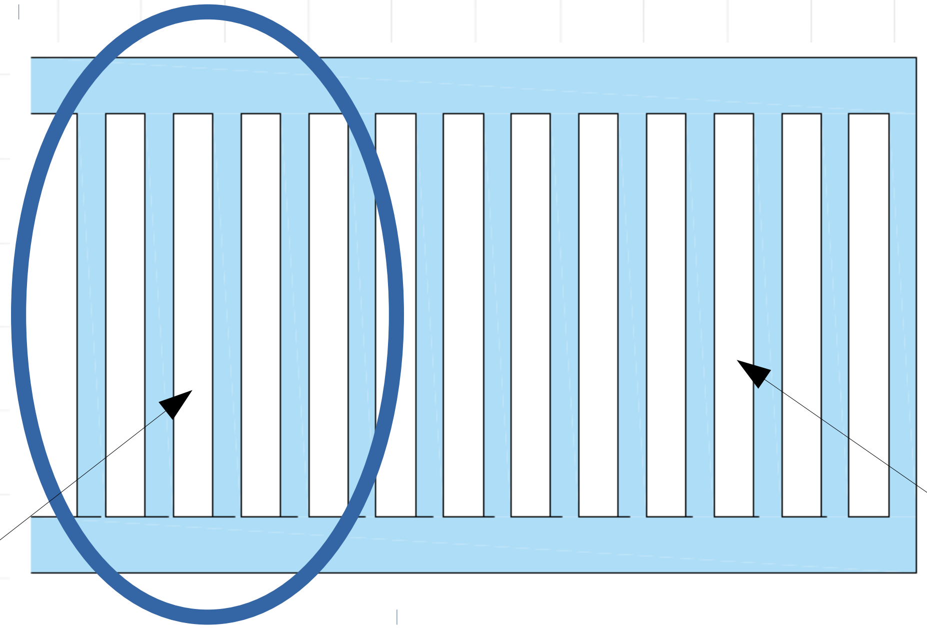
Working conditions:

- $p = 8.2$ bar
- $T_{sat} = -149$ C

Comments:

- LNG evaporated much faster for the same wall temperature profile as in R410A case
- Onset of superheated region is Re independent
- For the assumed water mass flow no solidification is expected
- T_{out} weakly dependent on η and Re

FLOW MALDISTRIBUTION IN PHE



ACTIVE Region (high velocity)

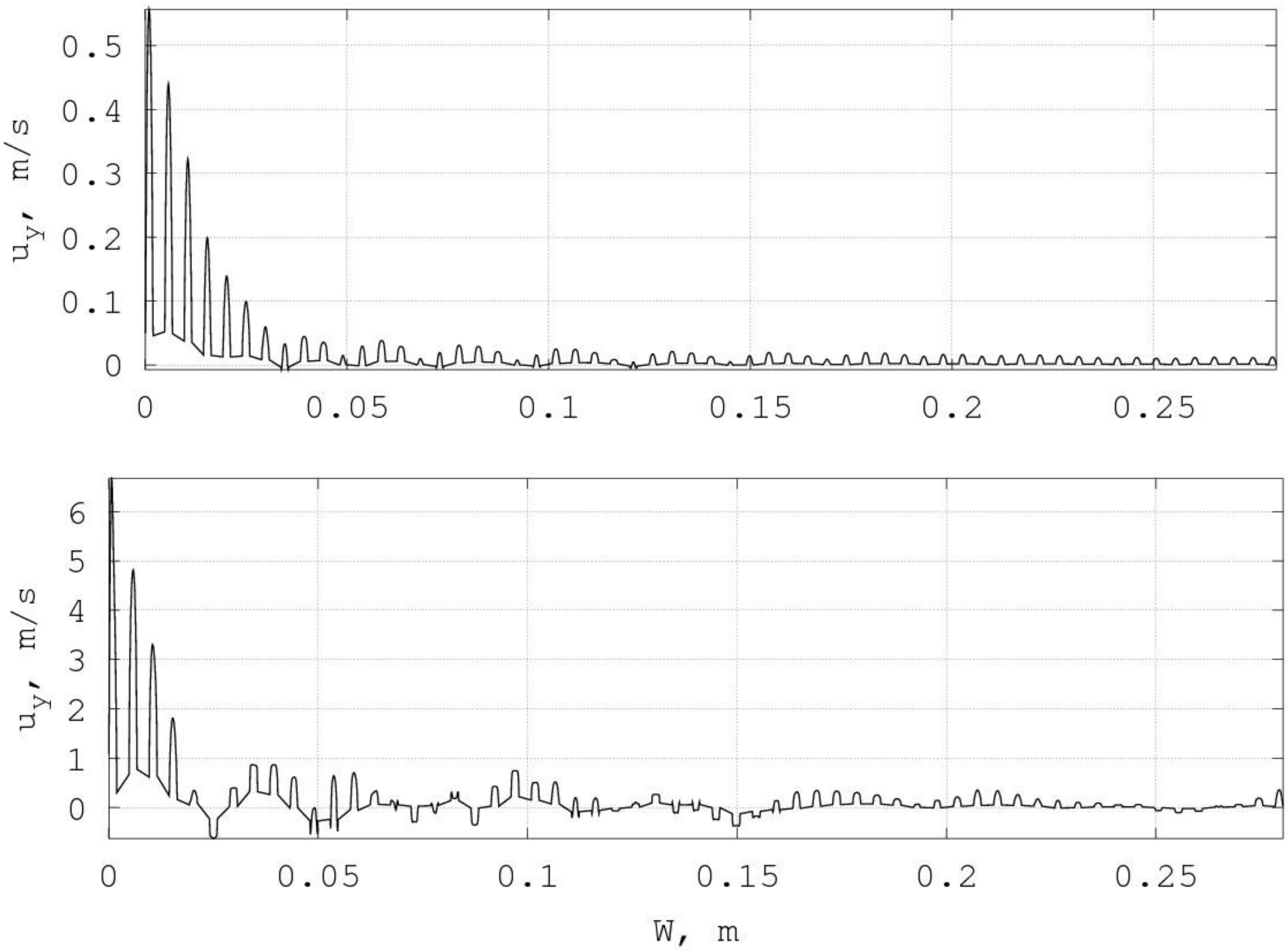
PASSIVE Region (low velocity)

Significant part of PHE can be NOT ACTIVE

Is it mostly a hydrodynamic problem ?

Is a phase change of primary or secondary importance ?

FLOW MALDISTRIBUTION IN PHE (ongoing research)



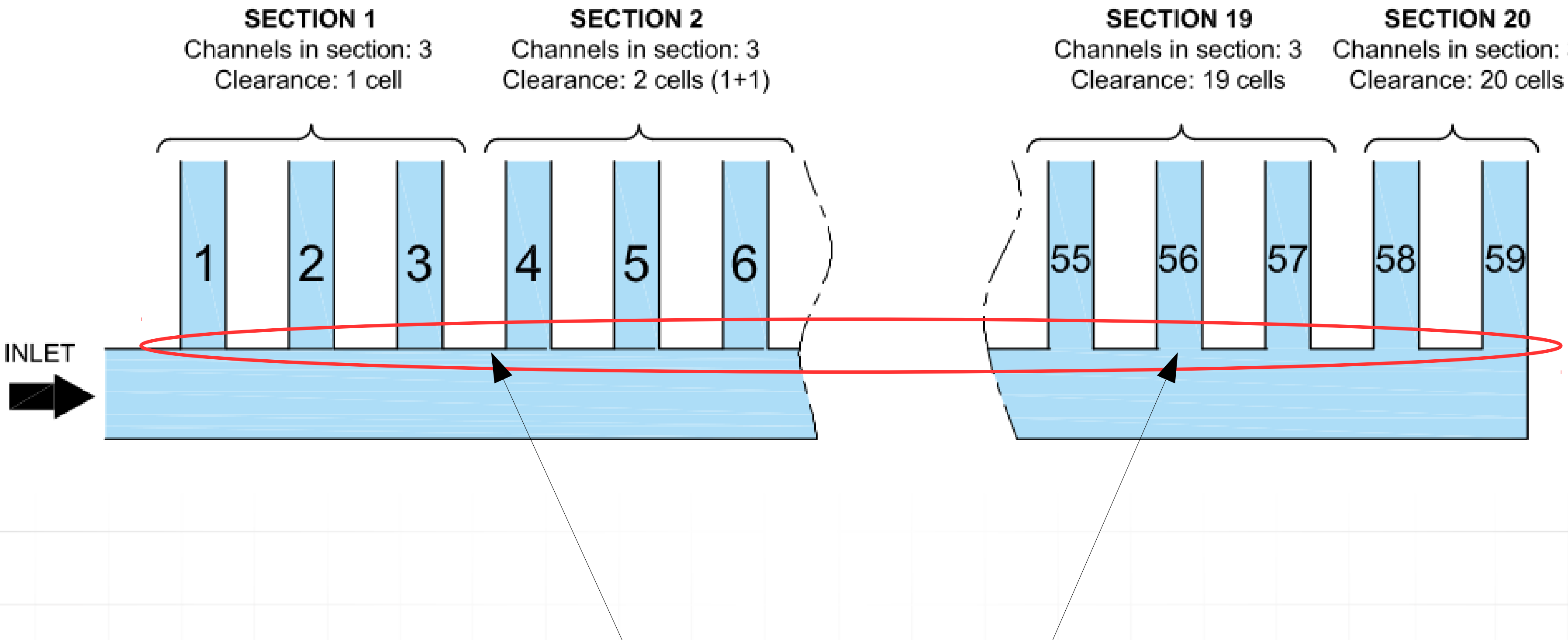
60 plates PHE

Re = 100

Re = 1000

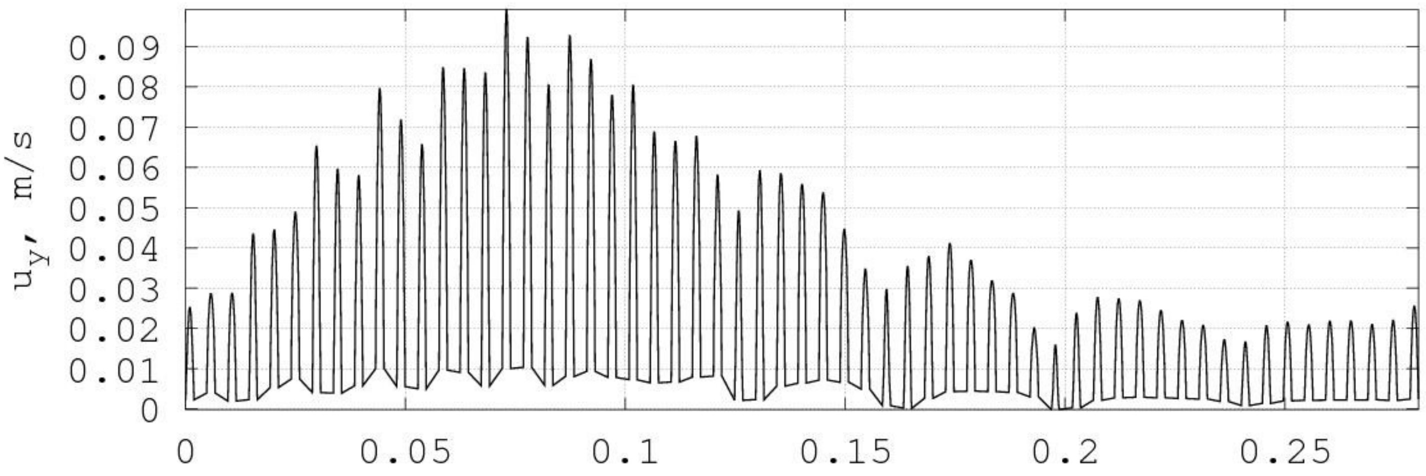
Single phase incompressible solution

GEOMETRICAL MODIFICATIONS OF PHE

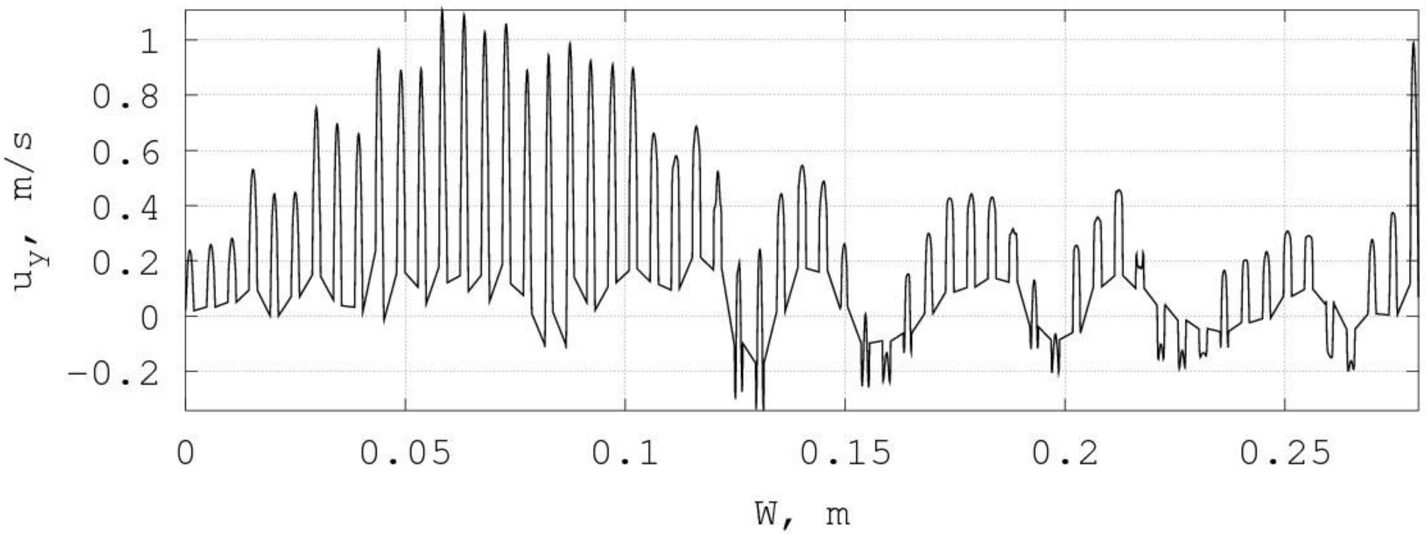


Consecutive channels inlet area gets larger with distance from the PHE INLET

FLOW MALDISTRIBUTION – IMPROVEMENTS



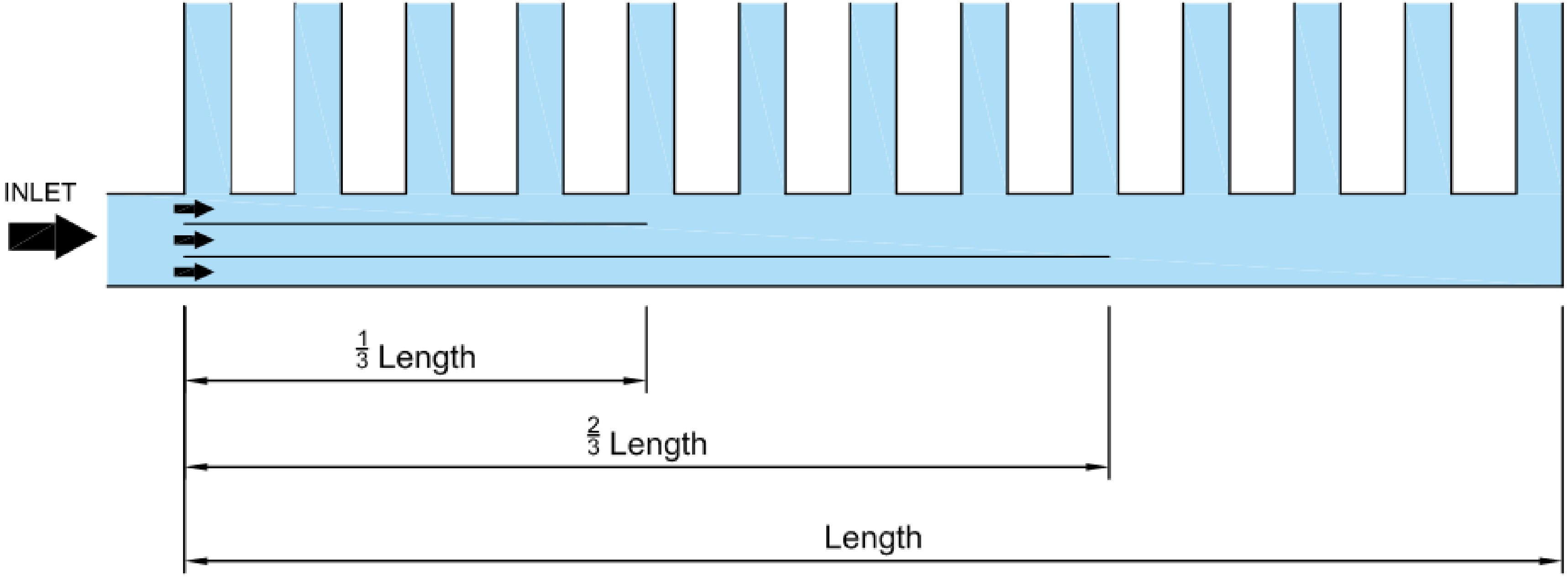
Re = 100



Re = 1000

FUTURE WORK

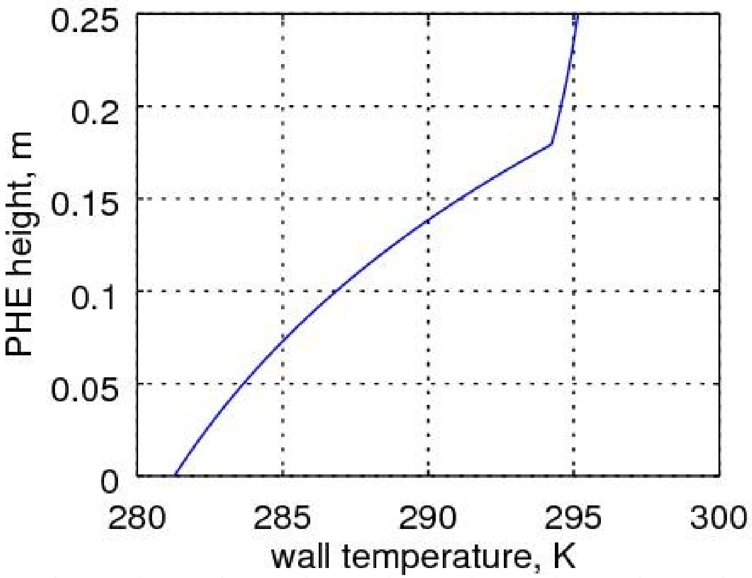
- Study of different geometrical PHE modifications:



- Usage of the proposed mathematical model with phase change to investigate flow maldistribution in PHE

Does the phase change is responsible for flow maldistribution ?

APPENDIX



Re	Mass flow
	kg/s
500	0.004
1000	0.008
1500	0.012